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(NASA-CR-161656) COAL GASIFICATION SYSTEMS
ENGINEERING AND ANALYSIS. APPENDIX B:
MEDIUM BTU GAS DESIGN Final Report (BDM
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COAL GASIFICATION SYSTEMS ENGINEERING AND ANALYSIS FINAL REPORT APPENDIX B: MEDIUM BTU GAS DESIGN

December 31, 1980

BDM/H-80-800-TR

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APPENDIX 8 INTRODUCTION AND OVERVIEW

This appendix is a compilation of reports presenting the technical results of conceptual design work based on five coal gasification technologies and an analysis of schedule requirements for projects based on these technologies. Cost analyses for these designs are contained in Appendix D.

1. Appendix B-1

This is a reference facility design report based on the Koppers-Totzek coal gasification process.

2. Appendix B-2

This is a reference facility design report based on the Texaco coal gasification process.

3. Appendix B-3

This is a reference facility design report based on the Babcock and Wilcox coal gasification process.

4. Appendix B-4

This contains facility definition level design reports based on the Lurgi and BGC/Lurgi processes.

5. Appendix B-5

This presents the results of the schedule analysis.

APPENDIX B-1

KOPPERS-TOTZEK COAL GASIFICATION PLANT REFERENCE FACILITY DESIGN

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1.0 INTRODUCTION

The United States, after a number of years of development based on plentiful and inexpensive oil and natural gas, is entering a period of time when it is essential to supplement these energy sources by the increased use of coal. Coal is the nation's most plentiful fossil fuel. Coal gasification is a means of accomplishing this. While utilization of coal through conversion to gaseous products is not new there is no industry within the US which might serve as a base for establishing cost, operational reliability and requirements, and design data for the large scale environmentally acceptable plants needed.

The Tennessee Valley Authority with systems engineering and analysis support from the George C. Marshall Space Flight Center has initiated a project which would establish the commercial base and demonstrate the requirements for gasifying coal in a large integrated facility. The project consists of gasifying 20,000 tons per day of Eastern coal in a four module plant-the construction of which is staggered to accommodate efficient use of construction manpower and product market development.

The purpose of this study is the systematic engineering and cost analysis of five selected gasifier technologies and the other associated systems required to produce medium BTU gas for delivery into a 600 psig distribution system in Northern Alabama. This fort in a series of design studies provides a reference basis for selection of the final plant configuration and choice of process technologies.

The methodology consists of selection of a systematic breakdown of the facility for purpose of study, identification of available technologies for each system, development of a definition level design and cost analysis for each system, conducting a series of trade studies using the definition design as a basis, and finally development of reference facility designs and cost analyses based on the previous work.

This report covers the work done in establishing a Reference Facility Design based on Koppers-Totzek coal gasification technology. It is one of three such designs, the other two being based on Babcock and Wilcox and Texaco coal gasification processes. The design covers a complete grass roots facility site at Murphy Hill, Alabama. The plant inputs are coal, water and electricity. The plant outputs are medium BTU gas (MBG) and solidified sulfur.

The design approach involved breaking each of four identical modules into process and support systems followed by analysis of each system through identification of available technologies and use of the comparative trade studies to select the preferred processes for each system.

2.0 SUMMARY

A four module, 20,000 TPD, based on KT coal gasification technology has been designed. The plant processes Kentucky No. 9 coal with provisions for up to five percent North Alabama coal. Coal transportation is by river barge except for the Alabama coal which is trucked to the site. Medium BTU gas with heat content of 305 BTU/SCF and not more than 200 ppm sulfur is the primary plant product. Sulfur is recovered for sale as prilled sulfur. Ash disposal is on site. The plant is designed for zero water discharge. Trade studies provided the basis for not using boiler produced steam to drive prime movers. Thus process derived steam in excess of process requirements is superheated for power use in prime movers. Electricity from the TVA grid is used to supply the balance of the plant prime mover power requirements. A study of the effect of minemouth coal cleaning showed that coal cleaning is not an economically preferred route.

The plant design was arrived at by a systematic procedure based on published design work, process trade studies, team engineering experience, a NASA provided module level definition of some twenty systems, and the TVA document, "Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," March, 1980.

The overall plant configuration is shown schematically in Figure 2.1. As shown, General Facilities, Instrumentation and Control, and Coal Handling serve plant wide functions. All other plant components are contained in four identical process modules. The Instrumentation and Control System operates to monitor overall plant performance and to control intermodule relationships. The total list of systems is given in Table 2.1.

The design procedure involved defining available processes to meet the requirements of each system, technical/economic trade studies to select the preferred processes, and engineering design and flow sheet development for each module. Cost studies assumed a staggered construction schedule for the four modules beginning spring 1981 and a 90 percent on stream factor.

The basis and results of the design study are given in Tables 2.2 and 2.3.

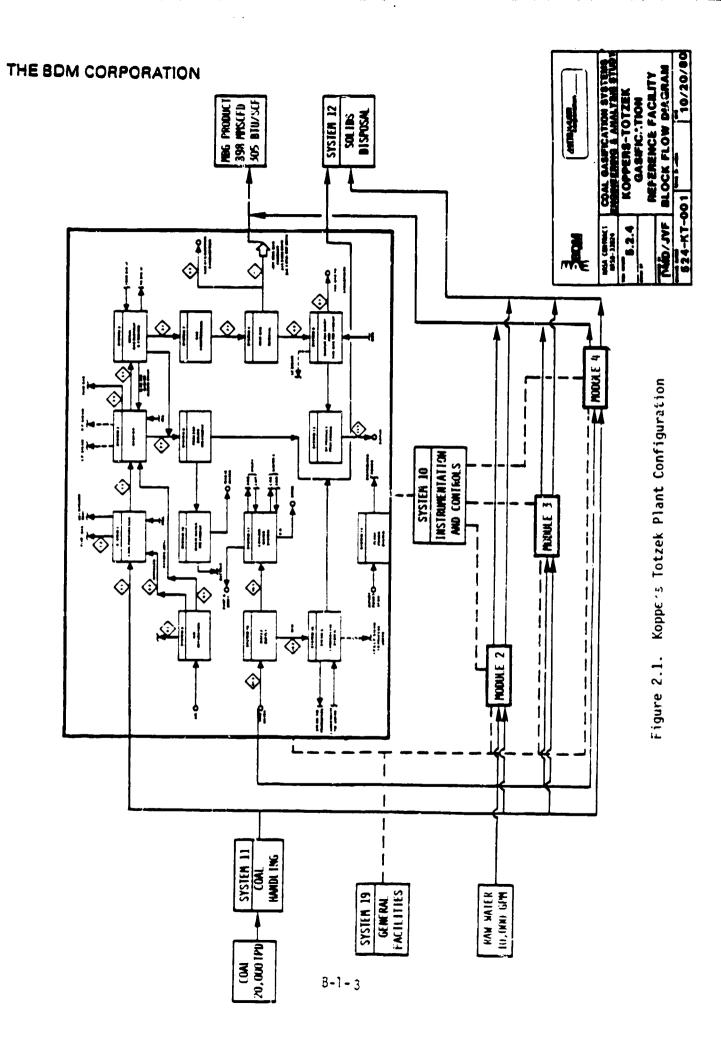


TABLE 2.1. LIST OF SYSTEMS

SYSTEM NO.	NUMBER OF PER MODULE	COST UNITS PER FACILITY	SYSTEM DESCRIPTION
1	1	4	COAL PREPARATION AND FEEDING
2	9	36	GASIFICATION
3	9	36	INITIAL GAS CLEANUP AND COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	2	8	AIR SEPARATION
7	1	4	COMPRESSION
8	1	4	PROCESS SOLIDS TREATMENT
10	0	1	INSTRUMENTATION AND CONT. IL
11	0	1	COAL HANDLING
12	0	1	SOLIDS DISPOSAL
13	1	4	BY-PRODUCT PROCESSING
14	1	4	PLANT POWER SYSTEM
15	1	4	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKE-UP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	0	1	GENERAL FACILITIES
7 8 10 11 12 13 14 15 16 17	1 0 0 0 1 1 1	4 4 1 1 1 4 4 4 4	COMPRESSION PROCESS SOLIDS TREATMENT INSTRUMENTATION AND CONTROL COAL HANDLING SOLIDS DISPOSAL BY-PRODUCT PROCESSING PLANT POWER SYSTEM STEAM GENERATION/DISTRIBUTION RAW WATER MAKE-UP COOLING WATER SYSTEM WASTE WATER TREATMENT

TABLE 2.2. DESIGN BASIS SUMMARY*

PLANT CAPACITY

PLANT TYPE

COAL TYPE

PLANT SERVICE FACTOR

PLANT LIFE

ELECTRICITY SOURCE

WATER

COAL RECEIPT

PRODUCT DELIVERY

PRODUCT QUALITY

SULFUR

COAL COST (1980)

LAND COST (1980)

CLEARING AND GRUBBING (1980)

ELECTRICITY COSTS (1980)

BY-PRODUCT CREDIT

ESCALATION

20,000 TPD

FOUR INDEPENDENT MODULES

KENTUCKY NO. 9

90 PER CENT

20 YEARS EACH MODULE

TVA GRID-4.16KV, 6.9KV, 13.8KV

TENNESSEE RIVER

BARGE +5% BY TRUCK

MBG AT 600 PSIG

MINIMUM 285 Btu/SCF

PRILLED

\$1.25/MM Btu

\$3,000/ACRE

\$2,000/ACRE

NONE

AS PER TVA SPECIFICATION

^{*}Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," Tennessee Valley Authority, March, 1980

TABLE 2.3. DESIGN STUDY RESULTS*

FEED COAL	6,570,000 10,000	
WATER PURCHASED ELECTRICITY	3,347	MM KWHY
MBG PRODUCT		M MSCFD MM MSCFY
MBG PRODUCT MBG QUALITY	305	BTU/SCF
SULFUR PRODUCT (PRILLED)	673 222,000	LTPD
SULFUR PRODUCT TOTAL CAPITAL REQUIREMENTS	\$2,371	MM
OPERATING AND MAINTENANCE COSTS		MM/YR MM/YR
COAL, CATALYST, CHEMICALS	•	·
PLANT OPERATING STAFF	344	PERSONS
MAINTENANCE/YEAR	\$65	MM

^{*} COSTS ARE IN 1980 DOLLARS

Designs in this study were based on published works, engineering experience and judgments, and consultations with appropriate process and equipment vendors followed by design calculations, flow process sheet development and documentation.

Cost estimates are based on normal conceptual level design estimating procedures as described in the "Coal Estimation and Economic Evaluation Methodology," BDM/W-80-258-TR-RV2, submitted as another part of this project work scope and reprinted as Appendix E. Schedule estimates are in accordance with overall guidelines furnished by NASA.

Details of the schedule are considered typical for projects of this type and while they do not represent a detailed analysis specific to this project they represent the complexity and illustrate schedule constraints associated with the subject project. Process flow diagrams are included for the main process systems, air separation system, cooling water system, and steam distribution system as aids in following written system descriptions.

Technical results of the study are reported in this appendix. A cost analysis is included as a part of Appendix D.

3.0 OVERALL FACILITY DESCRIPTION

This section presents and discusses the process design of a plant designed to gasify 20,000 tons per day of coal in four 5000 TPD modules and to produce a medium-Btu product fuel gas (MBG) with a nigher heating value of 305 Btu/SCF. The plant design is based upon using entrained flow of gasifiers as offered by Koppers-Totzek, and upon the design criteria (coal characteristics, product specifications, site data, etc.) set forth by the Tennessee Valley Authority in March, 1980.

3.1 Overall Plant Processing

The overall plant configuration is shown in the block flow diagram, (see Figure 2.1). Kentucky No. 9 coal is received by barge and either placed in a dead storage pile or processed in System 11, Coal Handling. Provisions have also been made to receive up to five per cent of plant capacity by truck. Dead storage provides for 90 days of supply in interruption.

Coal, either directly from barges or from dead storage, is processed in a Bradford breaker and conveyed to four process modules of 5000 TPD capacity each.

In each module, the raw coal fed to the plant from System 11 is crushed and dried in System 1 and fed to System 2, the Koppers-Totzek gasifiers. The feed coal is partially combusted (within the gasifier), with oxygen supplied from the air separation plant, to form a raw product gas which is rich in carbon monoxide and hydrogen. Roughly half of the coal ash is entrained with the gas. The remainder drains by gravity from the gasifier. The raw gas leaves the gasifiers passes through ash quenching and slag recovery and is then cooled to 350°F in waste heat recovery boilers. Heat recovered in the waste heat boilers and from the gasifiers is used to generate steam for the plant. The gas is then further cooled to 100°F in Venturi scrubbers and water wash towers to remove residual char and ash particulates, in System 3, Initial Gas Cleanup and Cooling System.

The raw gas leaves the Venturi scrubbers at 100°F at slightly above atmospheric pressure. It is then compressed in System 7 to about 640-650 psia and sent to System 4, the Acid Gas Removal, where essentially all of the hydrogen sulfide and carbonyl sulfide, and about 45 per cent of the carbon dioxide contained in the raw gas is removed. The acid gas removal system uses the Selexol process (licensed by the Allied Chemical Corporation) in which a physical solvent (the dimethyl ether of polyethylene glycol) absorbs and removes hydrogen sulfide, carbonyl sulfide and the concomitant carbon dioxide from the raw gas.

The acid gas (hydrogen sulfide, carbonyl sulfide and carbon dioxide) is recovered from the Selexol solvent and processed in System 5, Sulfur Recovery and Tail Gas Treatment, wherein over 99 per cent of the hydrogen sulfide and carbonyl sulfide is converted into 668 long tons/day of by-product elemental sulfur. The sulfur recovery unit utilizes the conventional Claus process to convert about 90 to 95 per cent of the hydrogen sulfide (and some of the carbonyl sulfide) in the acid gas into by-product sulfur. The residual tail gas from the Claus unit is then processed in a Beavon-Stretford unit to recover additional by-product sulfur.

Oxygen utilized for the partial combustion of coal and recycle char in the gasifiers is produced in System 6, a conventional air separation plant. Atmospheric air is compressed to about 100 psia, cooled to 100°F and then separated cryogenically into oxygen and nitrogen at about 2 psig. The oxygen is compressed to about 22 psia for use in the gasifiers. A small portion of the nitrogen is also compressed for use in transporting pulverized coal.

In addition to the systems contained in each module and described in Section 4, the plant consists of System 10, the overall plant monitoring and control system; System 11, the coal handling and storage system serving all four modules; and System 12, the solid waste disposal system serving the total facility.

The purpose of System 10 is to provide operational monitoring and supervisory master control of module operations. The system includes instrumentation to measure key characteristics of all major streams (mass, temperature, pressure, composition, density, and/or others as appropriate) and to display these characteristics in a central location. A dedicated computer is provided to record data of interest at the facility management level. Telecommunications capabilities are provided to communicate with module and systems operations as required to control total plant operation to meet delivery requirements.

System 11, the Coal Handling System, provides for the unloading of coal delivered to the plant either by barge or truck, reclaiming the coal from storage, reducing the size of coal, and transporting the coal to Coal Preparation and Feeding, System 1.

Raw coal is unloaded from barges by the barge unloading subsystem which is designed to unload up to ten 1500-ton capacity barges per shift. The coal is unloaded at an average rate of 1200 tons per hour on a five-day week basis. Coal is transferred by conveyor to a radial stacker which then produces a kidney-shaped coal pile containing live and dead storage. Coal is reclaimed from live storage and conveyed to a Bradford breaker where it is reduced in size to 2" x 0. Coal from trucks is unloaded into a chute from which it is conveyed to the Bradford breaker. Crushed coal from the Bradford breaker is transported to day storage silos from which it is transferred via vibrating

belt conveyors to coal preparation and feeding, System 1, for further processing. Coal fines from the Bradford breaker are collected and sent to the gasification unit, System 2.

The purpose of System 12 is to store solid waste generated by facility operation during the facility life.

This system consists of a lined impounding pit sized to contain 20 years of solid waste. It includes the conveyor system to move the solids to the pit and a leachate recovery area to recover and pump leachate to the process condensate system.

Module descriptions are the subject of Section 4.

3.2 Facility Balances

The key material inputs and product outputs for the overall plant may be summarized as:

INPUTS:	Raw Coal 98% Oxygen Raw Water Intake Imported Electric Power	20,000 short tons/day 16,400 short tons/day 10,000 gpm 425 MW
	Imported Lieutric rower	TES MA

OUTPUTS:	Product Gas	898 MM SCFD
	By-product Sulfur	668 long tons/day
	Solids to Disposal (Wet)	4,335 tons/day
	Solids to Disposal (Dry)	3,252 tons/day

Table 3.1 gives the thermal efficiency for the facility. Table 3.2 gives the final product characteristics.

TABLE 3.1. CONVERSION EFFICIENCY KOPPERS-TOTZEK PROCESS 20,000 TPD FACILITY

<u>INPUTS</u>	10 ⁶ BTU/HR	PER CENT
① COAL TO FACILITY	18,300	
② ELECTRIC POWER TO FACILITY	1,448	
OUTPUTS		
3 MBG FROM FACILITY	11,426	
4 COAL FINES FROM FACILITY	-0-	
EFFICIENCY		
COAL-TO MBG (3 + 1) x 100%		62.4
OVERALL PRODUCT EFFICIENCY 3 ÷ (1 + 2) x 100%		57.8
OVERALL FACILITY EFFICIENCY ($3 + 4$) + ($1 + 2$) x 100%		57.8

TABLE 3.2. PRODUCT CHARACTERISTICS

	VOLUME PER CENT
CARBON MONOXIDE	63.4
HYDROGEN	29.6
NITROGEN AND ARGON	1.5
CARBON DIOXIDE	4.9
METHANE	<u>0.5</u>
	100.0
HYDROGEN SULFIDE	4.0 PPM VOLUME
CARBONYL SULFIDE	16.0 PPM VOLUME
TOTAL SULFUR	20.0 PPM VOLUME
WATER VAPOR	7.0 LBS/MM SCF
HIGHER HEATING VALUE	305 BTU/SCF
TEMPERATURE	75 ^O F
PRESSURE	615 PSIA

4.0 MODULE DESCRIPTIONS

Each of the four modules in the plant consist of identical sequences of process systems as given in Table 2.1. Process flow diagrams and an equipment list are included in Section 5 for the principal process systems and such auxiliary systems, as cooling water and steam systems.

Each module receives 5000 TPD of coal from System 11, Coal Handling. The sequence of processing shown in Figure 4.1 is repeated in each of the four modules. Descriptions of each modular system are as follows.

4.1 System Descriptions

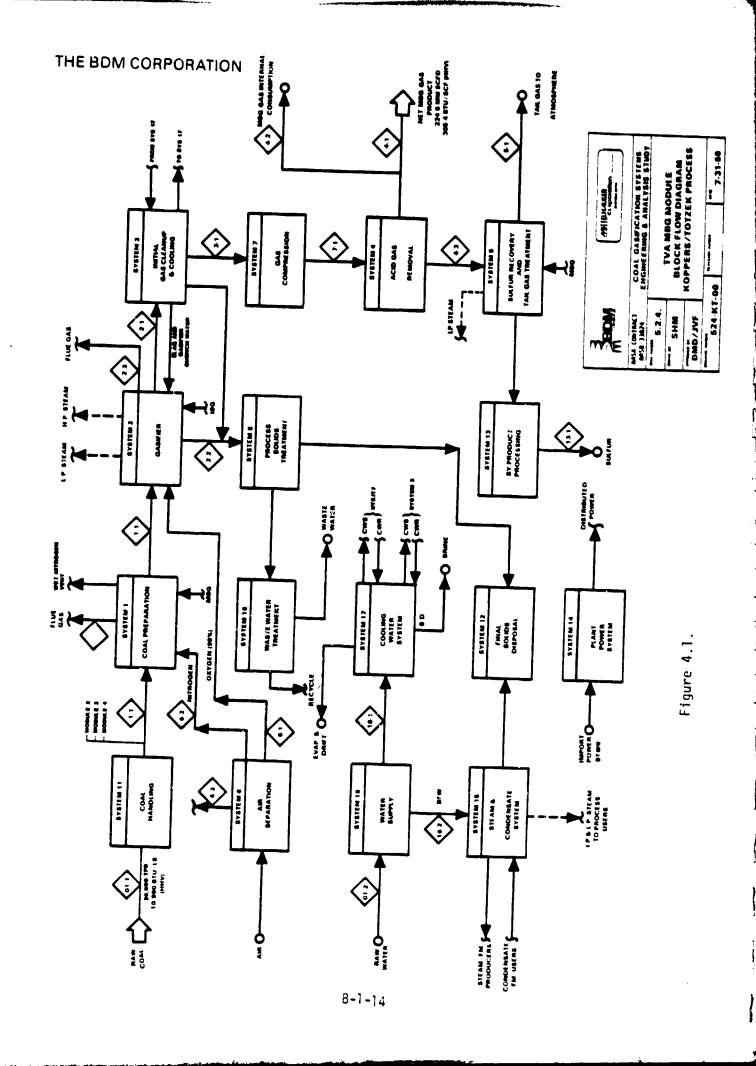
4.1.1 System 1 - Coal Preparation and Feeding

This system receives the rough processed coal from coal handling, System 11. The operations carried out in System 1 include: pulverizing coal to the required size, screening and recycle or oversized fractions, drying of processed coal to the required final moisture content, and transporting of the sized, dried coal to Gasification, System 2, for direct gasification. System 1 is shown schematically in Drawing Number 524-KT-01.

Crushed 2" x 0 coal is received from System 11 via a vibrating belt feeder. Tramp iron separators then remove ferrous metal items from the coal stream and discharge them through the tramp iron chute. The coal is then weighed on a belt scale after which it goes to a pulverizer-dryer. The pulverizers crush the coal to the required 70% through 200 mesh size, while simultaneously drying the coal to around 2% moisture. Drying is accomplished by a hot nitrogeous gas stream. The nitrogen gas stream is heated in a direct fired heater utilizing MBG product gas. Dried, pulverized coal is conveyed by the hot nitrogen gas stream to gasifier service bins. The pulverizer-dryers are controlled so as to maintain coal particle temperatures between 160-180°F, in order to avoid coking or devolatilization. Level controls on the gasifier service bins regulate the intermittent feeding of gasifier feed bins. Transport nitrogen gas is separated from the coal in the service bins prior to dust removal and subsequent venting. Gasifier feed bins are connected to the gasifier via variable speed screw feeders. These screw feeders discharge the coal to mixing heads where it is entrained in oxygen and low pressure steam, and transported via pipes to the gasifier burners. All fines are collected and returned to the service bins.

4.1.2 System 2 - Gasification

The Koppers-Totzek gasifier is of the high temperature, cocurrent entrained flow type. It is a proprietary unit licensed by Koppers Company. There are four feed points on the four-headed gasifiers, and two burner heads on each point. Each burner is arranged to produce a flame pattern which intersects the pattern of the adjacent burner on a head. Oxygen from the air



separation unit, System 6, and low pressure steam generated in the gasifier jacket, transport the pulverized coal to the gasifier. System 2 is shown schematically in Drawing Number 524-KT-02.

The coal, oxygen, and low pressure steam react to gasify the carbon and volatile matter in the coal. Due to the high temperature operation, as high as 3300°F, no tars, oils, phenols, etc., are produced. Each module contains eight operating and one spare gasifier train.

Approximately 50% of the ash in the coal leaves the gasifier in the form of liquid slag, out the bottom of the gasifier. Entrained slag droplets are quenched with water at the gasifier outlet, in order to drop the temperature below the ash fusion temperature. This prevents adherence of liquid ash to the waste heat boiler tubes. The gas, with entrained particulate ash, passes to the waste heat boiler located directly over the gasifier. Due to a reduction in velocity of the gas in the waste heat boiler, the heavier particles of ash are returned via a chute, to the slag quench tank below the gasifier. The waste heat boiler cools the gas to about 350°F, prior to cleanup. The waste heat boiler generates 49,500 pounds per hour of 1450 psig steam per gasifier.

The jacket of each gasifier produces 49,600 pounds per hour of low pressure steam from waste heat escaping the yasifier refractory lining. Of this, 7,065 pounds are used as feed to the reactor.

The oxygen feed rate is 0.90 0₂/1b dry coal. The higher heating value of the product gas produced is 305 Btu/SCF.

The Koppers-Totzek gasifiers operate at approximately 7 psig. Each module contains eight operating and one spare reactor.

4.1.3 System 3 - Initial Gas Cleanup and Cooling

The Koppers-Totzek gasification system includes a proprietary means for removal of particulate matter, as well as cooling the gas the final amount necessary to achieve temperatures compatible with sulfur removal systems. System 3 is included in Drawing Number 524-KT-02.

The raw gas exiting the waste heat boiler at around 350° F is washed in a pair of Venturi scrubbers arranged in series. The primary scrubber removes the bulk of the particles (around 90%). The gas from the secondary scrubber passes to the final gas cooler. The gas is cooled with water to around 95° F in this cooler; the total particulate removal efficiency of this system is around 99.9%. The cooled gas passes to System 7, compression.

The scrubbers and cooler produce a considerable quantity of particulate laden waste water. This water is constantly recycled after solids removal in a clarifier and cooling in a cooling tower. Due to losses in the cooling

loop and water entrained in the slag stream from the quench tank, makeup water must constantly be provided. Gas cleanup and cooling are integral parts of the K-T system. Thus, there are eight operating and one spare system in each module. Figure 4.2 presents a balance of the water system around the gasifier and cleanup system.

4.1.4 System 7 - Compression

The purpose of this unit is to compress the MBG from initial gas cleanup and cooling, System 3, to a pressure sufficient to provide a plant boundary pressure of 600 psig. Water produced in the multi-stage compression process is collected and recycled to System 3 to provide a portion of the make-up required. The product gas compressors for the Koppers-Totzek module are motor driven machines. Outlet pressure from the compression unit is around 650 psig in order to provide for pressure losses in acid gas removal, dehydration, and gas metering.

Each module contains:

(1)	Axial Compressor		34,500 HP
(2)	2-Stage Centrifugal	Compressor	34,600 HP
	1-Stage Centrifugal		21,300 HP

System 7 is shown schematically in Drawing Number 524-KT-07,

4.1.5 System 4 - Acid Gas Removal

The acid gas removal system utilizes the Selexol solvent process and received 645,000 lbs/hr of sour gas at 100°F and 665 psia. The sweet (desulfurized) gas leaves the Selexol absorber at about 615 psia and a temperature of 60°F. The component removals across the absorber are:

	Inlet Gas (mols/hr)	Outlet Gas (mols/hr)	% Removal
Methane	125	125	0.00
Hydrogen	8,125	8,123	0.02
Carbon monoxide	17,406	17,374	0.18
Carbon dioxide	2,483	1,335	46.23
Hydrogen sulfide	433	1	99.77
Carbonyl sulfide	49	4	91.84
Nitrogen	420	420	4.00
Water	42	5	88.10
	28,958	27,387	

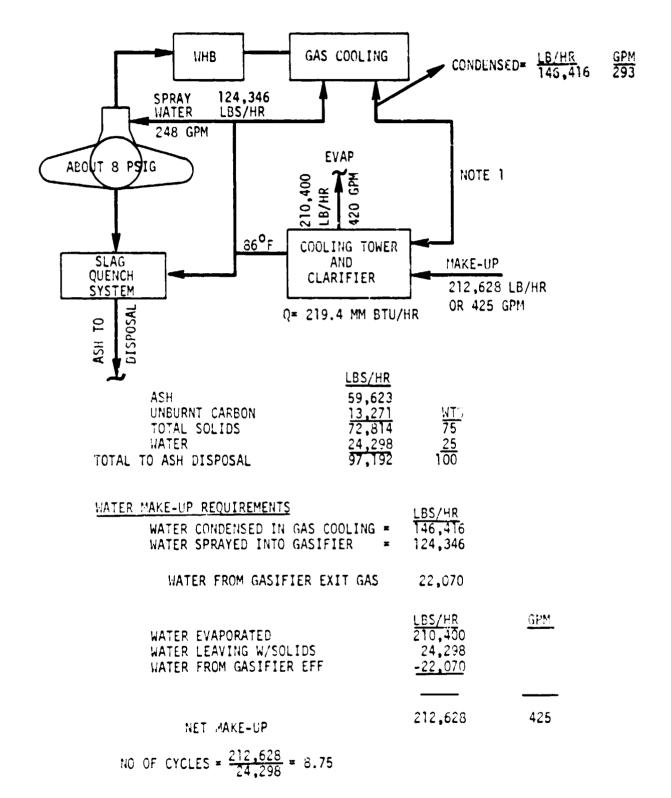


Figure 4.2. Process Cooling Tower Nater Dalance

The Selexol sulfur removal unit is a proprietary process licensed by Allied Chemical Company. The primary purpose of this system is to remove H₂S and COS from the product gas stream in order to obtain a sulfur concentration of not more than 200 ppmv. Secondarily, the process removes CO₂, which increases the gas heating value. System 4 is shown schematically in Drawing Number 524-KT-O4.

The Selexol process removes sulfur compounds and carbon dioxide by means of countercurrent contact in an absorber. The physical absorption of the compounds to be removed is done using the dimethyl ether of polyethylene glycol as a solvent. Chilled solvent is used. Solvent chilling is accomplished by means of an absorption refrigeration system utilizing low pressure gasifier jacket steam.

CO₂ removal from the rich solvent leaving the absorber is accomplished by a series of successively lower pressure flashes. Flashed gas is recycled to the absorber inlet. Sulfur compounds are removed in the stripping column. The stripping column utilizes a steam heated reboiler. The stripper overhead gas, after cooling to about 120°F, flows to the sulfur recovery and tail gas treatment unit, System 5.

The sour gas from the Selexol unit is rich in ${\rm CO_2}$, and at slightly above atmospheric pressure.

4.1.6 System 5 - Sulfur Recovery and Tail Gas Treatment

Acid gas from Acid Gas Removal, System 4, is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean H_2S acid gases. Typically, in a Claus-type sulfur plant, the acid gas is first passed through a knockout drum before entering the reaction furnace. The chemistry of the process involves converting H_2S to elemental sulfur according to the following equation:

$$2H_2S + SO_2 \rightarrow 3S + 2H_2O$$

The reactions are exothermic, and the heat liberated generates steam in the reaction furnace boiler and in the sulfur condenser. The sulfur from each condenser is drained to a recovery pit in By-Product Processing, System 13, and the tail gas from the final condenser is fed to a Beavon tail gas treating unit where essentially complete removal of the remaining sulfur compounds is achieved before discharge to the atmosphere. The Beavon sulfur removal process reduces the sulfur content in the tail gas to less than 100 ppm. In this system, hydrogenation and hydrolysis are used to convert essentially all sulfur compounds to hydrogen sulfide. This gas is then cooled and passed into a contactor where the hydrogen sulfide is absorbed by the redox solution and oxidized to elemental sulfur. The reduced redox solution is reoxidized by contact with air and subsequently recirculated to

the contactor. Elemental sulfur is removed in the air-blowing step as a froth which is pumped to a sulfur melter to be melted under pressure, separated from the redox solution and transferred to By-Product Processing, System 13. The decanted redox solution is returned to the system. System 5 is shown in Drawing Number 524-KT-Q5.

The system receives 48,370 lbs/hr of acid gas from the Selexol solvent regenerator at about 7 psig and 120° F.

The inlet gas composition is:

	<pre>Inlet Gas (mols/hr)</pre>
Hydrogen Carbon monoxide Carbon dioxide Hydrogen sulfide Carbonyl sulfide Nitrogen Water	3.3 31.3 1,148.9 432.0 44.0 0.0 153.0
	1.812.5

The recovery of by-product elemental sulfur from the sulfur recovery system is 167 long tons/day, which amounts to an overall sulfur recovery of about 99.9 percent.

4.1.7 System 6 - Air Separation

The purpose of this unit is to supply oxygen to the gasifiers. The process is non-proprietary and is offered in package form by several designer/manufacturers in the U.S. The gaseous oxygen stream of 98% purity is produced by distillation of liquified air. System 6 is shown schematically in Drawing Number 524-KT-06.

In the unit, which consists of multiple trains, 1,484,000 lbs/hr of atmospheric air is filtered and compressed to 110 psia by multi-stage compressors. Drivers are steam turbines with motors installed for start-up. The filtered-compressed air is cooled via cooled heat exchangers, then passes through reversing heat exchangers where it is cooled to slightly above its liquefaction temperature by heat exchange with the oxidant and waste gas streams, and by subsequent expansion through a turbine.

Water and carbon dioxide freeze out in the reversing exchangers, and are therefore removed from the inlet air stream. Subsequent switching of the exchanger sublimes the water and carbon dioxide into vapor which is vented as waste.

The cold air then enters the high pressure column for initial purification. The vapor at the top of the column is nitrogen containing 10 ppm of oxygen. Most of this is condensed in the high pressure condenser/low pressure reboiler and returned as reflux to the high pressure column. Some of this nitrogen vapor becomes the feed stream to the expansion turbine and some becomes low pressure product nitrogen gas after being superheated in the superheater and being warmed to about -150°F in a reversing heat exchanger.

Some of the nitrogen condensed in the condenser/reboiler becomes the liquid nitrogen product. It is first subcooled in the waste nitrogen core subcooler and then flashed into the liquid nitrogen separator. Liquid nitrogen product from the separator is transferred to the liquid nitrogen storage.

An impure reflux stream containing a small percentage of oxygen is withdrawn at an intermediate point in the column. It is subcooled and flashed into the top of the low pressure column.

The final purification of oxygen takes place in the low pressure column. The feeds to this column are impure reflux, and crude liquid oxygen from the high pressure column is subcooled and passed through the hydrocarbon absorbers before being flashed into the low pressure column. This procedures assures that all hydrocarbons are removed from the oxygen before purification. Processing the crude oxygen after it is liquified assures that the absorption is done at a relatively constant pressure and temperature and thus limits the possibility of accidental desorption.

As a further safeguard, a portion of the liquid oxygen in the low pressure sump is constantly withdrawn, passed through a guard hydrocarbon absorber and returned to the sump. This prevents trace quantities of hydrocarbons from accumulating in the low pressure sump.

The separated oxygen from the cold box (at 2 psig and 70°F) is compressed, in four stages of compression, to 40 psia for use in the gasifiers.

Each module contains a 23,000 HP axial air compressor powered by a 1450 psig superheated steam turbine, a 7500 HP centrifugal air compressor driven by a 25 psig steam turbine and a 1750 HP centrifugal oxygen compressor powered by an electrical motor. All turbine drives have electrical start-up motors.

The air separation plant was designed as two trains, each in operating service and each providing one-half of the oxygen requirement.

4.1.8 System 8 - Process Solids Treatment

Approximately 74,000 pounds per hour per module of ash and slag from the gasifier and gas cooling system along with biological sludges and solid wastes from process condensate treatment are treated in this unit. Gravity settlers separate dense solids from the waste streams.

Float thickeners-clarifiers treat slurry from the gravity settlers. Thickener and coagulant aids are added to facilitate solid-liquid separation.

Rotary drum filters filter sludges from the process condensate treating systems and the float thickeners.

Recovered water is sent to the process condensate treating system, and solids are conveyed to final solids disposal.

4.1.9 Sysiem 13 - By-Product Processing

Sulfur is the only plant by-product other than ash/char solids which are disposed of on-site. Molten sulfur is pumped to a prilling tower in a continuously flowing circulation system. In the tower, sulfur is dispensed in droplets through nozzles. Droplets fall counter current to a stream of cooling air and solidify prior to landing in the bottom prill collection section. From the prill tower, sulfur is conveyed to a storage building from which it is transferred by truck or barge for sale. This system is illustrated in Drawing Number 524-KT-13.

4.1.10 System 14 - Plant Power System

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

- (1) Develop the necessary voltage stepdown arrangement for plant requirements
- (2) Distribute the necessary power to the plant equipment.

TVA's incoming substation transformers receive power from its primary distributed voltage switching station and step down this voltage to a medium voltage to supply the plant electrical power requirement for motors, heaters, lighting, and other miscellaneous loads.

The Medium Voltage Electrical Distribution Systems is a secondary selection system (double ended supply) with several medium voltage buses. Each medium voltage bus receives power from its respective incoming substation transformer through an incoming breaker and supplies power to the medium voltage distribution system through the feeder breakers.

The Low Voltage Electrical Distribution System typically consists of multiple 480 V double-ended load centers and 480 V motor control centers (MCC's) supplying the power to 480 V loads throughout the plant. Two load centers are interconnected through a normally open tie breaker. In the event of loss of one load center transformer or its feeder, the 480 V loads of the affected load center are fed by the second load center through the tie breaker.

Each load center consists of an incoming line section, load center transformer, and low voltage section with metal enclosed draw out power circuit breakers.

Load center transformers are air cooled, dry type, 150°F temperature rise, with delta connected primaries and wye connected secondaries. All load center feeder circuit breakers are 1600A frame and 50,000A RMS symmetrical interrupting capacity. The 480 V motor feeder breakers are electrically operated with instantaneous and long time trip units.

480 V MCC's consist of starters, feeder circuit breakers and control devices, assembled in a common structure with horizontal and vertical buses.

A 125 volt DC system supplies control power for medium voltage and 480 volt plant switchgear control, protective relaying and annunciation. The system also supplies power for emergency lighting.

4.1.1: System 15 - Steam Generation/Distribution

All of the plant steam is process generated.

Twenty-five psig steam (396,000 pounds per hour per module) is generated in the gasifier jackets. The gasifier steam requirement is 50,000 pounds per hour per module. Steam in excess of the gasifier requirements is used along with steam from System 5, Sulfur Recovery, to drive gas compressor in System 7, the refrigerant cooling water in System 4 and other plant uses. Waste heat boiler steam, generated at 1450 psig, is used to drive gas compressors and a small amount is let down to 550 psig for use in the sulfur recovery system. One hundred psig steam is generated in the sulfur recovery system and from gas compression turbine exhaust. It is used for reboiler heating in System 4 and to deaerate the steam condensate system. The steam system is given in Drawing Number 524-KT-10. The balance is shown in Figure 4.3.

4.1.12 System 16 - Water Supply

The raw water treatment unit is designed to provide treated and untreated water for the following facility water systems:

- (1) Fire water
- (2) Service water
- (3) Potable water
- (4) Cooling water
- (5) Boiler feed water

Raw water is pumped from the river to a fire water-raw water storage tank.

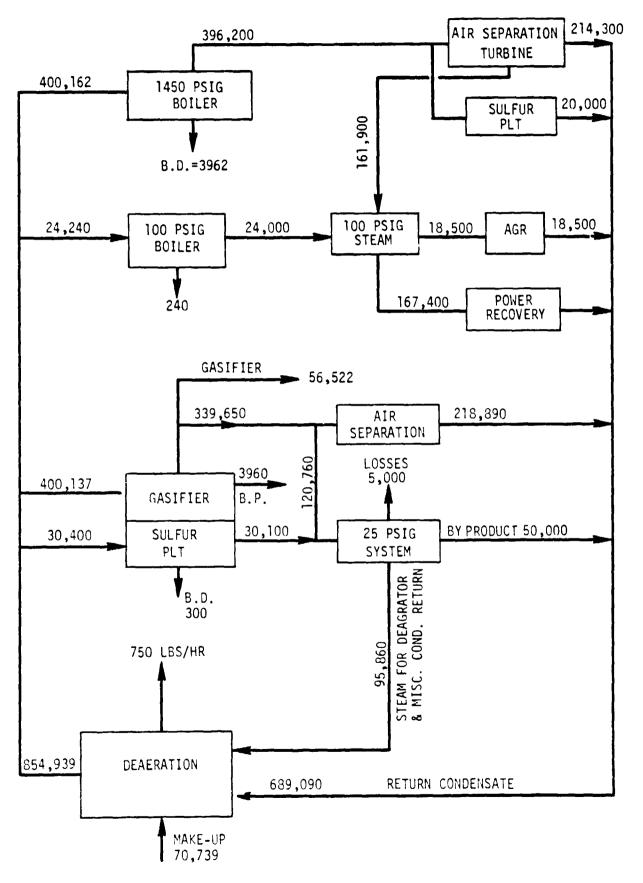


Figure 4.3. Steam System Balance-1bs/hr per Module B-1-23

The raw water-fire water storage tank provides surge capacity for water treatment as well as storage capacity for fire water. During an emergency, fire water is pumped from the tank to the fire water heater system. The fire water pumps are motor driven and have a diesel engine driven spare. The spare pump is equipped with automatic start-up capability in case of power failure.

The raw water is pumped from the raw water-fire water storage tank to the softener-clarifier. Lime, alum, and polyelectrolyte from the clarifier bulk chemical storage and feed system are added to the softener-clarifier, which is equipped with an internal flocculation mechanism. The alum and polyelectrolyte aid in the removal of suspended solids from the raw water. Lime is added during the clarification step to "cold soften" the raw water. Chlorine is added to the raw water to inhibit algae growth in the clarifier and sand filters and reduce organic contamination.

The underflow from the clarifier is a one wt percent sludge and is pumped to solids treatment for further processing.

The clarified and softened raw water from the softener-clarifier flows to the self-backwashing sand filters where additional suspended solids are removed. A pressure differential across the filter bed initiates the backwash cycle. The backwash flows by gravity to the sand filter backwash sump and is recycled to the softener-clarifier. The filtered water flows to the filtered water storage tank and is utilized as cooling tower make-up for the process cooling tower as service water for general plant use, as feed to the demineralizer package, and as feed to the potable water system.

Water intended for potable services is chlorinated and again filtered to meet American Water Works Association (AWWA) standards and stored in a tank sized to hold a day's potable water requirements. The chlorine residual is maintained at 0.5-1.0 ppm free chlorine in the tank.

Filtered water intended as feed to the demineralizer package is injected with sodium sulfide to remove trace amounts of chlorine which adversely affect the demineralizer resins and after filtered through activated carbon to remove any remaining organic contaminants and dissolve iron.

In the demineralizer, the mineral salts present in the water are removed by ion exchange. A two-step demineralization system, utilizing strong cation and strong anion exchangers in series, is provided. A degasifier following the strong cation and magnesium, which the anion exchangers remove anions such as chloride and sulfate. The strong anion exchanger also removes silica. The degasifier is provided to remove carbon dioxide and other dissolved gases.

The mixed bed polisher is provided to remove silica to 0.02 ppm and to polish returned turbine condensate for reuse.

The boiler feed water deaerating heaters operate at 30 psig and 250°F. The deaerators reduce the oxygen content of BFW to 0.005 cc/liter.

Hydrazine or sodium sulfite is injected into the storage compartment of the deaerators for chemical scavenging of any residual oxygen. Morpholine is injected into the suction of the boiler feed water pumps to protect the condensate systems.

The total water requirement is illustrated in Figure 4.4. Allowances for water treatment blowdown and contingency add to that shown to result in a total requirement of 2500 GPM per module.

Drawing Number 524-KT-16 shows System 16.

4.1.13 System 17 - Water Cooling

The purpose of this unit is to provide cooling water to the various process users in the facility.

The cooling tower system includes the tower and fans, side stream filters, circulating water pumps, cold water basin, blowdown system, chemical addition equipment, and distribution system.

Cooling water is pumped from the cold water basin, through the distribution system to the process heat exchangers where low-level, sensible heat is picked up, and back to the cooling tower. The cooling tower rejects low-level heat by evaporative cooling to air drawn through the cooling tower by the cooling tower fans.

A portion of the circulating water is passed through side stream filters to reduce loading to suspended solids, dirt and scale.

The dissolved solids level of the cooling water is maintained by a continuous blowdown stream to the process condensate system. Water level in the cooling tower basis is maintained by continuous make-up of the clean water from the raw water treatment system.

The blowdown stream is passed through a blowdown treatment system to recover chromate ions via ion exchange or by chemical reduction to chromium hydroxide and is sent to waste treatment for disposal.

Chlorine is added to the cooling water on a routine periodic basis to prevent algae growth. Chemical algicides are added periodically to further eliminate algae growth. Sulfuric acid is added to control pH, and zinc and chromate inhibitors are added to the cooling water for corrosion control. Occasionally, a polyphosphate dispersant is added to enhance the action of the inhibitors. This system is illustrated in Drawing Number 524-KT-17.

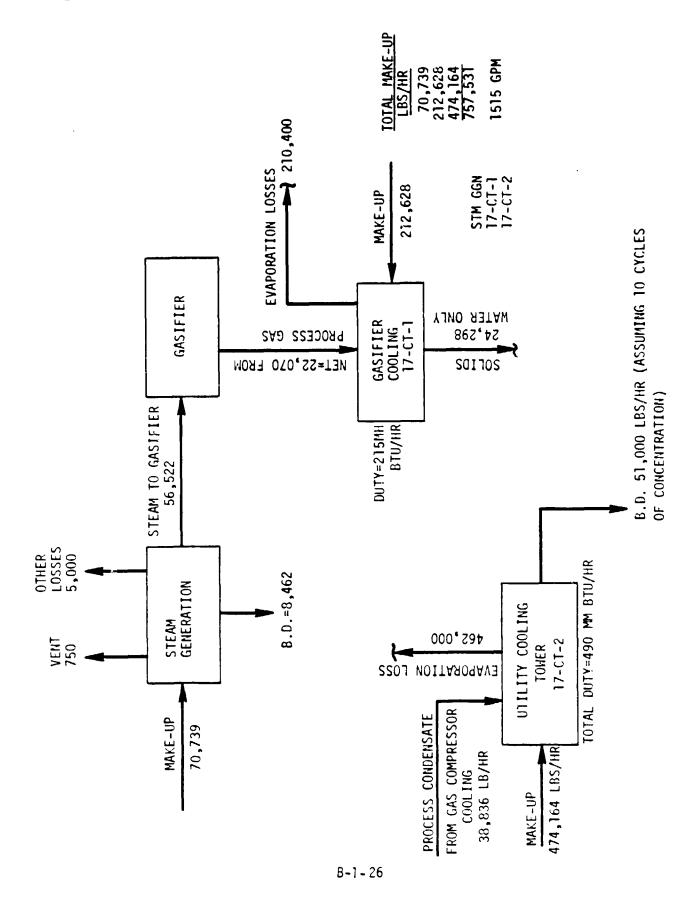


Figure 4.4. Overall Water Balance-1bs/hr per Module

4.1.14 System 18 - Waste Water Treatment

The purpose of this unit is to collect and treat all plant liquid effluent streams. The plant design is predicated on "zero discharge" and permits recycle and reuse of treated water. Streams treated include the following: .

- (1) Oily water sewers
- (2) Coal pile run-off
- (3) Storm water run-off
- (4) Demineralizer regenerant wastes and rinse water
- (5) Cooling tower blowdown

- (6) Sanitary waste water
 (7) Gasifier slag quench drains
 (8) Separated water from solids treatment
- (9) Filtrate from biological treatment.

Process operations include:

- (1) Oil Separator streams containing free and dissolved oil and treated in a gravity separator utilizing an emulsion breaking chemical and heat to separate the oil-water mixture
- (2) Sour Water Stripper water streams with appreciable H₂O and HN₂ residuals are steam-stripped to remove these contaminants
- (3) Equalization Basin liquid streams with extremely high or low pH are mixed in an equalizing basis and treated with sulfuric acid or caustic to change the mixed pH to a value of 6.0 - 8.0
- (4) Gravity Settling-Thickener liquid streams with high suspended or dissolved solids are treated in a gravity settler-thickener and mixed with lime, alum, coagulant aids, and polymers to facilitate separation and thickening
- (5) Multiple Effect Evaporation neutralized wastes and brines are evaporated to recover water and concentrate the solids.

The recovered, treated water is used as make-up to cooling towers or raw water supply. The resultant solids are conveyed to the solids disposal system. System 18 is illustrated in Drawing Number 524-KT-18.

4.1.15 System 19 - General Facilities

The purpose of this unit is to provide equipment or services to support the gasification facility at the facility level.

This unit is a general facility category and provides the following equipment and services:

- (1) Administration building
- (2) Laboratories
- (3) Change rooms
- (4) Warehouses
- (5) Maintenance buildings

- (6) Operation centers (7) Security offices (8) Plant air facility
- (9) Fire house
- (10) Visitor reception
- (11) Plant fencing
- (12) Plant lighting
- Roads, bridges (13)
- (14)Docking facilities
- (15) Interconnection pipe ways
- (16) Fire protection network
- (17) Flare stacks and neaders
- (18) Plant instrument air compressors (19) Environmental monitoring
- (20) Site preparation.

4.2 Module Balances

Material and energy balances have been determined for each module. Table 4.1 gives stream compositions and quantities for each intersystem stream. Table 4.2 gives an energy balance at the modular level.

TABLE 4-1
MATERIAL BALANCE
K-T PROCESS
INTEGRATED FACILITY/NODULE DEFINITION

1-1 PROJECT 1-1 PR		23.901	5.761	767						•		X9.45	376,817	3,806	629 000	140	
RAW WATER TO WATER TREATMENT LB-MOLS LB-S													226 000				2.453
RAW COM, FEED LIB-MOLS LIBS HR	253,637	23.901	15.449				,		4		59,650	376 817	39,853	416,670			
STREAM NUMBER STREAM ND SYMBOL PORMULA MEIGHT	12.01	SULFUR SANCON NO 28.016	13 a				SULPUR DIOXIDE SO2 26.13	N CYANIDE MCH			OTHER SOLIDS		18,016	GAS MOLECULAR WEIGHT	PRESSURE, PSTA		

TABLE 4.1 (continued)
MATERIAL BALANCE
K-T PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER			1-1		(2-1)		\(\sigma_{12}\)
STREAM ID			FLUE GAS	AS	PAN CAS TO COOLING	DOOL THE	ASH TO SOLIDS TREATMENT
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LBS	LB-MOLS HR	LBS	LBS LBS
CARBON	U	12.01	4	ı			G
HYDROGEN	H	2.016		•	8.125.4	16.300	
OXYGEN	0.0	32.0	47.6	1,52".2		•	
NITHOGEN & ARGON	N,	28,016	1,082.7	30,332.9 •	419.9	11,764	
SULFUR	5,	32,06					4
CHIORINE	ប	35.453	-				
CARBON MONOXIDE	8	28.01	1000		17.605.8	487,536	
CARBON DIOXIDE	8	44.01	345.1	15,187.9	2,483.2	109,286	
HETHANE	Ę,	16.042		-	125	2.005	
LTHYLENE	S H	28.052	•	-		1	
ETHANE	2,11,5	30.068		-	•		
PROPYLENE	C.U.	42.078	•				
PROPANE	S F	44.094			•		
HYDROGEN SULFIDE	HSS	34.076	1	-	433.2	14,762.	
CARBONYL SULFIDE	cos	60.075		-	48.7	2,926.	
CARBON DISULFIDE	cs,	76,13		-	_		
SULFUR DIOXIDE	so.	64,06	0.1	6.4		•	
RITROUS OXIDE	NO.	30.008	1	-		1	+
AMMONIA	NH.	17,031	1	ı	•	3	
HYDROGEN CYANIDE	нси	27,026	•	ı	-		
HYDROGEN CHIORIDE	HC1	36.461	_		14.3	521.	
NAPHTHA	ı		1		•		
TAR & OII.	ı		1	_	•	•	
ASH	1		1	-	_		59.650
_					_		14.334
SUBTOTAL, DRY			1,475.5	47,050.4	29,055.5	645,180.	73,84
WATER	н,0	18.016	197.5	3,558.2	10, 326.5	186.042.	24.661
TOTAL, WET	•		1,673.0	122,697	39, 382.0	831,222.	90,645
GAS MOLECULAR MEIGHT	HT		30.6		21.	1	
TEMPERATULE, OF			009		350	0	*
PRUSSURE, PSIA			14.4	7	15.4	9	
			cro	/40	14,952,083	186	
N. O. D. H. G. D. H.							49

IABLE 4.1 (continued)
MATERIAL SELANCES
X-T PROCESS
INTEGRATED FACILITY/MODEL DEFILATION

STREAM NUMBER			1-2		1-	1		\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\
STREAM ID			FI HE GAS		RAM CAS TO COMPRESSION	COMPRESSION	AB 584	MIC CAS PRODUCT
			LB-MOLS		TB-110LS	E ::3	LB-NOLS	LUS
	SYMBOL	FORMULA WEIGHT	H.	RII	1418	Z Z	HE I	
CARBON	U	12.91		ı				
ETHOGEN	H	2.016	1		8,125.4	16, 301.	7,308.6	14,234
OXXGER	70	32.0	125.9	4.029				
PITTHOGEN & ARGON	N,	28.936	2,863.3	90,218	419.9	11,764.	3//.9	10,58/
SULFUR	S	32.06	_				-	
CHIORINE	3	35,453	L		17 404 @	72.5 287	15.676.2	437.917
CAMMON MONOXIDE	00	28.01	913.3	771 07	2.483.2	109.286.	1,200.6	52,030
CAEDON DIOXIDE	203	44.01	7,716	887.28	125.	2,005.	112.5	1,804
TILLIAND	V	750 00			-			ı
POURNE		30.07	1					1
FRODYLENE	C2116	42.0.78	1	•	•	1	-	-
FOPARE	C316	44.094	1	1	•			1, 1,
HYDROGEN SULFIDE	358	34.076	ţ	•	433.2	14,762.	0.9511	14.36
CARRONYL SULFIDE	502	60.075	ı	1	48.7	2,926.	1.977	238.90
CARRON DISULFIDE	CS.	76,13						
SULFUR DIOXIDE	505	64.06	0 2651	16.98	-	•		
HITROUS OXIDE	NO	30,008	3	1				
AMMONIA	MH	17.031						
HYDROGEN CYANIDE	HCF	27.026	1	1				
HYPROGEN CHLORIDE	HC)	36,461		1 1 1				,
THE COLL					-		,	
100	1			,				•
OFHER SOLIDS					•		•	
SUBTOTAL. DRY			3,901.7	124,410	29.041.2	099.449	74,638.8	218.151
WATER	П,О	18.016	86 507	71314	2,244.4	40.435	3.6	64.8
TOTAL. WET	,		4, 107.6	131,724	31,285.6	685,095	24,642.4	\$18,216
GAS MALECULAR WEIGHT	E			عي		21.9	21.0	0.
TTMPERATURE, OF			607			100		93
PERSSURE, PSIA			7.71	7	11	6.1	9	519
11. 13. 13.			1,634	,634,993	11.6	11,878,140.	9,151,269	269
W.	i i			_	ŗ			

TABLE 4.1 (continued)
MATERIAL BALANCE
K-T PROCESS
INTEGRATED FACILITY/MODULE DEFINATION

STREAM NIMBER			1-1		(-)			
STREAM ID		-	MBG GAS INTERNAL CONSUMPTION	CONSUMPTION	ACID CAS TO SUL		SULFUR PLAN	
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LIBS	LB -FOLS	E S	LB-1:OLS HR	LIR
CARRON	U	12.01	1	1		L		
HYDROGEN	Н,	2.016	813.5	1,640	3.3	1	24.6	69
OXXGEN	0,	32.0					2 30	20, 30
NITROGEN & ARGON	NS	28.016	42.0	1,177	2		754.0	777-47
SULFUR	S	32.06		1 1				
CALCURINE MONOXIDE	170	20 01	2 072 1	48.743	31.3	877.	1	1
	000	44.01	133.6	5,880	1148.9	50,563.	1,175.3	51,725
MUTHANE	Ç.	16.0.2	12.5	201	1			1
	, II	28.052	1	-	1			
	Célifo	30,068		1			1	
PROPYLENE	CAHO	42.078	-	1	i .	•		
PROPANE	C.)H	44.094			1 000		16000	0.0784
HYDROGEN SULFIDE	HJS	34.076	0.1059	3.61	41.264	14.726.	O'ME	75.0
CARBONYL SULFIDE	sos	60.075	0.4427	26.59	07·hh	7,000.		
CARBON DISULFIDE	cs,	76.13			L			
SULPUR DIOXIDE	505	64,06		_	*	1		
HITROUS OXIDE	NO	30.008						
AMMONIA	NK	17.031	•		•			
	- (27.026				1		
HYDROGEN CHLORIDE	-	36,461		, ,				•
TAR & OII.			1	1				1
ASH					•	-		_
OTHER SOLIDS			1	1		_	*	•
. HUBTOTAL, DRY			2,742.3	57,671	1,659.92	68,833.	2,153.9	78,502
WATER	н,о	18.016	0.9986	17.99	153.4	2,764,	147.3	2,654
TOTAL, WET	ı .		2,743.6	57,689	1,013,32	71,597.	2.301.2	81,156
CAS MONECULAR WEI	GHT		21.0	0'		8	35.3	
TEMPERATURE, OF			09		120		001	
PRESSURE, PSIA			615	5	20	,	4.44	-
SCER			1,041,247	757	194,880	7	24427	
W.T.S						T	_	

TABLE 4.1 (continued)

Divinues of the PROCESS

F.1 PROCESS
INTEGRATED FACILITY/RODULE DEFINITION

STREAM NUMBER			6-1	<u> </u>	6-2	^		6-3 7
STREAM ID			OXIDANT TO GASIFIER	ASIFIER	INERT CAS TO COAL	- PREPARATION	AIR SEP	AIR SEPARATION VENT
			LB-FOLS	LES	1 B-:10LS	0 2	LB-FOLS	Liss
	SYMBOL	FORMULA WEIGHT	NH.	111111111111111111111111111111111111111				
CARBON	g	12.01	t i	1	4	t		***************************************
HEOFOGEN	H	2.016	1	1				
ОХХСЕИ	70	32.0	10.501.4	336.045				
HITROGEN & ARGON	N 2	28,016	214.3	6,004	3,000	84,048	36,270	1,016,140
SULFUR	5.	32.06		•	ŧ			
CHLORINE	73	35,453				+		
CAPBOR MONOXIDE	00	28.01	ı		•			
CARBON DIOXIDE	505	41.01	1		ı			
PETTIANE	CII	16.042	l	1		_	1	
ETHYLENE	C. II.	28.052	1	•	-		-	
1 THANE	CZHZ	30.068	-	•	1		,	
PESPILENE	CZHO	42.078	ı	ı	•	ı		
PEOPANE	9 H C	44.094	•	1	t	-		-
HTDROGEN SULFIDE	H 35 W	34.076		1	1		•	-
	cos	69.075	•	1		1	•	ı
ш	CS,	76,13	•	•		•		-
	505	64.06			1	ı	1	
KITROUS OXIDE	130	30.608				-	-	
	NH.	17.031		_	1	1		
CYANIDE	нси	27.026	ı,	1	1	1		-
HYDROGEN CHLORIDE	нсл	36,461	1		*	1		-
RAPHTHA				1	4		-	
TAR & OIL	1		ŀ	t	1			
ASH	ı		1	1	-			•
CTHER SOLIDS				•	1 1	ı	•	
SURTOTAL, DRY			10,715.7	342,049	3,000	84,048	36,270	1,016,140
интек	ور عر	18.016	1	1	0	0		
FOTAL, WET	ı		10,715.7	342,049		84,048	36,270	1,016,140
LECULA	I.I.		31.92		28.0		28.0	
TEMPERATURE, OF			071		06		98	
PPESSURE, PSIA			21.6		14.4		991	9
54. FB			7,890,4	98	1,140,	8	44.44	
W.25	:		1	-			13,782,600	004

MAYER A.1 (continued)

	DEFT. TOOM
	• :
PROCESS	THEFGRAPHS PACILITY/NODE C DOFF. NOW
- ¥	THYEGRATED

STREAM ID STRE	STREAM NUMBER			1-1		(11-1)	(12-1)
SVHBOL FORMULA WEIGHT Link List			+ -;	SOUR CAS TO ACTO	GAS REMOVAL	COAL TO COAL PREPARATION 1.B-MOLS Les	S TO DIS
C		SYMBOL		LIN	H.		
Coloration 1	ROUNT	U	12.61	•	•	253,637	
SERIFFE SERI	EN	Ε,	2.016	8,125,4	16, 381.	17,925	
A		D2	32.0	0 007	776.11	5.761	
The matching column	N C ARGON		28.016	419.9	111/04:	15,449	
Main Monoxide Co.		ماد	32.00			767	9.00
December 2002 11.0.2 12.483.2 1092.286. 12.5 12.005. 12.5 12.005. 12.5 12.005. 12.5 12.005. 12.5 12.005. 12.5 12.005. 12.5 12.005.	AFBON MONOXIDE	30	28.01	17,405.8	487,536.		ı
10, 0.4 10,	ARBON DIOXIDE	တ	9.1.6	2.483.2	109,286.		
Figure C C C C C C C C C	LTHANE	CH	16.642	125.	2,005.		
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No.		CATA	30.068	1	-		t
A		2 I I	42.078	_			
STATE STAT	-) E. C.	44.094	3 667			
NH DISCULPIDE CS. 76.13 NH DISCULPIDE CS. 76.13 NH DISCULPIDE CS. 76.13 NH DISCULPIDE NO. 64.06 NGS OXIDE NO. 64.06 NGRA CANIDE HCI 27.026 NGRA CHICARIDE HCI 36.461 NGRA CHICAR		H 25	34.076	433.4	2 926		
No DOXIDE SO			76 13				
17.031 17.031 17.031 17.031 17.031 17.031 17.031 17.031 17.031 17.031 17.031 17.031 17.036 1		205	64.06		1		
17.031		≘	30,008		1		
SGEN CYANIDE HCd 27,026		NH.	17,031	•			
CEN CHIORIDE HC1 36,461	TUROGEN CYANIDE	HCr	27.026				
F SOLIDS TALL, DRY H 20 18,016 29,041,2 644,660, 76,817 R SOLIDS TALL, DRY H 20 18,016 29,082.8 645,409, 70,15CUIAR WEIGHT 100, 11,041,832.	2	HC1	36.461	1	1		
R SOLIDS TAL, DRY H, WET A'LIECULIAR WEIGHT SURE, PSIA 11,041,832.	THA THA			-	i	F	
0 18,016 29,041,2 644,660, 376,817 29,082,8 645,409, 416,670 100, 100, 100, 11,041,832.	AR & OIL	-				059 65	59.650
0 18,016 29,041,2 644,660, 376,817 41.6 749, 39,853 22,19 645,409, 416,670 100, 100, 100, 11,041,832.		-				22017	14,334
0 18,016 41.6 749. 39,853 29,082.8 645,409. 416,670 100	HIER SOLIDS			2 170 66	644.660.	376,817	73.984
29,082,8 645,409, 416,670 98,6 20,19 100. 665.	ATER	H ₂ O	18,016	41.6	749.	39,853	24,661
100. 100. 111,041,832.	OTAL, WET	7		29,082.8	645,409,	416,670	98,645
100. 665. 11,041,832.	AS MOLECULAR WEIGH	H				2	
905): 11,041,832.	EMPERATURE, OF			100.		1	
	RESSURE, PSIA			11 041 8	337	1	
				-			67

TABLE 4.1 (continued)
MATERIAL BALANCE
K-T PROCESS
INTEGRATED PACILITY/MODULE DEFINITION

	ORIGINAL PAGE IS	TAB MX: K INTEGRAT	TABLE 4.1 (continued) MATERIAL BALANCE K-T PROCESS INTEGRATED PACILITY/MODULE DEFINITION	
STREAM NUMBER		(1-61)	(1-91)	(16.2)
STREAM ID		10 STORAG	ING WATER	SE SE
ZYMBOL	BOL FORMULA WEIGHT	LB-MOLS LBS	LB-MOLS LBS	H H HE
CARBON	12.01			
	32.0			
NITROGEN 6 ARGON N2	28.036	1.64.24		
CHLORINE C1	32,06			
2		•		
CARBON DIOXIDE CO.	14.01			
ETHYLENE				
ETHANE				
PROPANE	44.094			
CARBONYI SHIPTOP CAS				
	76.13			
SULPUR DIOXIDE SO				
	10.008	ė		
N CHLORIDE	36.461			
TAP COTT				
ASH				
ER SOLIDS				
SUBTOTAL, DRY		476.4 15,273	686 790	20.740
	877.87			26.102
GAS MOLECULAR WEIGHT		6/2*61		
TEMPERATURE, OF		250		
PRESSURE, PSIA				
MAD		1	1,370	140
		***************************************	**************************************	

TABLE 4.2 MODULE ENERGY BALANCE, 10⁶ BTU PER HOUR %-T PROCESS

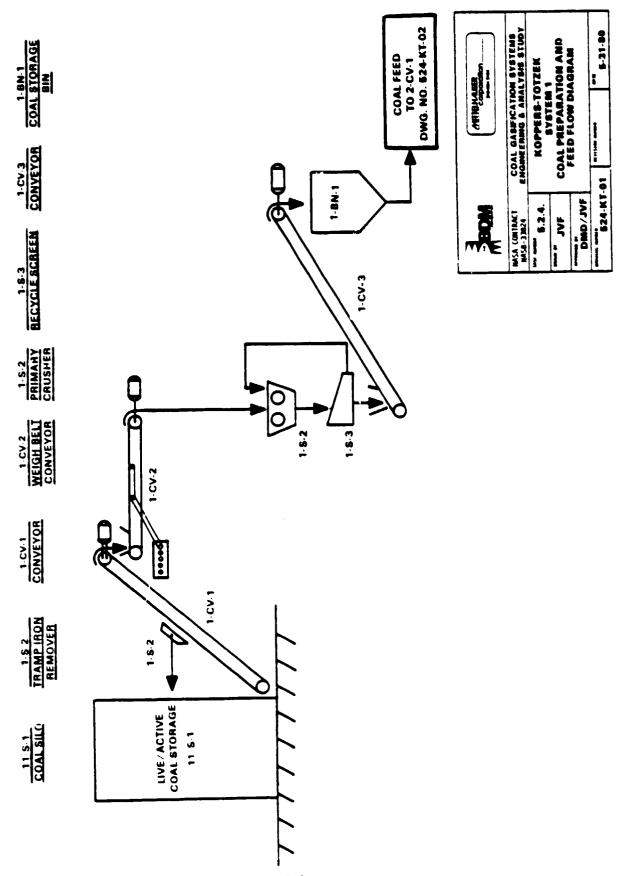
TOTAL.	4575.	4594.5		2856.5	92.1 32.0	3597.2	362.3	-1449.9 -25.0 -1112.6
SENSIBLA	200			300	\$3.9 17.0 0.5			-336.5
LATENT	19.5			~	36.2			-1113.4
ÄHH	4575.			2856.5	3			
PHASE	Solid Cas Liquid			2m2 5c[02 5c]	644 644 6014			
EMTHALFT BALANCE IMLETS	STPEAN COM! Air to Air Sepin	SUBIOTAL IN	OUTLETS	STPLAM MEG Ash	Astr Septo Vent Flug Gan Sulfur		HEAT AND SHAFT WING ELECTRIC POSER SHAFT FORM	HELT RESETTION HEAT THEUT (STEAN) KAUTATION/FUGITIVE LIGSS SUBSOTAL MET VALUE TOTAL ABSOLUTE VALUE

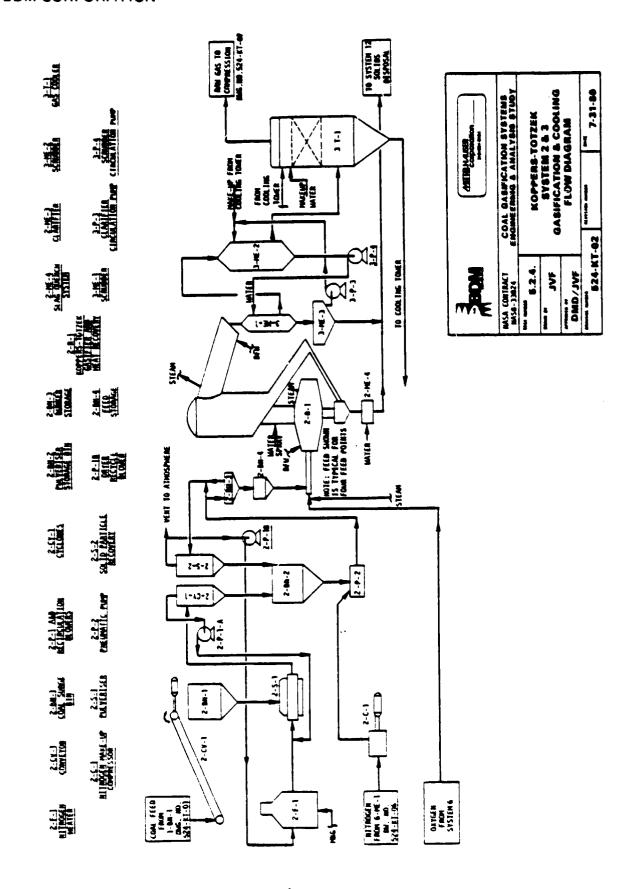
SALZBEE: BET EVIDALENY ENTY HEAT & SHAPT WORD X 1007 × 10.43
TOTAL ABSOLUTE VALUE OF HIAT & SHALL BONY

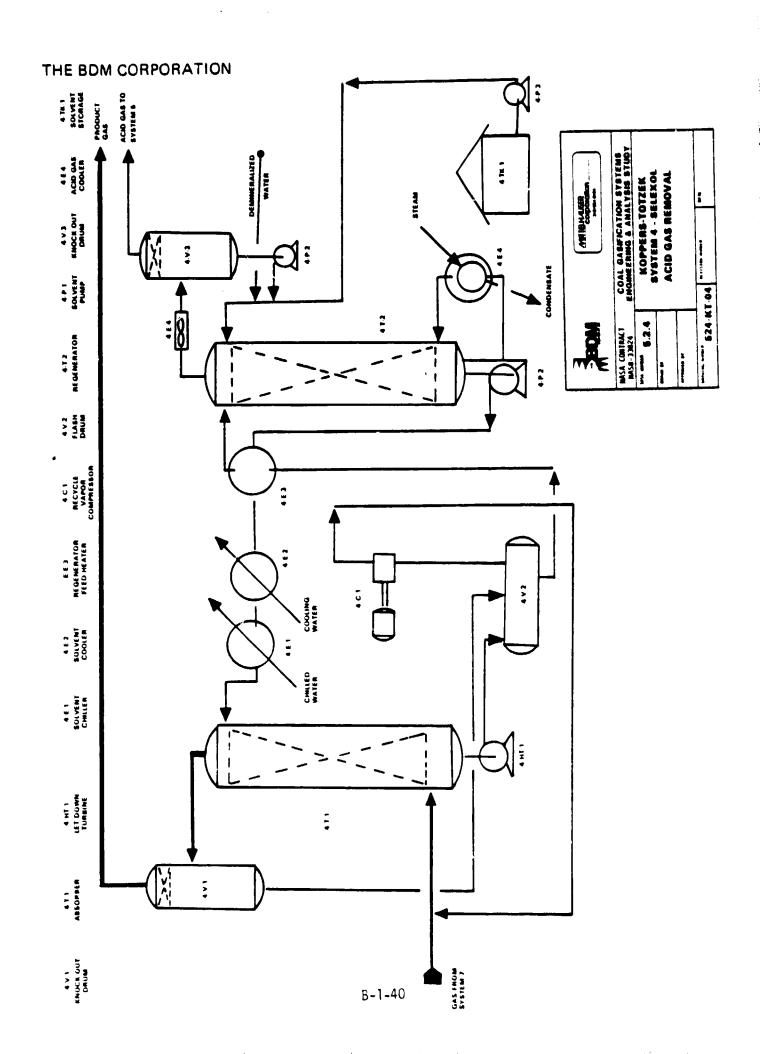
EASIS: 60°F, LIGGID WATER

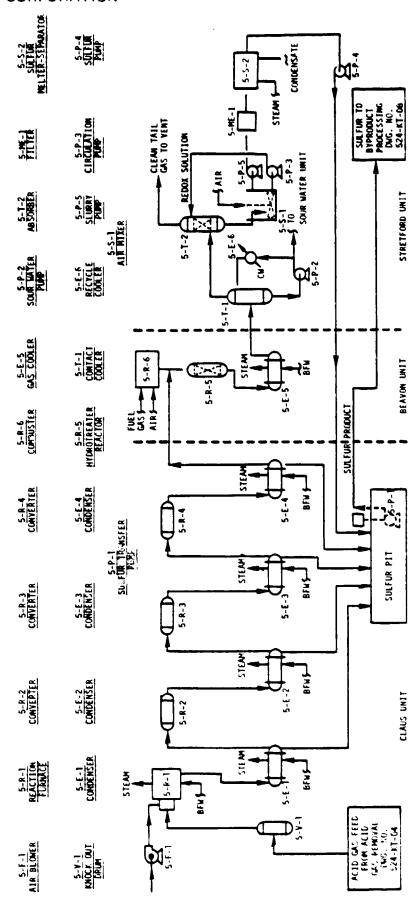
5.0 PROCESS FLOW DIAGRAM AND EQUIPMENT LIST

This section contains the process flow diagrams and the list of equipment for each module. The facility contains four times the quantity shown here except for the coal silos which are on a facility basis and System 5, which has two trains in Module 1 and only one train in subsequent modules.









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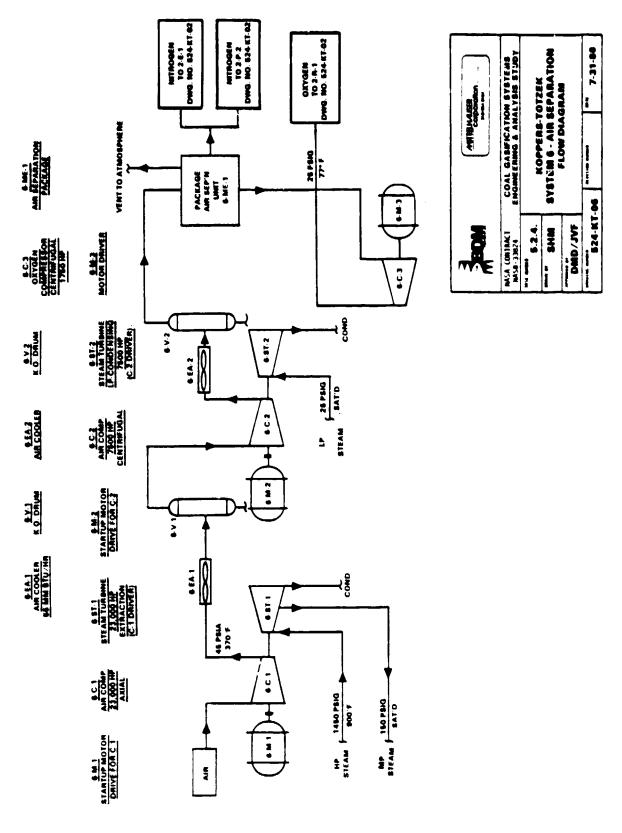
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COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY

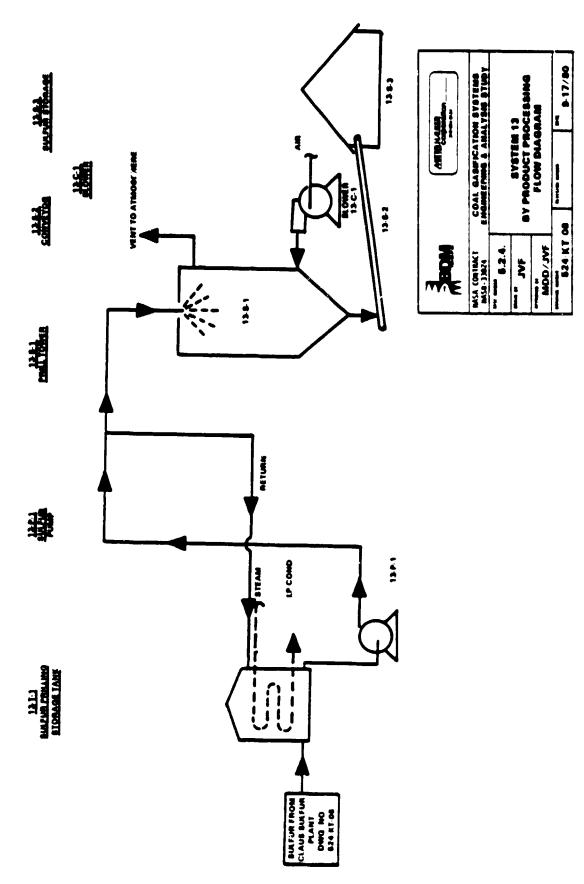
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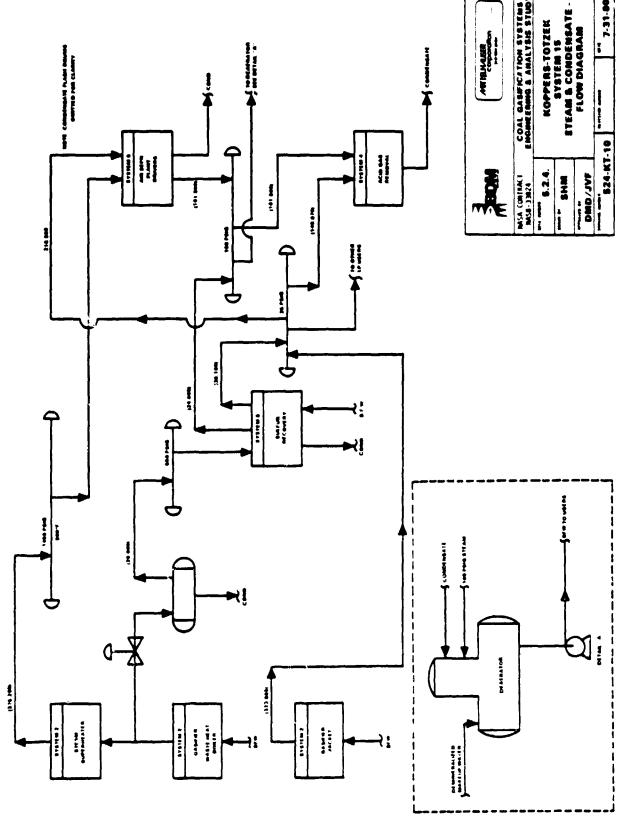
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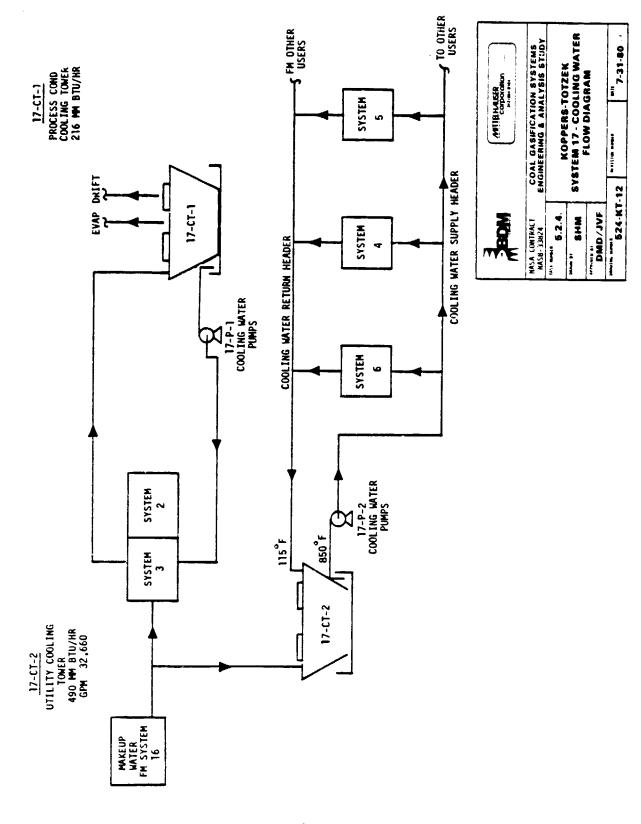
SYSTEM 5 SULFUR RECOVERY



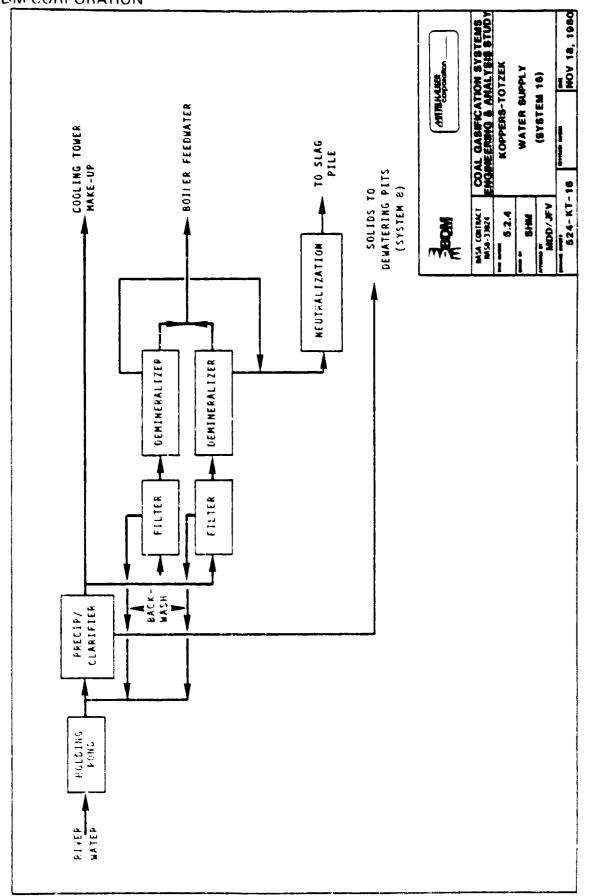
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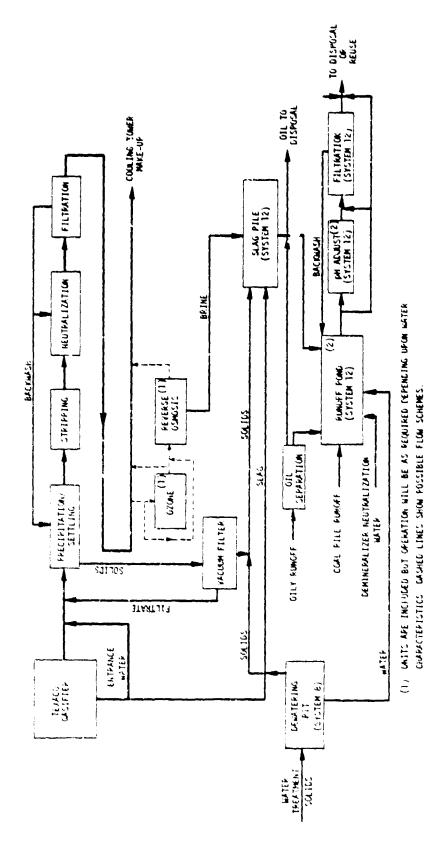






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MSA CONTRACT

MSSA CONTRACT

MSS-33824

ENGINEERING & ANALYSIS STUDY

KOPPERS-TOTZEK

WASTEWATER TREATMENT

(SYSTEM 18)

MDD/JVF

MDD/JVF

MDD/JVF

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B-1-48

THE BDM CORPORATION TVA MBG MODULE KOPPERS-TOTZEK SYSTEM NO. 1 COAL PREPARATION AND FEEDING

ITEM NUMBER	SERVICE	QUA		DESCRIPTION
1-\$-1	LIVE COAL	SPARE	OPERATION	·
	SILO	0	4	SILO
Particular to the constitution				11.750 TONS/ EACH
I-S-I	TRAMP IRON REMOVER	0	7	MAGNETIC
-EV-1	LIVE COAL CONVEYOR	0	1	CONVEYOR
I-EV-2	WEIGH BELT CONVEYOR	0	1	SCALE AND CONVEYOR
 I-BN-1	SIZED COAL STORAGE BIN	0	2	
I-S-2	PRIMARY CAOL CRUSHER	0	1	• <u>-</u>
1-CV-3	SIZED COAL CONVEYOR	0	1	CONVEYOR
1 - S-3	RECYCLE SCREEN	0	1	
	!			
		İ		

THE BDM CORPORATION TVA MBG MODULE KOPPER-TOTZEK SYSTEM NO. 2 GASIFICATION

ITEM	SERVICE	QU	ANTITY	Beage
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
2-R-1	KOPPERS-TOTZEK GASIFIER	1	8	
2-P-1	RECIRCULATION BLOWER	9	8	
2-P-2	TRANSFER NITROGEN BLOWER	9	8	
2-P-3	CLARIFIER CIRCULATION PUMP	9	8	-
2-P-4	SCRUBBER CIRCULATION PUMP	9	8	
2 - F-1	TRANSPORT NITROGEN HEATER	1	8	. <u></u> ., <u>.</u>
2-CV-1	CRUSHED COAL CONVEYOR	1	8	
2-BN-1	PULVERIZER SURGE BIN	1	8	
2- BN-2	PULIVERIZED COAL STORAGE BIN	,	8	

TVA MBG MODULE
KOPPERS-TOTZEK
SYSTEM.NO. 2

ITEM	SERVICE		NTITY	DESCRIPTION
NUMBER	JUNTION	SPARE	OPERATION	DESCRIPTION
- BN- 3	BUNKER STORAGE	1	8	
- BN- 4	REACTOR FEED STORAGE	1	8	
-CY-1	CYCLONES	2	16	and the state of t
?-S-1	COAL PULIVERIZER	1	8	Annual Company of the Company
?-S-2	SOLID PARTICLE RECOVERY	2	16	
2-ME 1,2	RAW GAS SCUBBERS	2	16	
2-ME-3	CLARIFIER	1	8	
3-T-1	RAW GAS COOLER	1	8	

TVA MBG MODULE K-T GASIFIER SYSTEM NO. 4 - ACID GAS REMOVAL

ITEM	SERVICE	A con-	ANTITY	DESCRIPTION
NUMBER		SPARE	OPERATION	DESCRIPTION
4-V-1	SOLVENT KNOCK-OUT DRUM	0	1	
4-V-2	SOLVENT FLASH DRUM	0	1	
I-V-3	REGENERATOR KNOCK-OUT	0	1	
4-T-1	ACID GAS ABSORBER	0	1	
4-T-2	SOLVENT REGENERATOR	0	1	· •··
4-E-1	SOLVENT CHILLER	0	1	SHELL & TUBE EXCHANGER
4-E-2	SOLVENT COOLER	0	1	SHELL & TUBE EXCHANGER
4-E-3	REGENERATOR FEED PREHEAT EXCHANGER	0	1	SHELL & TUBE EXCHANGER
4-E-4	ACID GAS COOLER	0	1	AIR FAN EXCHANGER
4-E-5	REGENERATOR REBOILER	0 B-1-52	1	SHELL & TUBE EXCHANGER

TVA MBG MODULE K-T GASIFIER SYSTEM: NO: 4 ACID GAS REMOVER

ITEM	SERVICE	1	ANTITY	DECORTOTION
NUMBER		SPARE	OPERATION	DESCRIPTION
4-P-1	SOLVENT CIRCULATION PUMP	1	1	
1-P-2	REGENERATOR REFLEX PUMP	1	1	
1-P-3	SOLVENT TRANSFER PUMP	1	1	
4-TK-1	SOLVENT STORAGE TANK	0	1	······································
4-ME-1	TEG GAS DRYING PACKAGE	0	1	INCLUDES TEG CIRCULAT PUMP, TEG CONTRACTOR AND TEG REGENERATOR
4-C-1	RECYCLE VAPOR COMPRESSOR	0	1	•
4-HT-1	RICH SOLVENT LET DOWN TURBINE	0	1	
			OR C	HNAL PAGE IS POOR QUALITY

TVA MBG MODULE K-T GAS1FIER SYSTEM NO. 5 SULFUR RECOVERY & TAIL GAS CLEAN-UP

ITEM	SERVICE	QUA		
NUMBER		SPARE	OPERATION	DESCRIPTION
5-P-1	SULFUR PRODUCT PUMP	1	1	
5-P-2	STRETFORD SOUR WATER PUMP	1	1	<u> </u>
5-P-3	REDOX SOLUTION CIRCULATION PUMP	1	1	
5-P-4	STRETFORD SULFUR TRANSFER PUMP	1	1	
5-P-5	REACTION FURNACE AIR BLOWER	1	1	
5-S-1	AIR OXIDATION MIX TANK	0	1	
5-S-2	STRETFORD SULFUR SEPARATOR MELTOR	0	1	
5- S-3	MOLTER SULFUR STORAGE PIT	. 0	1	·

TVA MBG MODULE

KOPPERS-TOTZEK SYSTEM NO. 5 SULFUR RECOVERY & TAIL GAS CLEAN-UP

ITEM	SERVICE	QUANTITY		DECEDIBLION	
NUMBER		SPARE	OPERATION	DESCRIPTION	
5-V-1	ACID GAS KNOCK - OUT DRUM	0	1		
5-R-1	REACTION FURNACE	0	1	ACID GAS BURNER AND STEAM GENERATION	
5-R-2,3,4	SULFUR CONVERTERS	0	3	HORIZONTAL REACTORS	
5-R-5	BEAVON HYDROTREATOR REACTOR	0	1	HORIZONTAL GAS HYDROLYSIS REACTOR	
5-R-6	COMBUSTOR	0	1	HOT GAS GENERATOR	
5-T-1	BEAVON UNIT CONTACT COOLER	0	1		
5-T-2	STRETFORD ABSORBER	0	1		
5-E-1,2,3,4	SULFUR CONDENSORS	0	4	SHELL & TUBE EXCHANGER	
5-E-5	BEAVON UNIT GAS COOLER	0	1	SHELL & TUBE EXCHANGER	

NOTE: THIS ENTIRE UNIT IS DUPLICATED IN MODULE 1 ONLY.

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TION TVA MBG MODULE KOPPERS-TOTZEK SYSTEM NO. 5 SULFUR RECOVERY & TAIL GA'S CLEAN-UP

ITEM			TITY	000000000000000000000000000000000000000
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
5-E-6	STRETFORD RECYCLE COOLER	0	1	
5-P-1	LIQUID SULFUR PUMP	1	1	SUMP PUMP
5-P-2	SOUR WATER CIRCULATION PUMP	1	1	
5-P-3	REDOX SOLUTION CIRCU- LATION PUMP	1	1	
5-P-4	LIQUID SULFUR PUMP	1	1	
5-P-5	SULFUR SLURRY PUMP	1	l	MOYNO PUMP
5-ME-1	FILTER	0	1	SULFUR SLURRY FILTER

TVA MBG MODULE KOPPERS-TOTZEK

SYSTEM NO. 6 - AIR SEPARATION

1 EM	SERVICE	QUANTITY		DECEDITION	
NUMBER	JEN 10L	SPARE	OPERATION	DESCRIPTION	
6-ME-1 A/B	AIR SEPARATION PLANT	0	2	PACKAGE SYSTEM 2052 TPD 95% 0 ₂	
6-C-1 A/B	AIR COMPRESSOR	0	2	AXIAL COMP. 23,000 H.P.	
6 -C-2 A/B	AIR COMPRESSOR	0	2	CENTRIFUGAL COMP. 7500 H.P.	
6-C-3 A/B	O ₂ COMPRESSOR	O	2	CENTRIFUGAL COMP. 1750 H.P.	
6-ST-1 A/B	STEAM TURBINE DRIVER	0	2	DRIVER FOR C-1 CONDENSING/EXTRACTION H.P. TURBINE	
6-ST-2 A/B	STEAM TURBINE DRIVER	v	2	DRIVER FOR C-2 CONDENSING L.P. TURBINE	
6-M-1	ELECTRIC MOTOR	0	1	START-UP MOTOR FOR AIR PLANT C-1 COMP.	
6-M-2	ELECTRIC MOTOR	0	1	START-UP MOTOR FOR AIR PLANT C-2 COMP.	
6-M-3	ELECTRIC MOTOR	0	2	DRIVER FOR C-3	

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TVA MBG NODULE KOPPERS-TOTZEK

SYSTEM NO. 6 - AIR SEFARATION

				•
1TEM	SERV1C E	QUAN	TITY	DECCRIPTION
NUMBER		SPARE	OPERATION	DESCRIPTION
6 -EA-1	AFTER COOLER	0	2	AIR COOLED EXGR. Q = 55 NMBTU/HR
6-EA-2	AFTER COOLER	0	2	AIR COOLED EXGR. Q = 55 MMBTU/HR
6-V-1	C-1 COMPR. K.O. DRUM	0	2	
6-V-2	C-2 COMPR. K.O. DRUM	0	2	

B-1-58

TVA MBG MODULE KOPPERS-TOTZEK SYSTEM NO. 7 - GAS COMPRESSION

ITEM	SERVICE	QUAN	TITY	DECONTACTION	
NUMBER		SPARE	OPERATION	DESCRIPTION	
7-C-1	PDT GAS COMPRESSOR	0	1	38,500 H.P. AXIAL MACHINE	
7-C-2	PDT GAS COMPRESSOR	0	1	34,600 H.P. DOUBLE SUCTION CENTRIFUGAL	
7-C-3	PDT GAS COMPRESSOR	0	1	21,300 H.P. CENTRIFUGAL	
7-M-1	ELECTRIC MOTOR	0	1	DRIVER FOR C-1	
7-M-2	ELECTRIC MOTOR	0	1	DRIVER FOR C-2 & C-3	
7-EA-1	C-1 COMPRESSOR AFTERCOOLER	0	1	AIR COOLED HEAT EXCHANGER Q = 100 MMBTU/HR	
7-EA-2	C-2 COMPRESSOR INTERCOOLER	0	2	AIR COOLED HEAT EXCHANGER Q = 48 MMBTU/HR	
7-EA-3	C+2 COMPRESSOR AFTERCOOLER	0	1	AIR COOLED HEAT EXCHANGER Q = 43 MMBTU/HR	
7-EA-4	C-3 COMPRESSOR	0	1	AIR COOLED HEAT EXCHANGER Q = 54 MMBTU/HR	
L	<u> </u>	R_1_50	L	Land the second	

TVA MBG MODULE KOPPERS-TOTZEK SYSTEM NO. 7 - GAS COMPRESSION

ITEM	SERVICE	QUAN'	TITY	05500107104
NUMBER		SPARE	OPERATION	DESCRIPTION
7-E-1	TRIM COOLER	.0	1	SHELL & TUBE HEAT EXCHANGER Q = 6 MMBTU/HR
7 - V-1	CONDENSATE K.O. DRUM	0	1	
7-V-2 A/B	CONDENSATE K.O. DRUM	0	2	
7-V-3	CONDENSATE K.O. DRUM	0	1	
7-V-4	CONDENSATE K.O. DRUM	0	1	
	L	1 60	<u> </u>	

TVA MBG MODULE KOPPERS-TOTZEK SYSTEM NO. 13 - BY-PRODUCT PROCESSING

ITEM	SERVICE	QUAN	TITY	2555077774011
NUMBER		SPARE	OPERATION	DESCRIPTION
13-T-1	MOLTEN SULFUR STORAGE TANK	0	2	STEAM HEATED 500 TONS EACH
13-P-1	PRILL TOWER FEED PUMP	1	1	MOLTEN SULFUR PUMP
13-5-1	PRILLING TOWER	0	1	
13-S-2	PRILLED SULFUR CONVEYOR	0	1	
13-0-1	AIR BOILER	1	1	
13-5-3	PRILLED SULFUR STORAGE	0	1	5000 TONS STORAGE BUILDING

TVA MBG MODULE KOPPERS-TOTZEK SYSTEM NO. 17 - COOLING WATER

	3,3,6,1 10	QUANTITY		
ITEM NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
	 			
17-CT-1	PROCESS CONDENSATE COOLING TOWER	0	1	2 CELLS Q = 215 MMBTU/HR GPM BY K-T
17-CT-2	UTILITY COOLING TOWER	0	1	8 CELLS Q = 490 MMBTU/HR GPM = 32,660
17-P-1 A - E	COOLING WATER CIRCULA- TION PUMPS	1	4	
17-P-2 A - S	COOLING WATER CIRCULA- TION PUMPS	3	16	1,450 H.P.
			<u> </u>	<u></u>

TVA MRG MODULE KOPPERS-TOTZEK SYSTEM NO. 17 - COOLING WATER

ITEM	CERVICE	QUANTITY		050001071011	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION	
17-CT-1	PROCESS CONDENSATE COOLING TOWER	0	1	2 CELLS Q = 215 MMBTU/HR GPM BY K-T	
17-CT-2	UTILITY COOLING TOWER	0	1	8 CELLS Q = 490 MMBTU/HR GPM = 32,660	
17-P-1 A - E	COOLING WATER CIRCULA- TION PUMPS	1	4		
17-P-2 A - S	COOLING WATER CIRCULA- TION PUMPS	3	16	1,450 H.P.	

APPENDIX B-2

TEXACO COAL GASIFICATION PLANT REFERENCE FACILITY DESIGN

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10

1.0 INTRODUCTION

The United States, after a number of years of development based on plentiful and inexpensive oil and natural gas, is entering a period of time when it is essential to supplement these energy sources by the increased use of coal. Coal is the nation's most plentiful fossil fuel. Coal gasification is a means of accomplishing this. While utilization of coal through conversion to gaseous products is not new there is no industry within the US which might serve as a base for establishing cost, operational reliability and requirements, and design data for the large scale environmentally acceptable plants needed.

The Tennessee Valley Authority with systems engineering and analysis support from the George C. Marshall Space Flight Center has initiated a project which would establish the commercial base and demonstrate the requirements for gasifying coal in a large integrated facility. The project consists of gasifying 20,000 tons per day of Eastern coal in a four module plant-the construction of which is staggered to accommodate efficient use of construction manpower and product market development.

The purpose of this study is the systematic engineering and cost analysis of five selected gasifier technologies and the other associated systems required to produce medium BTU gas for delivery into a 600 psig distribution system in Northern Alabama. This effort in a series of design studies provides a reference basis for selection of the final plant configuration and choice of process technologies.

The methodology consists of selection of a systematic breakdown of the facility for purpose of study, identification of available technologies for each system, development of a definition level design and cost analysis for each system, conducting a series of trade studies using the definition design as a basis, and finally development of reference facility designs and cost analyses based on the previous work.

This report covers the work done in establishing a Reference Facility Design based on Texaco coal gasification technology. It is one of three such designs, the other two being based on Babcock and Wilcox and Koppers-Totzek coal gasification processes. The design covers a complete grass roots facility site at Murphy Hill, Alabama. The plant inputs are coal, water and electricity. The plant outputs are medium BTU gas (MBG) and solidified sulfur.

The design approach involved breaking each of four identical modules into process and support systems followed by analysis of each system through identification of available technologies and use of the comparative trade studies to select the preferred processes for each system.

2.0 SUMMARY

A four module, 20,000 TPD, based on Texaco coal gasification technology has been designed. The plant processes Kentucky No. 9 coal with provisions for up to five percent North Alabama coal. Coal transportation is by river barge except for the Alabama coal which is trucked to the site. Medium BTU gas with heat content of 291 BTU/SCF and not more than 200 ppm sulfur is the primary plant product. Sulfur is recovered for sale as prilled sulfur. Ash disposal is on site. The plant is designed for zero water discharge. Trade studies provided the basis for not using boiler produced steam to drive prime movers. Thus process derived steam in excess of process requirements is superheated for power use in prime movers. Electricity from the TVA grid is used to supply the balance of the plant prime mover power requirements. A study of the effect of minemouth coal cleaning showed that coal cleaning is not an economically preferred route.

The plant design was arrived at by a systematic procedure based on published design work, process trade studies, team engineering experience, a NASA provided module level definition of some twenty systems, and the TVA document, "Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," March, 1980.

The overall plant configuration is shown schematically in Figure 2.1. As shown, General Facilities, Instrumentation and Control, and Coal Handling serve plant wide functions. All other plant components are contained in four identical process modules. The Instrumentation and Control System operates to monitor overall plant performance and to control intermodule relationships. The total list of systems is given in Table 2.1.

The design procedure involved defining available processes to meet the requirements of each system, technical/economic trade studies to select the preferred processes, and engineering design and flow sheet development for each module. Cost studies assumed a staggered construction schedule for the four modules beginning spring 1981 and a 90 percent on stream factor.

The basis and results of the design study are given in Tables 2.2 and 2.3.

TABLE 2.1. LIST OF SYSTEMS

SYSTEM NO.	NUMBER OF PER MODULE	COST UNITS PER FACILITY	SYSTEM DESCRIPTION
1	1	4	COA! PREPARATION AND FEEDING
2	4	16	GASIFICATION
3	4	16	INITIAL GAS CLEANUP AND COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	3	10	AIR SEPARATION
7	1	4	COMPRESSION
8	1	4	PROCESS SOLIDS TREATMENT
10	0	1	INSTRUMENTATION AND CONTROL
11	0	1	COAL HANDLING
12	0	1	SOLIDS DISPOSAL
13	1	4	BY-PRODUCT PROCESSING
14	1	۵	PLANT POWER SYSTEM
15	1	4	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKE-UP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	0	1	GENERAL FACILITIES

TABLE 2.2. DESIGN BASIS SUMMARY*

PLANT CAPACITY

PLANT TYPE COAL TYPE

PLANT SERVICE FACTOR

PLANT LIFE

ELECTRICITY SOURCE

WATER

COAL RECEIPT
PRODUCT DELIVERY

PRODUCT QUALITY

SULFUR

COAL COST (1980) LAND COST (1980)

CLEARING AND GRUBBING (1980)

ELECTRICITY COSTS (1980)

BY-PRODUCT CREDIT

ESCALATION

20,000 TPD

FOUR INDEPENDENT MODULES

KENTUCKY NO. 9

90 PER CENT

20 YEARS EACH MODULE

TVA GRID-4.16KV, 6.9KV, 13.8KV

TENNESSEE RIVER

BARGE +5% BY TRUCK

MBG AT 600 PSIG

MINIMUM 285 Btu/SCF

PRILLED

\$1.25/MM Btu \$3,000/ACRE

\$2,000/ACRE

NONE

AS PER TVA SPECIFICATION

^{*}Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," Tennessee Valley Authority, March, 1980

TABLE 2.3. DESIGN STUDY RESULTS*

FEED COAL WATER PURCHASED ELECTRICITY MBG PRODUCT MBG PRODUCT MBG QUALITY SULFUR PRODUCT (PRILLED) SULFUR PRODUCT TOTAL CAPITAL REQUIREMENTS OPERATING AND MAINTENANCE COSTS COAL, CATALYST, CHEMICALS	1,078 354 291 668 222,000 \$2,091 \$128	GPM MM KWHY M MSCFD MM MSCFY BTU/SCF LTPD LTPY
PLANT OPERATING STAFF	271	PERSONS
MAINTENANCE/YEAR	\$57	MM

^{*} COSTS ARE IN 1980 DOLLARS

Designs in this study were based on published works, engineering experience and judgments, and consultations with appropriate process and equipment vendors followed by design calculations, flow process sheet development and documentation.

Cost estimates are based on normal conceptual level design estimating procedures as described in the "Coal Estimation and Economic Evaluation Methodology," BDM/W-80-258-TR-RV2, submitted as another part of this project work scope and reprinted as Appendix E. Schedule estimates are in accordance with overall guidelines furnished by NASA.

Details of the schedule are considered typical for projects of this type and while they do not represent a detailed analysis specific to this project they represent the complexity and illustrate schedule constraints associated with the subject project. Process flow diagrams are included for the main process systems, air separation system, cooling water system, and steam distribution system as aids in following written system descriptions.

Technical results of the study are reported in this appendix. A cost analysis is included as a part of Appendix D.

3.0 OVERALL FACILITY DESCRIPTION

This section presents and discusses the process design of a plant designed to gasify 20,000 tons per day of coal in four 5000 TPD modules and to produce a medium-Btu product fuel gas (MBG) with a higher heating value of 291 Btu/SCF. The plant design is based upon using entrained flow of gasifiers as offered by Texaco, and upon the design criteria (coal characteristics, product specifications, site data, etc.) set forth by the Tennessee Valley Authority in March, 1980.

3.1 0 . 1 . Plant Processing

like overall plant configuration is shown in the block flow diagram, (see Figure 2.1). Kentucky No. 9 coal is received by barge and either placed in a dead storage pile or processed in System 11, Coal Handling. Provisions have also been made to receive up to five per cent of plant capacity by truck. Dead storage provides for 90 days interruption of supply.

Raw coal from System 11, Coal Handling, is ground and slurried with water. This slurry, containing 40 percent water, is pumped into three Texaco gasifiers per module operating at 690 psia pressure. The feed is partially combusted with oxygen from the air separation plant heating the mixture to approximately 2450°F. Ash becomes molten slag. Water contained in the feed is vaporized and reacts with the partial combustion products via the steam carbon and water gas shift reactions to produce a product rich in carbon monoxide and hydrogen. Raw gas is quenched within the reactor to 455°F. Molten slag is solidified in granules by the quenching operation and removed by lock hoppering along with quench water. After water-solids separation, the ash is disposed of on-site in a controlled disposal area. In the gas cooling and cleanup system, the gas is further cooled to 100°F before being treated in the acid gas removal system. The acid gas removal system uses the Selexol process (licensed by the Allied Chemical Corporation) in which a physical solvent (the dimethyl ether of polyethylene glycol) absorbs and removes hydrogen sulfide, carbonyl sulfide and the concomitant carbon dioxide from the raw gas.

The acid gas (hydrogen sulfide, carbonyl sulfide and carbon dioxide) is recovered from the Selexol solvent and processed in System 5, Sulfur Recovery and Tail Gas Treatment, wherein over 99 per cent of the hydrogen sulfide and carbonyl sulfide is converted into 668 long tons/day of by-product elemental sulfur. The sulfur recovery unit utilizes the conventional Claus process to convert about 90 to 95 per cent of the hydrogen sulfide (and some of the carbonyl sulfide) in the acid gas into by-product sulfur. The residual tail gas from the Claus unit is then processed in a Beavon-Stretford unit to recover additional by-product sulfur.

Oxygen utilized for the partial combustion of coal and recycle char in the gasifiers is produced in System 6, a conventional air separation plant. Atmospheric air is compressed to about 100 psia, cooled to 100 F and then separated cryogenically into oxygen and nitrogen at about 2 psig. The oxygen is compressed to about 815 psia for use in the gasifiers.

In addition to the systems contained in each module and described in Section 4, the plant consists of System 10, the overall plant monitoring and control system; System 11, the coal handling and storage system serving all four modules; and System 12, the solid waste disposal system serving the total facility.

The purpose of System 10 is to provide operational monitoring and supervisory master control of module operations. The system includes instrumentation to measure key characteristics of all major streams (mass, temperature, pressure, composition, density, and/or others as appropriate) and to display these characteristics in a central locati n. A dedicated computer is provided to record data of interest at the facility management level. Telecommunications capabilities are provided to communicate with module and systems operations as required to control total plant operation to meet delivery requirements.

System 11, the Coal Handling System, provides for the unloading of coal delivered to the plant either by barge or truck, reclaiming the coal from storage, reducing the size of coal, and transporting the coal to Coal Preparation and Feeding, System 1.

Raw coal is unloaded from barges by the barge unloading subsystem which is designed to unload up to ten 1500-ton capacity barges per shift. The coal is unloaded at an average rate of 1200 tons per hour on a five-day week basis. Coal is transferred by conveyor to a radial stacker which then produces a kidney-shaped coal pile containing live and dead storage. Coal is reclaimed from live storage and conveyed to a Bradford breaker where it is reduced in size to 2" x 0. Coal from trucks is unloaded into a chute from which it is conveyed to the Bradford breaker. Crushed coal from the Bradford breaker is transported to day storage silos from which it is transferred via vibrating belt conveyors to coal preparation and feeding, System 1, for further processing.

The purpose of System 12 is to store solid waste generated by facility operation during the facility life.

This system consists of a lined impounding pit sized to contain 20 years of solid waste. It includes the conveyor system to move the solids to the pit and a leachate recovery area to recover and pump leachate to the process condensate system.

Module descriptions are the subject of Section 4.

3.2 Facility Balances

The key material inputs and product outputs for the overall plant may be summarized as:

INPUTS:	Raw Coal	20,000 short tons/day
	98% Oxygen	17,080 short tons/day
	Day Waton Intako	10 000 apm

Imported Electric Power 198 MW

OUTPUTS: Product Gas
By-product Sulfur
Solids to Disposal (Wet)

1,178 MM SCFD
668 long tons/day
3,312 tons/day

Solids to Disposal (Dry) 3,305 tons/day

Table 3.1 gives the thermal efficiency for the facility. Table 3.2 gives the final product characteristics.

TABLE 3.1 CONVERSION EFFICIENCY TEXACO PROCESS 20,000 TPD FACILITY

INPUTS	10° BTU/HR	PER CENT
O COAL TO FACILITY	18,300	
② ELECTRIC POWER TO FACILITY	675.6	
OUTPUTS		
3 MBG FROM FACILITY	13,050.8	
4 COAL FINES FROM FACILITY	-0-	
EFFICIENCY		
COAL-TO MBG (③ ÷ ①) x 100%		71.3
OVERALL PRODUCT EFFICIENCY 3 ÷ (1) + 2) x	100%	68.8
OVERALL FACILITY EFFICIENCY (3 + 4) ÷ (1 +	②) x 100%	68.8

TABLE 3.2. PRODUCT CHARACTERISTICS

	VOLUME PER CENT
CARBON MONOXIDE	51.2
HYDROGEN	37.1
NITROGEN AND ARGON	1.3
CARBON DIOXIDE	9.9
METHANE	0.5
	100.0
HYDROGEN SULFIDE	48 PPM VOLUME
CARBONYL SULFIDE	161 PPM VOLUME
TOTAL SULFUR	209 PPM VOLUME
WATER VAPOR	7 LBS/MM SCF
HIGHER HEATING VALUE	291 BTU/SCF
TEMPERATURE	75 ^O F
PRESSURE	615 PSIA

4.0 MODULE DESCRIPTIONS

Each of the four modules in the plant consist of identical sequences of process systems as given in Table 2.1. Process flow diagrams and an equipment list are included in Section 5 for the principal process systems and such auxiliary systems, as cooling water and steam systems.

Each module receives 5000 TPD of coal from System 11, Coal Handling. The sequence of processing shown in Figure 4.1 is repeated in each of the four modules. Descriptions of each modular system are as follows.

4.1 System Descriptions

4.1.1 System 1 - Coal Preparation and Feeding

System 1, Coal Preparation and Feeding, receives 2"x0 coal feed from System 11, Coal Handling. Each module processes 5,000 TPD of ROM coal in one of two 100 percent units. This system, shown in Drawing Number 524-TX-O1, grinds coal and produces a 60 wt percent coal slurry which is fed to System 2, Coal Gasification. This design is patterned after the Parson's design in EPRI report AF-880, "Preliminary Design Study for an Integrated Coal Gasification Combined Cycle Power Plant."

ROM coal is received by conveyor from storage silos and passed through a magnetic tramp iron remover. A prescreening system selects the 2"xl/4" coal for further size reduction in a cage mill. The combined cage mill product and minus 1/4" coal from the prescreening operation are further processed in a wet rod mill. The discharge from the rod mill containing approximately fifty percent water is diluted and screened to produce the desired size consistancy with oversized material being recycled to the mill. A centrifuge is used to reduce the water content of the coal slurry. Final water content is controlled at 40 percent by water addition to the centrifuge product in a mix tank. Final slurry control is obtained by density measurement in the mix tank pump around loop. The slurry feed is transferred to and maintained in two parallel agitated tanks prior to transfer to a run tank in System 2, Coal Gasification. Alternate use of the two parallel tanks allows precise control of slurry content before transferring to the next system.

4.1.2 System 2 - Coal Gasification

The Texaco Coal Gasification Process uses a coal slurry feed, consisting of fresh ground coal together with recycled fine slag and carbon with a total solids content of 60% by weight. The slurry is pumped from mix tanks in the grinding and slurry section to the gasifier slurry tank. A circulating pump circulates the slurry through this tank and supplies slurry to the suction of the high pressure charge pumps. The charge pumps raise the slurry pressure sufficient to feed the gasifiers at 690 psia.

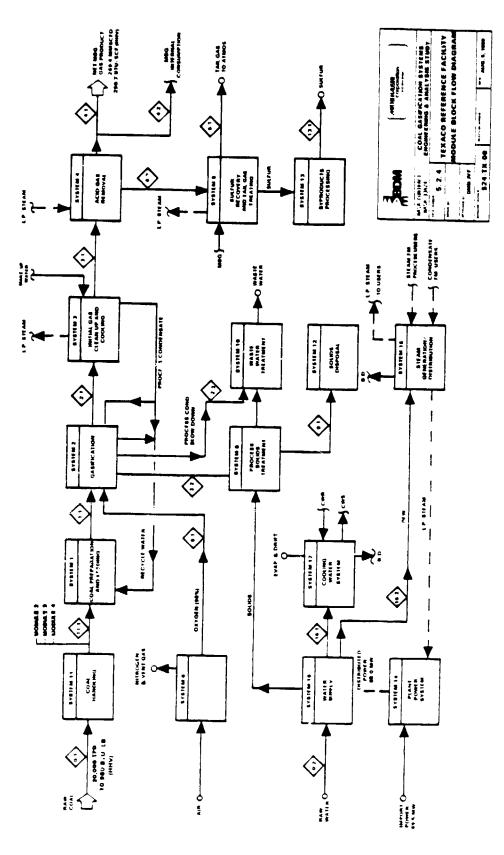


Figure 4.1

The coal-water slurry is fed through a specially developed burner into a refractory-lined gasifier reactor. Partial combustion with oxygen takes place at a pressure of 675 psig, and a temperature of 2450 degrees F to produce a gas consisting mainly of CO, H₂, CO₂, and steam. Most of the sulfur in the coal is converted to H₂S with the balance converted to COS. Nitrogen and argon from the oxygen feed appear in the gas together with most of the nitrogen from the coal. The gas contains a small amount of methane. Some unconverted carbon and all of the ash are removed in the form of slag. The gas is essentially free of uncombined oxygen.

The upper section of the gasifier is the refractory-lined chamber in which the partial oxidation reaction takes place.

The gas from the gas generator reaction section passes straight down into the quench section of the gasifier along with the molten ash where it is immediately quenched with water from 2450 to 455 degrees F. Raw gas is sent to System 3, where residual carbon is scrubbed and the gas is further cooled.

Most of the ash in the coal feed agglomerates into essentially carbon-free molten slag droplets, which are quenched and solidified in the lower quench section of the reactor. This slag is settled through the quench water into the lock hopper. The lock hopper is periodically dumped onto a screen, from which the slag is conveyed to the solids treatment system.

Water from the gasifier quench chamber is cooled and combined with water from the carbon-scrubber lower section. Both streams contain fine slag and unconverted coal. The water stream from the lock hopper and water from the final product cooler separator join this stream, and the total flows into a flash pot.

Separation of water and solids takes in place in a clarifier where the fine slag and unconverted coal settle out, leaving a clarified water overflow that is pumped back to the carbon scrubber via a gray water drum. Makeup water is added at this point.

The clarifier underflow is fed to a centrifuge for dewatering. The concentrated underflow is returned to the coal grinding and slurrying section and mixed with coal-feed slurry. The filtrate is returned to the clarifier.

This system processes 5000 TPD of ROM coal in each plant module. Each module consists of three operating and one spare complete trains.

4.1.3 System 3 - Initial Gas Cleanup and Cooling

Raw gas at 675 psig and 455°F enter this system from the coal gasification system. The gas is cooled to 380°F in a steam generator producing 110 psig steam. Further cooling to 330°F in a second steam generator produces 50 psig steam. Boiler feed water to the steam generators is heated to 300°F by heat exchange with the gas from the second generator thereby lowering the gas temperature to 297°F. Water condensate is collected in knockout drums following each steam generator and heat exchanger and recycled for quenching in the gasification system. The gas is next processed in a Texaco proprietary scrubbing unit for recovery of soot from the gas stream. The soot-water blowdown from this unit is recycled to System 1, slurry preparation.

Following the gas scrubbing unit the gas is cooled by air coolers to 140°F. Water condensate is recycled to the gasification quench system and the gas scrubbing unit. The gas is further cooled to 105°F by exchange with cooling water and contacted with the system makeup water in an ammonia scrubber. Water from the ammonia scrubber is used as makeup in the coal slurry preparation unit. Product gas at 100°F and 625 psig is delivered to System 4, Acid Gas Removal. This system is given in Drawing Number 524-TX-03.

4.1.4 System 7 ~ Raw Gas Compression

The Texaco gasifiers operate at sufficient pressure to meet plant delivery requirements without product compression.

4.1.5 System 4 - Acid Gas Removal

The acid gas removal system utilizes the Selexol solvent process and received 721,825 lbs/hr of sour gas at 100°F and 640 psia. The sweet (desulfurized) gas leaves the Selexol absorber at about 630 psia and a temperature of 75°F. The component removals across the absorber are:

	Inlet Gas (mols/hr)	Outlet Gas (mols/hr)	% Removal
Methane Hydrogen Carbon monoxide Carbon dioxide Hydrogen sulfide Carbonyl sulfide Nitrogen Water	154 11,038 15,254 5,467 454 29 375 48	154 11,032 15,251 2,938 1 5 375	0.00 0.02 0.02 46.26 99.78 86.21 0.00 91.67
	32,819	29,579	

The Selexol sulfur removal unit is a proprietary process licensed by Allied emical Company. The primary purpose of this system is to remove H₂S and COS from the product gas stream in order to obtain a sulfur concentration of not more than 200 ppmv. Secondarily, the process removes CO₂, which increases the gas heating value. System 4 is shown schematically in Drawing Number 524-TX-O4.

The Selexol process removes sulfur compounds and carbon dioxide by means of countercurrent contact in an absorber. The physical absorption of the compounds to be removed is done using the dimethyl ether of polyethylene glycol as a solvent. Chilled solvent is used. Solvent chilling is accomplished by means of an absorption refrigeration system utilizing low pressure gasifier jacket steam.

CO₂ removal from the rich solvent leaving the absorber is accomplished by a series of successively lower pressure flashes. Flashed gas is recycled to the absorber inlet. Sulfur compounds are removed in the stripping column. The stripping column utilizes a steam heated reboiler. The stripper overhead gas, after cooling to about 120°F, flows to the sulfur recovery and tail gas treatment unit, System 5.

The sour gas from the Selexol unit is rich in ${\rm CO}_2$, and at slightly above atmospheric pressure.

4.1.6 System 5 - Sulfur Recovery and Tail Gas Treatment

Acid gas from Acid Gas Removal, System 4, is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean H₂S acid gases. Typically, in a Claus-type sulfur plant, the acid gas is first passed through a knockout drum before entering the reaction furnace. The chemistry of the process involves converting H₂S to elemental sulfur according to the following equation:

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O$$

The reactions are exothermic, and the heat liberated generates steam in the reaction furnace boiler and in the sulfur condenser. The sulfur from each condenser is drained to a recovery pit in By-Product Processing, System 13, and the tail gas from the final condenser is fed to a Beavon tail gas treating unit where essentially complete removal of the remaining sulfur compounds is achieved before discharge to the atmosphere. The Beavon sulfur removal process reduces the sulfur content in the tail gas to less than 100 ppm. In this system, hydrogenation and hydrolysis are used to convert essentially all sulfur compounds to hydrogen sulfide. This gas is then cooled and passed into a contactor where the hydrogen sulfide is absorbed by the redox solution and oxidized to elemental sulfur. The reduced redox solution is reoxidized by contact with air and subsequently recirculated to

the contactor. Elemental sulfur is removed in the air-blowing step as a froth which is pumped to a sulfur melter to be melted under pressure, separated from the redox solution and transferred to By-Product Processing, System 13. The decanted redox solution is returned to the system. System 5 is shown in Drawing Number 524-TX-05.

The system receives 132,719 lbs/hr of acid gas from the Selexol solvent regenerator at about 7 psig and 120°F .

The inlet gas composition is:

	<pre>Inlet Gas (mois/hr)</pre>
Hydrogen	4.5
Carbon monoxide	3.5
Carbon dioxide	2 , 529.2
Hydrogen sulfide	453.1
Carbonyl sulfide	24.4
Nitrogen	0.0
Water	244.1
	3,058.8

The recovery of by-product elemental sulfur from the sulfur recovery system is 167 long tons/day, which amounts to an overall sulfur recovery of about 99.9 percent.

4.1.7 System 6 - Air Separation

The purpose of this unit is to supply oxygen to the gasifiers. The process is non-proprietary and is offered in package form by several designer/manufacturers in the U.S. The gaseous oxygen stream of 98% purity is produced by distillation of liquified air. System 6 is shown schematically in Drawing Number 524-TX-06.

In the unit, which consists of multiple trains, 1,628,000 lbs/hr of atmospheric air is filtered and compressed to 110 psia by multi-stage compressors. Drivers are steam turbines with motors installed for start-up. The filtered-compressed air is cocied via cooled heat exchangers, then passes through reversing heat exchangers where it is cooled to slightly above its liquefaction temperature by heat exchange with the oxidant and waste gas streams, and by subsequent expansion through a turbine.

Water and carbon dioxide freeze out in the reversing exchangers, and are therefore removed from the inlet air stream. Subsequent switching of the exchanger sublimes the water and carbon dioxide into vapor which is vented as waste.

The cold air then enters the high pressure column for initial purification. The vapor at the top of the column is nitrogen containing 10 ppm of oxygen. Most of this is condensed in the high pressure condenser/low pressure reboiler and returned as reflux to the high pressure column. Some of this nitrogen vapor becomes the feed stream to the expansion turbine and some becomes low pressure product nitrogen gas after being superheated in the superheater and being warmed to about -150°F in a reversing heat exchanger.

Some of the nitrogen condensed in the condenser/reboiler becomes the liquid nitrogen product. It is first subcooled in the waste nitrogen core subcooler and then flashed into the liquid nitrogen separator. Liquid nitrogen product from the separator is transferred to the liquid nitrogen storage.

An impure reflux stream containing a small percentage of oxygen is withdrawn at an intermediate point in the column. It is subcooled and flashed into the top of the low pressure column.

The final purification of oxygen takes place in the low pressure column. The feeds to this column are impure reflux, and crude liquid oxygen from the high pressure column is subcooled and passed through the hydrocarbon absorbers before being flashed into the low pressure column. This procedures assures that all hydrocarbons are removed from the oxygen before purification. Processing the crude oxygen after it is liquified assures that the absorption is done at a relatively constant pressure and temperature and thus limits the possibility of accidental desorption.

As a further safeguard, a portion of the liquid oxygen in the low pressure sump is constantly withdrawn, passed through a guard hydrocarbon absorber and returned to the sump. This prevents trace quantities of hydrocarbons from accumulating in the low pressure sump.

The separated oxygen from the cold box (at 2 psig and 70° F) is compressed, in six stages of compression, to 815 psia for use in the gasifiers.

Modules one and three have three trains each while modules two and four have two trains each of air separation.

4.1.8 System 8 - Process Solids Treatment

Approximately 69,003 pounds per hour per module of ash and slag from the gasifier and gas cooling system along with biological sludges and solid wastes from process condensate treatment are treated in this unit.

Gravity settlers separate dense solids from the waste streams.

Float thickeners-clarifiers treat slurry from the gravity settlers. Thickener and coagulant aids are added to facilitate solid-liquid separation.

Rotary drum filters filter sludges from the process condensate treating systems and the float thickeners.

Recovered water is sent to the process condensate treating system, and solids are conveyed to final solids disposal.

4.1.9 System 13 - By-Product Processing

Sulfur is the only plant by-product other than ash/char solids which are disposed of on-site. Mo!ten sulfur is pumped to a prilling tower in a continuously flowing circulation system. In the tower, sulfur is dispensed in droplets through nozzles. Droplets fall counter current to a stream of cooling air and solidify prior to landing in the bottom prill collection section. From the prill tower, sulfur is conveyed to a storage building from which it is transferred by truck or barge for sale. System 13 is illustrated in Drawing Number 524-TX-13.

4.1.10 System 14 - Plant Power System

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

- (1) Develop the necessary voltage stepdown arrangement for plant requirements
- (2) Distribute the necessary power to the plant equipment.

TVA's incoming substation transformers receive power from its primary distributed voltage switching station and step down this voltage to a medium voltage to supply the plant electrical power requirement for motors, heaters, lighting, and other miscellaneous loads.

The Medium Voltage Electrical Distribution Systems is a secondary selection system (double ended supply) with several medium voltage buses. Each medium voltage bus receives power from its respective incoming substation transformer through an incoming breaker and supplies power to the medium voltage distribution system through the feeder breakers.

The Low Voltage Electrical Distribution System typically consists of multiple 480 V double-ended load centers and 480 V motor control centers (MCC's) supplying the power to 480 V loads throughout the plant. Two load centers are interconnected through a normally open tie breaker. In the event of loss of one load center transformer or its feeder, the 480 V loads of the affected load center are fed by the second load center through the tie breaker.

Each load center consists of an incoming line section, load center transformer, and low voltage section with metal enclosed draw out power circuit breakers.

Load center transformers are air cooled, dry type, 150°F temperature rise, with delta connected primaries and wye connected secondaries. All load center feeder circuit breakers are 1600A frame and 50,000A RMS symmetrical interrupting capacity. The 480 V motor feeder breakers are electrically operated with instantaneous and long time trip units.

480 V MCC's consist of starters, feeder circuit breakers and control devices, assembled in a common structure with horizontal and vertical buses.

A 125 volt DC system supplies control power for medium voltage and 480 volt plant switchgear control, protective relaying and annunciation. The system also supplies power for emergency lighting.

4.1.11 System 15 - Steam Generation/Distribution

In the absence of a waste heat boiler on the gasifier outlet there is no high pressure steam generated. Gas cooling and the sulfur plant provide 913,000 lb/hr of 125 psia saturated steam used as follows:

Selexol regenerator - 240,000 lb/hr Electric power generation - 661,000 lb/hr Turbine condensate reheat - 12,000 lb/hr

Gas cooling and the sulfur plant produce 173,000 lb/hr of 65 psia saturated steam used as follows:

Selexol unit refrigeration - 144,000 lb/hr Sulfur heating - 2,000 lbs/hr Steam tracing - 2,000 lbs/hr BFW deaeration - 25,000 lbs/hr

Steam condensate is recovered at 30 psia and 250° F. Total BFW circulation rate is 1,086,500 lbs/hr.

There are no fired steam boilers contained in the design. System 15 is shown in Drawing Number 524-TX-10.

4.1.12 System 16 - Water Supply

The raw water treatment unit is designed to provide treated and untreated water for the following facility water systems:

- (1) Fire water
- (2) Service water
- (3) Potable water

(4) Cooling water

(5) Boiler feed water

Raw water is pumped from the river to a fire water-raw water storage tank.

The raw water-fire water storage tank provides surge capacity for water treatment as well as storage capacity for fire water. During an emergency, fire water is pumped from the tank to the fire water heater system. The fire water pumps are motor driven and have a diesel engine driven spare. The spare pump is equipped with automatic start-up capability in case of power failure.

The raw water is pumped from the raw water-fire water storage tank to the softener-clarifier. Lime, alum, and polyelectrolyte from the clarifier bulk chemical storage and feed system are added to the softener-clarifier, which is equipped with an internal flocculation mechanism. The alum and polyelectrolyte aid in the removal of suspended solids from the raw water. Lime is added during the clarification step to "cold soften" the raw water. Chlorine is added to the raw water to inhibit algae growth in the clarifier and sand filters and reduce organic contamination.

The underflow from the clarifier is a one wt percent sludge and is pumped to solids treatment for further processing.

The clarified and softened raw water from the softener-clarifier flows to the self-backwashing sand filters where additional suspended solids are removed. A pressure differential across the filter bed initiates the backwash cycle. The backwash flows by gravity to the sand filter backwash sump and is recycled to the softener-clarifier. The filtered water flows to the filtered water storage tank and is utilized as cooling tower make-up for the process cooling tower as service water for general plant use, as feed to the demineralizer package, and as feed to the potable water system.

Water intended for potable services is chlorinated and again filtered to meet American Water Works Association (AWWA) standards and stored in a tank sized to hold a day's potable water requirements. The chlorine residual is maintained at 0.5-1.0 ppm free chlorine in the tank.

Filtered water intended as feed to the demineralizer package is injected with sodium sulfide to remove trace amounts of chlorine which adversely affect the demineralizer resins and after fil. red through activated carbon to remove any remaining organic contaminants and dissolve iron.

In the demineralizer, the mineral salts present in the water are removed by ion exchange. A two-step demineralization system, utilizing strong cation and strong anion exchangers in series, is provided. A degasifier following the strong cation and magnesium, which the anion exchangers remove anions such as chloride and sulfate. The strong anion exchanger also removes

silica. The degasifier is provided to remove carbon dioxide and other dissolved gases.

The mixed bed polisher is provided to remove silica to 0.02 ppm and to polish returned turbine condensate for reuse.

The boiler feed water deaerating heaters operate at 30 psig and 250° F. The deaerators reduce the oxygen content of BFW to 0.005 cc/liter.

Hydrazine or sodium sulfite is injected into the storage compartment of the deaerators for chemical scavenging of any residual oxygen. Morpholine is injected into the suction of the boiler feed water pumps to protect the condensate systems.

The total water requirement is illustrated in Figure 4.2. Allowances for water treatment blowdown and contingency add to that shown to result in a total requirement of 2500 GPM per module.

Drawing Number 524-TX-16 shows System 16.

4.1.13 System 17 - Water Cooling

The purpose of this unit is to provide cooling water to the various process users in the facility.

The cooling tower system includes the tower and fans, side stream filters, circulating water pumps, cold water basin, blowdown system, chemical addition equipment, and distribution system.

Cooling water is pumped from the cold water basin, through the distribution system to the process heat exchangers where low-level, sensible heat is picked up, and back to the cooling tower. The cooling tower rejects low-level heat by evaporative cooling to air drawn through the cooling tower by the cooling tower fans.

A portion of the circulating water is passed through side stream filters to reduce loading to suspended solids, dirt and scale.

The dissolved solids level of the cooling water is maintained by a continuous blowdown stream to the process condensate system. Water level in the cooling tower basis is maintained by continuous make-up of the clean water from the raw water treatment system.

The blowdown stream is passed through a blowdown treatment system to recover chromate ions via ion exchange or by chemical reduction to chromium hydroxide and is sent to waste treatment for disposal.

Chlorine is added to the cooling water on a routine periodic basis to prevent algae growth. Chemical algicides are added periodically to further

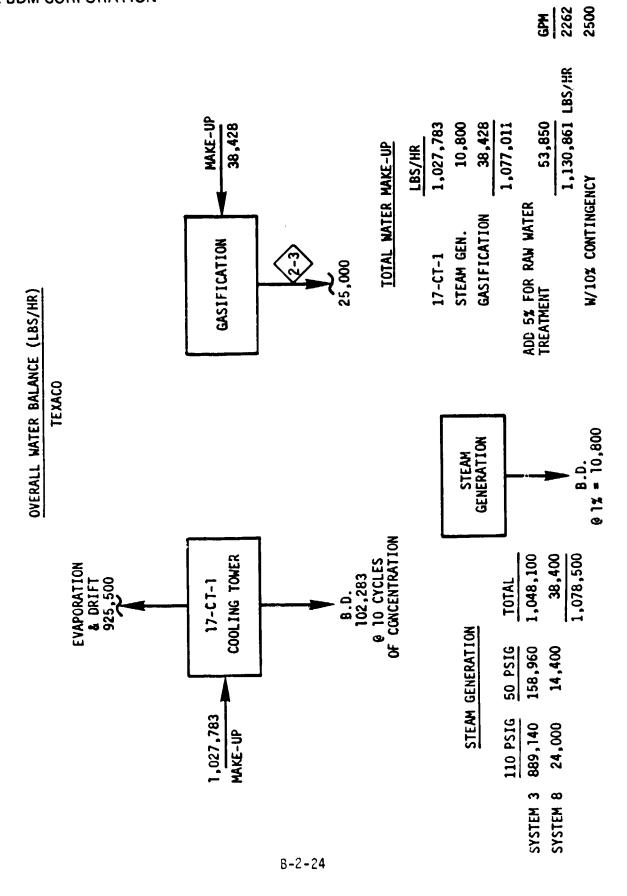


Figure 4.2. Water Balance

eliminate algae growth. Sulfuric acid is added to control pH, and zinc and chromate inhibitors are added to the cooling water for corrosion control. Occasionally, a polyphosphate dispersant is added to enhance the action of the inhibitors. This system is illustrated in Drawing Number 524-TX-12.

4.1.14 System 18 - Waste Water Treatment

The purpose of this unit is to collect and treat all plant liquid effluent streams. The plant design is predicated on "zero discharge" and permits recycle and reuse of treated water. Streams treated include the following:

- (1) Oily water sewers
- (2) Coal pile run-off
- (3) Storm water run-off(4) Demineralizer regenerant wastes and rinse water

- (5) Cooling tower blowdown
 (6) Sanitary waste water
 (7) Gasifier slag quench drains
- Separated water from solids treatment (8)
- (9) Filtrate from biological treatment.

Process operations include:

- (1) Oil Separator streams containing free and dissolved oil and treated in a gravity separator utilizing an emulsion breaking chemical and heat to separate the oil-water mixture
- (2) Sour Water Stripper water streams with appreciable H₂O and HN₂ residuals are steam-stripped to remove these contaminants
- (3) Equalization Basin liquid streams with extremely high or low pH are mixed in an equalizing basis and treated with sulfuric acid or caustic to change the mixed pH to a value of 6.0 - 8.0
- (4) Gravity Settling-Thickener liquid streams with high suspended or dissolved solids are treated in a gravity settler-thickener and mixed with lime, alum, coagulant aids, and polymers to facilitate separation and thickening
- (5) Multiple Effect Evaporation neutralized wastes and brines are evaporated to recover water and concentrate the solids.

The recovered, treated water is used as make-up to cooling towers or raw water supply. The resultant solids are conveyed to the solids disposal system. System 18 is illustrated in Drawing Number 524-TX-18.

4.1.15 System 19 - General Facilities

The purpose of this unit is to provide equipment or services to support the gasification facility at the facility level.

This unit is a general facility category and provides the following equipment and services:

- (1) Administration building

- (1) Administration buildin
 (2) Laboratories
 (3) Change rooms
 (4) Warehouses
 (5) Maintenance buildings
 (6) Operation centers
 (7) Security offices
 (8) Plant air facility
 (9) Fire house

- (10) Visitor reception
- (11) Plant fencing

- (11) Plant fencing
 (12) Plant lighting
 (13) Roads, bridges
 (14) Docking facilities
 (15) Interconnection pipe ways
 (16) Fire protection network
 (17) Flare stacks and headers
 (18) Plant instrument air compressors
 (19) Environmental monitoring
 (20) Site preparation.

4.2 Module Balances

Material and energy balances have been determined for each module. Table 4.1 gives stream compositions and quantities for each intersystem stream. Table 4.2 gives an energy balance at the modular level.

TABLE 4.1
MATERIAL BALANCE
TEXACO PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER		<u> </u>	0-1-0		0-2		1-1	
STREAM ID		•	COAL PIED	ď	EAN VATE	ATER	COAL SLIEBY TO CASIFICATION	1 T
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LBS	LB-MOLS HR	LBS	LB-MOLS LBS	
CARBON	Ü	12.01		253.637		_	٦,	
HYDROGEN	LB2	2.016		17.925			23 663	_
NITROGEN & ARGON	72	28.016		5.761			5,761	T-
	S.2	32.06		15,449			15,449	П
CHLORINE	CI	35.453		767		_	464	_
CARBON MONOXIDE	ဗ	28.01					1	
CARBON DIOXIDE	03	44.01						_
METHANE	CB	16.042		•		-		_
ETHYLENE	Call	28.052		-		-	1	7
FTHANE		30.068						_
PROPYLENE	E,F	42.078		-		•		_
PROPANE	C He	44.094						7
HYDROGEN SULFIDE	H.S.	34.076				-		7
CARBONYL SULFIDE	sgo	60.075		-		-	1	_
CARBON DISULFIDE	CS	76.13		,		-		-т
SULFUR DIOXIDE	SO	90.19		1		-		7
HITROUS OXIDE	NO	30.008		•		-	•	-
AMMONIA	NH,	17,031		•				
	HCM	27.026		-		#	1	_
HYDROGEN CHLORIDE	HC1	36.461		-		-		_
NAPHTHA	1							_
TAR & OIL	-			-		-		7
АЅИ	1			29,650		-	59,650	_
OTHER SOLI							(v)	_
SUBTOTAL, DRY				376,817		•	376,817	7
WATER	Изо	18,016		39,853			39,853	_
TOTAL, WET	•		;	416,670	•		416,670	
GAS MOLECULAR WEIGHT	GHT			-		-		\neg
TEMPERATURE, OF				•		-	140	_
PRESSURE, PSIA				1		-	-	
SCFH				•		ı		-
GPM				1		2,533		_

MATERIAL BALANCE TEXACO PROCESS INTEGRATED FACILITY/MODULE DEFINITION

		•	<				<	
			2-1	^	71		\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\	
STREAM ID		•	BAN GAS TO C	COOLING	CASIFIER TO SOLIDS TREATHERIT	TREATHERIT	PROCESS CONDENSATE MICHIGAN	DOLLER
	SYMBOL	FORMULA WEIGHT		LBS	LB-MOLS HR	LBS	LB-MOLS LBS	
CARBON	U	12.91		t				
HYDROGEN	Ä	2.016	11.037.5	22.252				
OXXCEN	20	32.0						
HITROGEN & ARGON	N2	28.016	375.0	11,153		-		
SULFUR	5.	32.06		•				
CHLORINE	ច	35,453	i.			1		
CARBON MONOXIDE	00	28.01	15,254,2	427.220		1	•	
CARRON DIOXIDE	co	44.01	5,466.7	240.589		t		
MITHANE	CB	16.042	154.2	2,474		•	•	
ETHYLENE	. E-3	26.052	•				•	
F.THANE		30.068	-			•		
PROPYLENE	C B	42.078	_			-		
	C'H'S	44.094	1	B ₁		-	•	
HYDROGEN SULPIDE	H Se	34.076	454.2	15.677		•		
CARBONYL SULFIDE	cos	60.075	29.2	1,754		-	-	
CARBON DISULFIDE	CS,	76.13	1	1		-		
SULFUR DIOXIDE	505	64.06					•	
HITROUS OXIDE	NO	30.008				-		
AMMONIA	IN	17.031	104.2	1,775		•	1	
HYDROGEN CYANIDE	HCM	27.026	1	,		_		
HITOROGEN CHLORIDE	HC1	36.461				1		
HAPHTIIA			-			•		
TAR & OIL	,		1		•	1	•	
ASH	-					59. 650	1	
OTHER SOLIES			-	-		2,577	•	
SUSTOTAL, DRY			32.875.2	722,744		62,227	•	
WATER	H ₂ 0	18.016	66,543.9	1.198.855		*	25,000	98
TOTAL, WET	' 		99,419.1	1,921,599		62,227	25.000	9
-	HT		19.33					
			455			1		
PPESSURE, PSIA			069			-		
SCFH			37,729,548		Alesantielly no			
GPM	•	- ',	-	-:	free water			3

MATERIAL BALANCE TEXACO PROCESS INTEGRATED FACILITY/MODULE DEFINITION

			_	1-1	_	Ÿ	√
STREAM NUMBER					Linearia	HOG INTERNAL	. CONSIDERTION
STREAM ID		COOLED CAS TO ACID CAS REMOVAL	CAS REPOVAL	LB-MOLS	LBS	LB-MOLS	LBS
SYMBOL	FORMULA WEIGHT	HR	XX	HR	PAR		
U	12.01	2 10 11	22.252	10.965.8	22,107.5	67.2	135.5
HYPROGEN B2	33.0	71,77	'		2 300 11	2.28	67.8
- 1	28.016	375.0	11,153	372.72	11,082.9	8317	
	32.06						
2	35,453		22, 23,	15 157 8	A2A_567.0	32.9	1,502,1
ONOXIDE	28.01	15.255.2	277.77	2 919 6	128.491.6	11.9	181.8
	44.01	2,400.1	777	153.26	2.458.6	0.94	131
	16.042	134.2			1		
	28.052						
	30.068			1			
L PRODYLENE	42.078			1	•		
	44.094		15. 477	1.464	39.0	900.0	40Z-0
HYDROGEN SULFIDE B'S	34.076	3.36	187	4.77	286.6	P.0.0	1 142
- 1	60.075	7.77					
SULFIDE	76,13				t .		
	64.06			1			
	30.008			1	•	•	
AMMONIA	17.031				•	,	
N CYANIDE	27.026			•	•	•	
HYDROGEN CHLORIDE HCL	36.461					1	
					•	-	
· oil							
					•		275 017
OTHER SOLIDS		22, 221, 0	720.969	29,575,095	589,035.3	181.255	3.010.690
SUBTOTAL, DRY		3677775	8.8	4.36	78.5	6.0	7,791
WATER H20	18.016	(:)*	300 .00	20 579 655	589,113.6	181.322	3,611.453
1.43		32,618.5	171,623	1	10 07	l	19.92
GAS MOLECULAR WEIGHT		21.99	9	3	, , , , , , , , , , , , , , , , , , ,	37	
TEMPERATURE, OF		001		613	S	9	513
PRESSURE, PSIA		12 424 67	1 (42)	11,22	11,225,403	2	710100
		72.131					

MATERIAL BALANCE TEXACO PROCESS INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER			***	(1)	\(\frac{1}{2}\)			[]
STREAM ID			ACID GAS TO SILFUR RECOVERY	LEUR RECOVERY	TAIL GAS FROM SIGRUR SECONERY	LFUR SECOVERY	OXYGEN TO	OXYGEN TO CASIFICATION
	SYMBOL	FORMULA WEIGHT	LB-MOLS HR	LBS	LB-MOLS HR	LBS	LB-MOLS HR	HR
CARBON	U	12.01		1	ŝ	9	•	t
HYDROGEN	H	2.016	4.5	1.6	25.8	52		•
OXYGEN	ja	32.0					10,906	349,000
RITHOGEN & ARGON	N	28.016	-		1,000.1	28,036	220.2	6,825
SULFUR	S	32.06						
CHLORINE	23	35.453	-					
CARBON MONOXIDE	05	28.01		98.0			:	
CARRON DIOXIDE	60	14.01	77877	110.11	7,387.4	113,871		
MILLIANE.	J.	16.042						
CONTRACTO	1-127	20.05				•		
o product eve	22.0	2000					1	•
,	1 2 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1	44.094	-					
HYDROGEN SULFIDE	H S	34.076	453.05	15.438.1	0.00279	0.095	1	
CARBONYL SULFIDE	cos	60.075	24.4	1,46	1	1		
CARBON DISULFIDE	CS	76.13	1					
SULFUR DIOXIDE	505	64.06	-		1			•
NITROUS OXIDE	NO	30.008	+	1				•
	NH.	17.031	-					•
	HCM	27.026						
HYDROGEN CHLORIDE	HC1	36.461	-					
GAFHTIA								
ASH ASH	1		-	_		•	-	
OTHER SOLING				-		_	1	
SUBTOTAL, DRY			3,014.65	128,321.1	3,613.9	141.959	921.11	35.83
WATER	н,0	18.016	244.1	4,397.7	167.9	3.025		
TOTAL, WET			3,258,75	132,718.8	3,781.8	14/, 984	11,126	355,833
GAS MOLECULAR WEIGHT	iliT		67.04		38.36		7	31 98
TEMPERATURE, OF			120		001		7	CO.
PRESSURE, PSIA			n		4.41		1	238 146
SCFH			1,236,696	96	1.415.191	183	77,1	Verio 277
€PM			-	The second of the second		L		

MATERIAL BALANCE TEXACO PROCESS INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER		- (P-1)	(1-1;)	(1-(1)
STREAM ID		PROCESS SOLIDS TO DISPOSAL	COM. TO COM. PREPARATION	13 13000
SYMBOL	BOL FORMULA WEIGHT	LB-MOLS LBS HR HR	LB-HOLS LBS HR	LB-MOLS LBS
nonato		4	253,637	
EN	2.016		17,925	
	12.0		23,901	
MITROGEN 6 ARGON N2	28.016	•	5,761	301.31
	32.06		767	
	35.453			
CARRON MONO : E				
S KOIC				
D ETHYLENE				
	9	1		
PROPERTY OF THE CAME			•	
NO BOLEM CHESTOR			3	
SULFIDE				
1				1
SULFUR DIOXIDE SO		1	1	
		5		
N CYANIDE				• • •
HYDROGEN CHIGRIDE HCI				
NAPHTHA				
6 OIL		059 65	65,65	
ASE COLUMN		3, 203		
SCHOOL DEV		68.853	376,817	477.4 15,305
WATER	18.016	150	39,853	
WET		69,003	416,670	477.4 15,305
GAS MOLECULAR WEIGHT				
TEMPERATURE OF				
PERSONE PSIA		*		
SOFI				
T (1)			•	

MATERIAL BALANCE TEXACO PROCESS I''GGRATED FACILITY/HODULE DEFINITION

STREAM NUMBER			\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\	(16-2)	
STREAM ID			10 COOL 18	NAKE-UP MATER TO STEAM CENERATORS	
	SYMBOL	FORMULA WEIGHT	LB-MULS LDDS	11	HR
САКЬОИ	ú	12.01			
HYDROGEN	В	2.016			
	p	32.0			
RITROGEN & ARGON	, K	28.016			
CHLORINE	20	35.452			
CAPRON MONOXIDE	ပ္ပ	28.01			
CARRON DIOXIDE	ر02	14.01			
METHANE	, io	16.042			
ETHYLENE	C,D	28.052			
ETHANE	<u>C_2</u> !! 6	30.06			
C CROPYLEAN	35.6	870.75			
HYDROGEN SULFIDE	33.5	34.076			
CAPBONYL SULFIDE	cos	60.075			
CARBON DISULFIDE	CS	76,13			
SULFUR DIOXIDE	\$0.5	64.06			
RITHOUS OXIDE	NO.	30.008			
AMMORIA	NIN N	17.031			
ETCHOREN CYANIDE	HCM	27,026			
HYDROCEN CHLORIDE	HC1	36.361			
ZAPHTHA					
770 • 200	֡֡֜֝֝֡֜֜֜֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֓֡֓֜֓֡֓֡֓֜֓֓֡֓֜֜֡֓֓֡֓֜֡֓֓֜֡֓֜֡				
OTHER SOLIDS					
SUBTOTAL. DRY					
WATER	N,O	18.016			
TOTAL, WET	4				
GAS MOLECULAR WEIGHT	H		1		
TEMPERATURE, OF			1		
PERSSURE, PSIA			4		
				22	
X.76			Ser.	77	

TABLE 4.2
ENERGY BALANCE, 10⁶ BTU PER HOUR
TEXAGO PROCESS
INTEGRATED FACILITY/MODILE DEFINITION

	TOTAL	4575 14 0	4589		3262.7 38.4 4.7 57.7	3381.2	1207.8	168.9 0 -1301.1 0 -42.4 -1174.6 1174.6
INTEGRATED FACILITY/MODULE DEFINITION	SENSIBLE	••			9. 6 5. 0 9. 6			-551.0
TED FACILITY/M	LATENT	2			3.2			-750.1
INTEGRA	HIIV	4575			3262.7 30.2 57.7			
	PHASE	Solid Cas Liquid			Gas Solid Gas Solid Gas			
	ENTHALPY BALANCE INLETS	STREAM COAL AIR SEP'N RAW MAKET	SUBFOTAL IN	OUTLETS	STREAM MBG Ash Tail Gas Sulfur Air Sup'n Vent Flue Gases	SUBTOTAL OUT	NET INCET-OUTLET	HEAT AND SHAFT WORK ELECTRIC POWER SHAFT PORK HEAT REJECTION HEAT INPUT (STEAM) KADIATION/FUGITIVE LOSS SUBTOTAL NET VALUE TOTAL ABSOLUTE VALUE

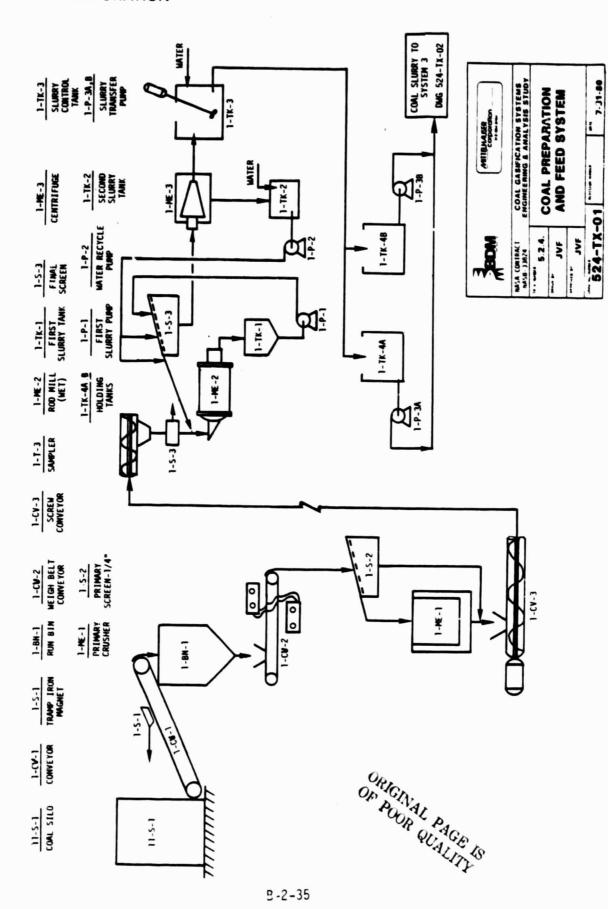
BALANCE: NET ENTHALPHY + NET HEAT & SHAFT WORK X 100X = 2.8X TOTAL ABSOLUTE VALUE OF HEAT & SHAFT WORK

BASIS: 60°F; LIQUID WATER

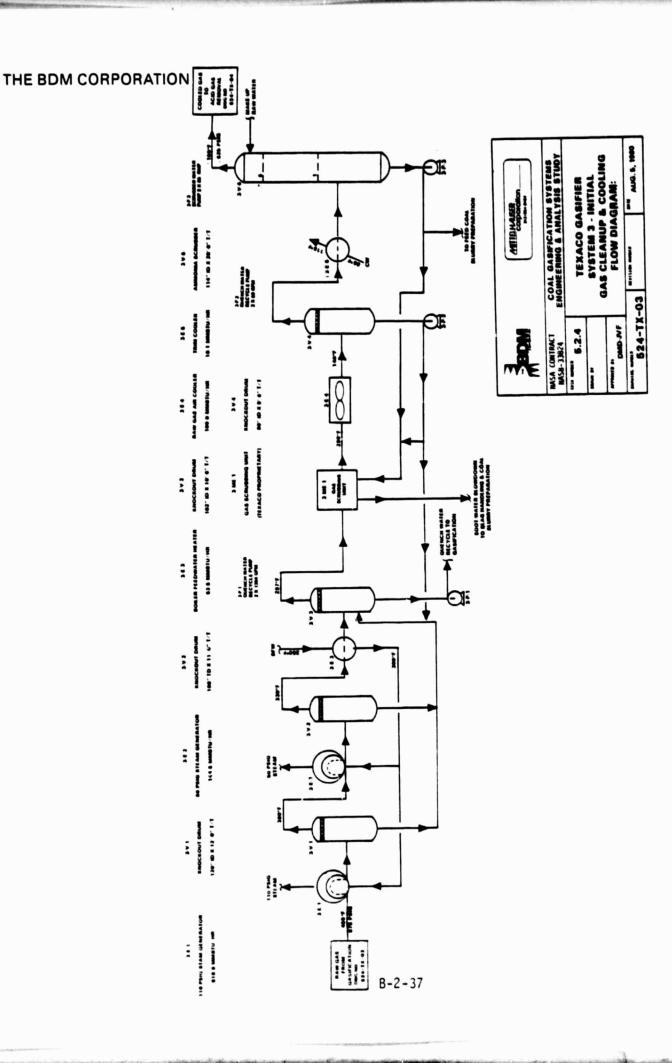
5.0 PROCESS FLOW DIAGRAM AND EQUIPMENT LIST

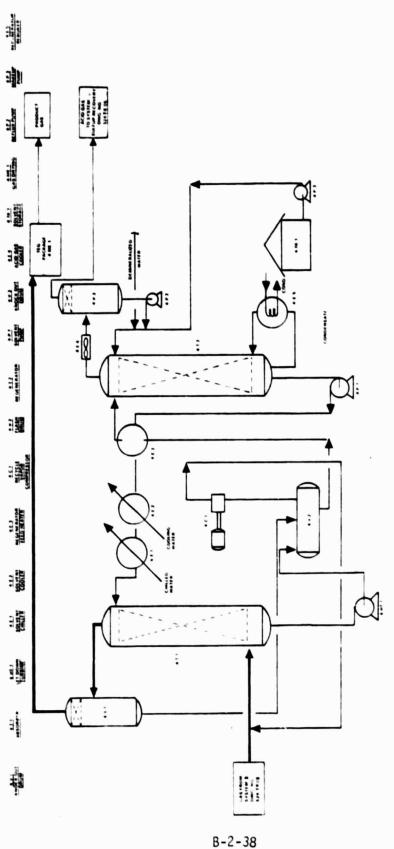
This section contains the process flow diagrams and the list of equipment for each module. The facility contains four times the quantity shown here except for the coal silos which are on a facility basis and System 5, which has two trains in Module 1 and only one train in subsequent modules.

1

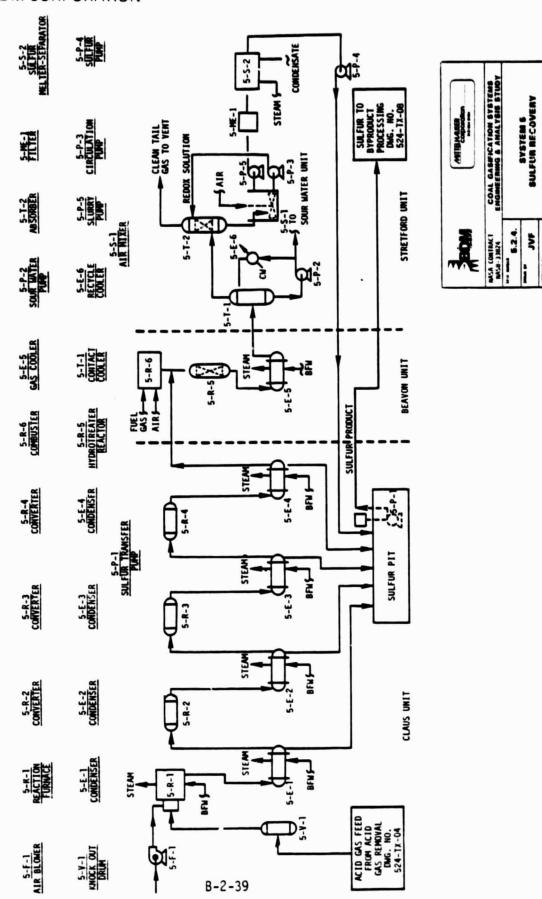


B-2-36

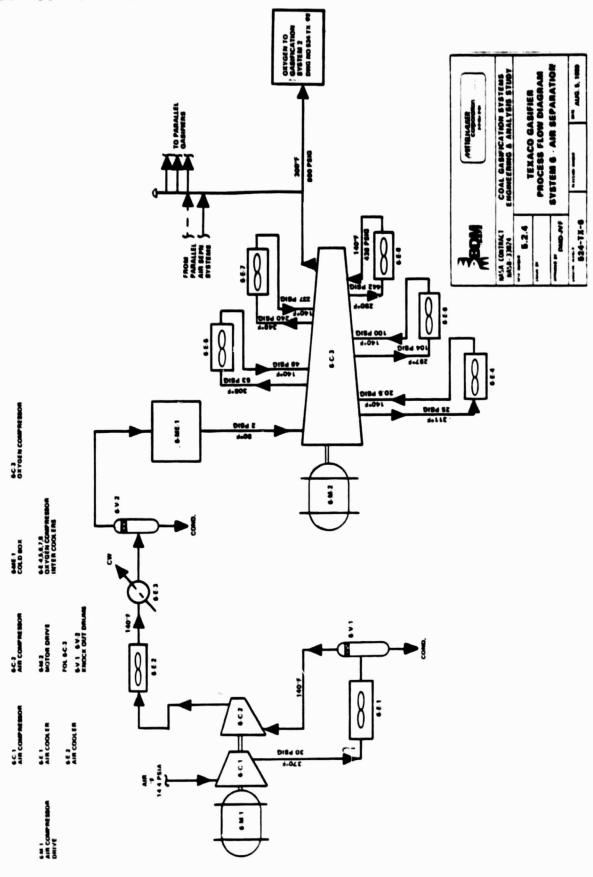


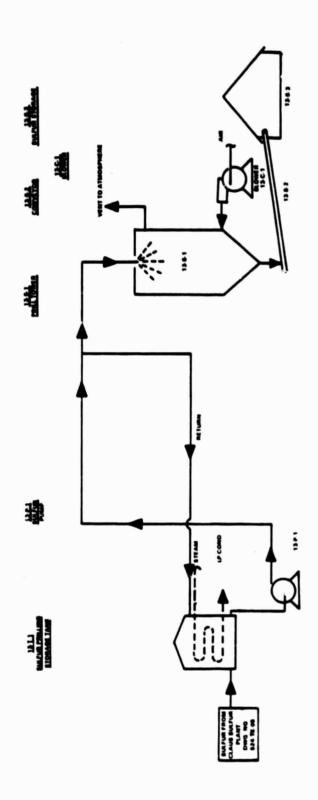


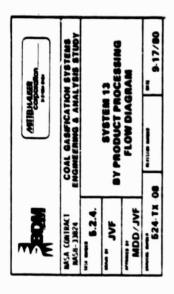
MITEHALER

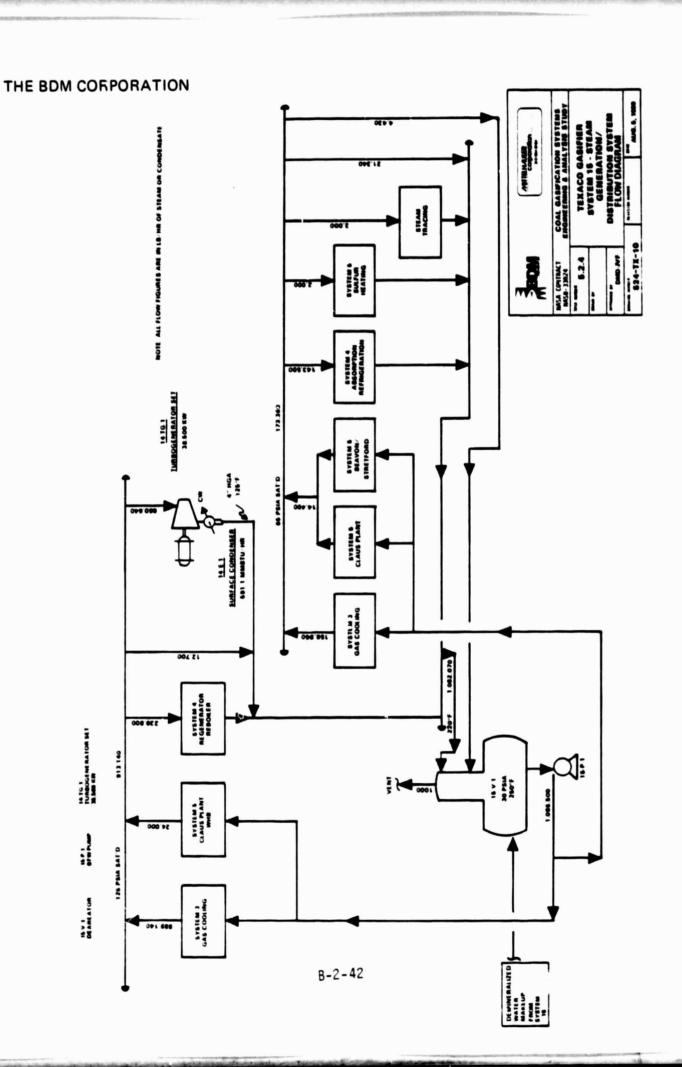


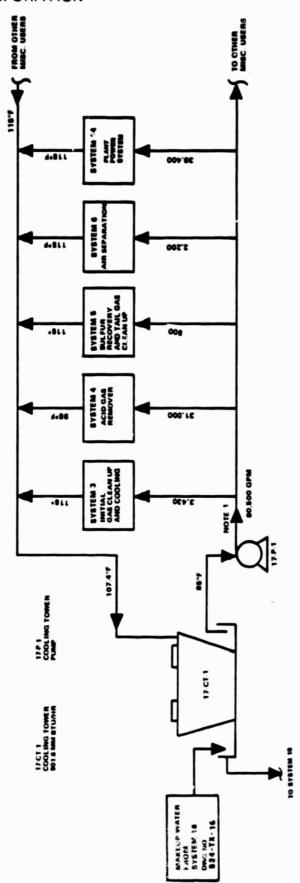
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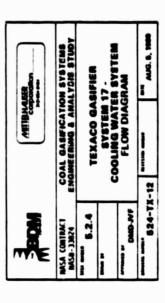


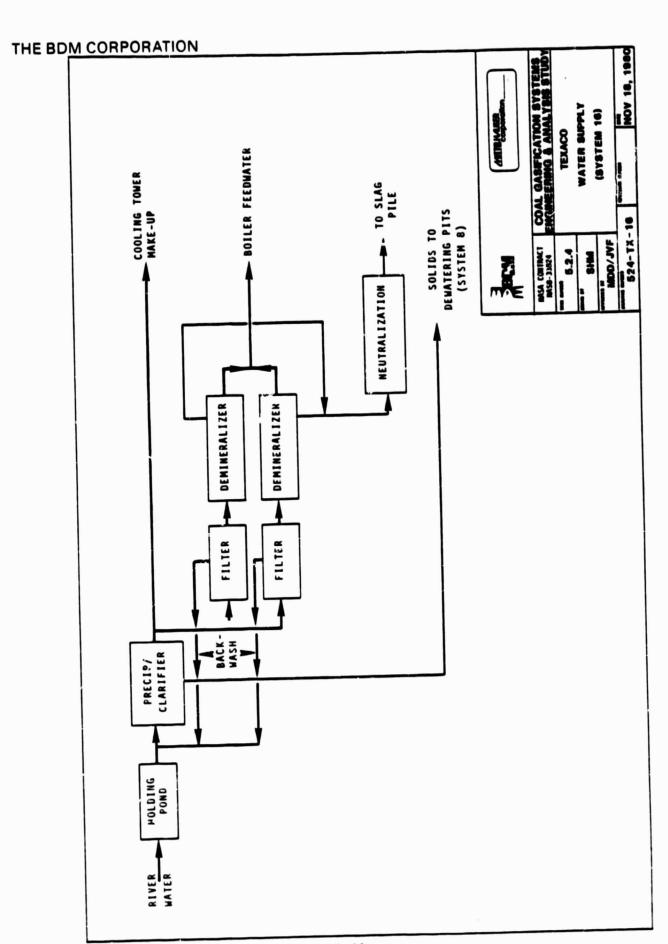


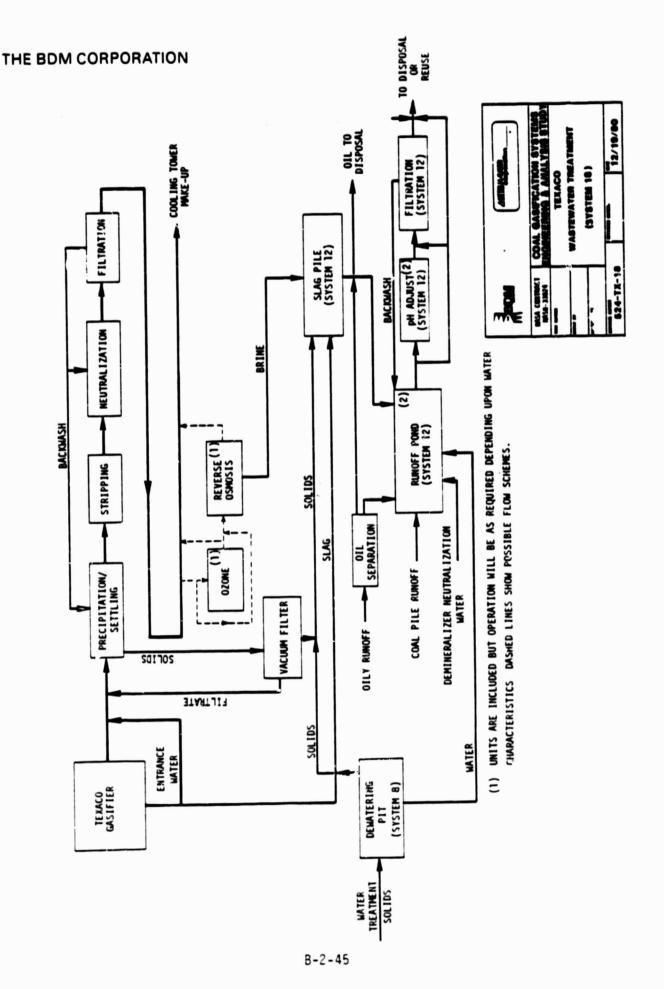


B-2-43

HOTE 1: TOTAL 80,500 GPM SICLUDES 6% CONTRIGENCY ALLOWANCE







N EQUIPMENT LIST
TVA MBG MODULE
TEXACO GASIFIER
SYSTEM NO: 1 - COAL PREPARATION & FEEDING

ITEM	SERVICE	QUA	NTITY	000000000000000000000000000000000000000
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
11-5-1	COAL STORAGE			COAL SILO
1-5-1	TRAMP IRON REMOVAL	NAPP .		MAGNETIC
1-CV-1	COAL TRANSPORT TO SYSTEM 1			COVERED CONVEYOR
1-BN-1	FEED SURGE	1	1	700 TON COAL BIN
1-CV-2	CONTROL FEED RATE	1	1	WEIGH BELT CONVEYOR
1-ME-1	CRUSH COAL TO	1	1	CAGE MILL
1-5-2	SEPARATE COAL AT 1/4"	1	1	VIBRATING SCREEN
1-CV-3	TRANSPORT COAL TO SLURRY PREPARATION	1	1	SCREW CONVEYOR
1-5-3	SAMPLE COAL	1	1	

EQUIPMENT LIST
TVA MBG MODULE
TEXACO GASTETER

TEXACO GASIFIER
SYSTEM NO: 1 - COAL PREPARATION & FEEDING

ITEM	SERVICE	QUA	YTITM	DECEDITATION
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
1-ME-2	FINE GRIND COAL	1	1	ROD MILL (WET)
	4			,
1-P-1	PUMP SLURRY TO SCREENING OPERATION			CENTRIFUGAL SLURRY PUMP
1-P-2	CIRCULATE WATER FOR SCREENING OPERATION			
1-TK-1	ROD MILL EFFLUENT TANK	1	1	CONICAL BOTTOM SLUR TANK
1-S-3	FINAL PRODUCT SIZE CONTROL	1	1	WET SCREEN
1-TK-4 A&B	HOLDING TANKS		1	STIRRED FEED SLURRY TANKS
1-ME-3	SLURRY WATER REDUCTION	1	1	CENTRIFUGE
1-TK-3	SLURRY COMPOSITION CONTROL	1	1	SLURRY TANK WITH DENSITY MONITORI FOR MAKEUP WATER CONTROL
1-TK-2	COLLECT CENTRIFUGE EFFLUENT	1	1	DATA SURGE TANK FOR SCREENING OPERAT

EQUIPMENT LIST TVA ::BG MODULE TEXACO GASIFIER

SYSTEM NO: 1 - COAL PREPARATION & FEEDING

ITEM	SERVICE	QUA	ANTITY	05000107104
NUMBER	SERVICE	SPARE	DPERATION	DESCRIPTION
1-P-3 A&B	TRANSPORT SLURRY TO SYSTEM 2		1	CENTRIFUGE SLURRY TRANSFER PUMPS

EQUIPMENT LIST TVA MBG MODULE TEXACO GASIFIER SYSTEM NO: 2 - COAL GASIFICATION

ITEM	SERVICE	QUA	NTITY	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
2-V-1	COAL SLURRY RUN TANK	1	3	ENCLOSED STIRRED TANK WITH PUMP AROUND SYSTEM
2-P-1	PUMP AROUND AND POSITIVE SUCTION SLURRY PUMP			
2-P-2	GASIFIER FEED PUMP	1	3	POSITIVE DISPLACEMEN HIGH PRESSURE SLURRY PUMP
2-R-1	COAL GASIFICATION	. 1	3	TEXACO REALTOR SYSTE
2-V-2	SLAG REMOVAL	1	3	TEXACO SLAG REMOVAL SYSTEM

EQUIPMENT LIST TVA MBG MOUGHE THYACO CASIFIER

SYSTEM NO: 3 - INITIAL GAS CLEANUP & COOLING

ITEM	SERVICE	QUAN	ITITY	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
3-2-1	110 PSIG STEAM GENERATOR	0	5	818.9 MM Btu/hr TOTAL KETTLE TYPE, 15,000 SQ FT/UNIT SHELL: CS; TUBES: SS
3-E-2	50 PSIG STEAM GENERATOR	0	2	144.5 MM Btu/hr TOTAL RETTLE TYPE, 10.080 SQ FT/UNIT SHELL: CS; TUBES: SS
3-E-3	SOILER FEEDWATER HEATER	0	1	53.5 MM Btu/hr TYPE AEL 17,650 SQ FT SHELL: CS; TUBES: SS
3-1-4	RAW GAS AIR COOLER	0	1	100.0 MMBtu/hr AIR COOLER FOUR BAYS, EA. 28'x30 6 TUBE ROWS, SS TUBES
3-E-5	TRIM COOLER	0	1	10.0 MM Btu/hr TYPE AEL 5650 SQ FT SHELL: CS; TUBES: SS
3-V-1	KNOCKOUT DRUM	0	1	VERTICAL VESSEL 120" ID x 12'-0" T/T C.S. DT: 450°F; DP: 700 P
3-V-2	KNOCKOUT DRUM	0	ì	VERTICAL VESSEL 108" ID x 11'-0" T/T C.S. DT: 380°F; DP: 700 P
3-7:-3	FEROCKOUT DRUM	0	1	VERTICAL VESSEL 102" ID x 10'-6" T/T

EQUIPMENT LIST Tha Meg Midule

TETACO GASIFIER

SYSTEM NO: 3 - INITIAL GAS CLEANUP & COOLING

ITCH	T	OUA	NTITY	T
ITEM NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
3-V-4	KNOCKOUT DRUM	0	1	VERTICAL VESSEL 90" ID x 9'-6" T/T C.S. DT: 200°F; DP: 700 PSI
3-V-5	AMMONIA SCRUEBER	0	1	VERTICAL VESSEL 114" ID x 20'-0" T/T C.S. w/6 SS SIEVE TRAY DT: 150°F; DP: 700 PSI
3-P-1	HOT FROCESS CONDENSATE FUMP	1	2_	CENTRIFUGAL PUMP CS/CI DESIGN GPM = 1565 P = 60 PSI
3-8-2	COOL PROCESS CONDENSATE PUMP	1	2	CENTRIFUGAL PUMP CS/CI DESIGN GPM = 72 P = 50 PSI
3-P-3	AMMONIA SCRUBBER BOITOMS PUMP	1	2	CENTRIFUGAL PUMP CS/CI DESIGN GPM = 93 P = 50 PSI
3-ME-1	GAS SCRUBBING UNIT			TEXACO PROPRIETARY

EQUIPMENT LIST TVA MBG MODULE TEXACO GASIFIER SYSTEM NO. 4 - ACID GAS REMOVAL

ITEM	SERVICE	QUA	YTITY	DECCRIPTION
NUMBER	JENTICE	SPARE	OPERATION	DESCRIPTION
4-V-1	SOLVENT KNOCK-OUT DRUM	0	1	
4-V-2	SOLVENT FLASH DRUM	0	1	,
4-V-3	REGENERATOR KNOCK-OUT	0	1	
4- T-1	ACID GAS ABSORBER	0	1	
4- T-2	SOLVENT REGENERATOR	0	1	
4-E-1	SOLVENT CHILLER	0	1	SHELL & TUBE EXCHANGER
4-E-2	SOLVENT COOLER	0	1	SHELL & TUBE EXCHANGER
4-E-3	REGENERATOR FEED PREHEAT EXCHANGER	0	1	SHELL & TUBE EXCHANGER
4-E-4	ACID GAS COOLER	0	1	AIR FAN EXCHANGER
4-E-5	REGENERATOR REBOILER	0 8-2-52	1	SHELL & TUBE — EXCHANGER

EQUIPMENT LIST TVA MBG MODULE TEXACO GASIFIER SYSTEM NO: 4 ACID GAS REMOVER

ITEM	SERVICE	QUA	ANTITY	
NUMBER	JENVICE .	SPARE	OPERATION	DESCRIPTION
4-P-1	SOLVENT CIRCULATION PUMP	1	1	
4-P-2	REGENERATOR REFLEX PUMP	1	1	
4-P-3	SOLVENT TRANSFER PUMP	1	1	
4-TK-1	SOLVENT STORAGE TANK	0	1 .	
4-ME-1	TEG GAS DRYING PACKAGE	0	1	INCLUDES TEG CIRCULATION PUMP, TEG CONTRACTOR AND TEG REGENERATOR
4-C-1	RECYCLE VAPOR COMPRESSOR	0	1	
4-HT-1	RICH SOLVENT LET DOWN TURBINE	0	1	

EQUIPMENT LIST

TVA MBG MODULE
TEXACO GASIFIER
SYTEM NO. 5 SULFUR RECOVER & TAIL GAS CLEAN-UP

ITEM	CERVICE	QUANTITY		
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
5-P-1	SULFUR PRODUCT PUMP	1	1	
5-P-2	STRETFORD SOUR WATER PUMP	1	1	
5-P-3	REDOX SOLUTION CIRCULATION PUMP	1	1	-
5-P-4	STRETFORD SULFUR TRANSFER PUMP	1	1	
5-P-5	REACTION FURNACE AIR BLOWER	1	1	
5-S-ì	AIR OXIDATION MIX	0	1	
5-S-2	STRETFORD SULFUR SEPARATOR MELTOR	0	1	
5-S-3	MOLTER SULFUR STORAGE PIT	0	1	

EQUIPMENT LIST TVA MBG MODULE TEXACO GASIFIER ORIGINAL PAGE IS OF POOR QUALITY

TEXACO GASIFIER
SYSTEM NO. 5 SULFUR REGOVERY & TAIL GAS CLEAN-UP

NOTE: TH	IS ENTIRE UNIT IS DUDI ICATE			GAS CLEAN-UP
ITEM	OUANTITY		TITY	DESCRIPTION
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
5-V-1	ACID GAS KNOCK - OUT DRUM	0	1	
5-R-1	REACTION FURNACE	0	1	ACID GAS BURNER AND STEAM GENERATION
5-R-2,3,4	SULFUR CONVERTERS	0	3	HORI ZONTAL REACTORS
5-R-S	BEAVON REACTORS	0	1	HORIZONTAL GAS HYDROLYSIS REACTOR
5-T-1	BEAVON UNIT CONDENSER	0	1	
5-T-2	STRETFORD ABSORBER	0	1	
5-E-1,2,3,4	SULFUR CONDENSORS	0	4	SHELL & TUBE EXCHANGER
5-E-5	BEAVON UNIT GAS COOLER	0	1	SHELL & TUBE EXCHANGER
5-E-6	STRETFORD RECYCLE COOLER	0	1	

EQUIPMENT LIST TVA MBG MODULE TEXACO GASIFIER

OF POOR QUALITY

SYSTEM NO: 6 - AIR SEPARATION

TEM	CERVICE	QUANTITY		
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
6-ME-1	AIR SEPARATION PLANT	o	1	PACKAGE SYSTEM 1657 TPD 98% OXYGEN
6-C-1	AIR COMPRESSOR	0	1	AXIAL COMPRESSOR
6-C-2	AIR COMPRESSOR	0	1	CENTRIFUGAL COMPRESSOR
6-C-3	ONYGEN COMPRESSOR	0	1	CENTRIFUGAL COMPRESSOR
6-M-1	AIR COMPRESSOR DRIVER	0	1	ELECTRIC MOTOR 24,100 Hp
6-M-2	OXYGEN COMPRESSOR DRIVER	0	1	ELECTRIC MOTOR 12,400 Hp
6-E-1	AIR INTERCOOLER	0	1	AIR COOLED EXCHANGER
é-E-2	AIR AFTERCOOLER	0	1	AIR COOLED EXCHANGER

EQUIPMENT LIST

TVA NEG MODULE

TEMACO GASIFIER SYSTEM NO: 6 - AIR SEFARATION

QUANTITY [FEM SERVICE DESCRIPTION JMBER SPARE OPERATION 6-E-3 AIR TRIM COOLER 0 1 AIR COOLED EXGR. 6-E-4 1st STAGE OXYGEN COOLER 0 1 5.28 MM Btu/hr 6-E-5 2nd STAGE OMYGEN COOLER 0 AIR COOLED EXGR. 1 5.09 MM Btu/hr 6-E-6 3rd STAGE OXYGEN COOLER 0 AIR COOLED EXG. 1 4.85 MM Btu/hr 6-E-7 4th STAGE ONYGEN COOLER 0 AIR COOLED EXGR. 1 6.45 MM Btu/hr 5th STAGE OXYGEN COOLER 6-E-8 0 1 AIR COOLED EXGR. 4.63 MM Btu/hr

EQUIPMENT LIST
TVA MBG MODULE
TEXACO GASIFIER
SYSTEM NO. 13 BY PRODUCT PROCESSING

ITEM	SERVICE	QUA	ANTITY	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
13-T-1	MOLTEN SULFUR STORAGE TANK	0	2	STEAM HEATED 500 TONS EACH
13-P-1	PRILL TOWER FEED PUMP	1	1	MOLTEN SULFUR PUMP
13-S-1	PRILLING TOWER	0	1	
13-S-2	PRILLED SULFUR CONVEYOR	0	1	
13-C-1	AIR BLOWER	1	1	
13-S-3	PRILLED SULFUR STORAGE	0	1	5000 TONS STORAGE BUILDIN

APPENDIX B-3

BABCOCK AND WILCOX COAL GASIFICATION PLANT REFERENCE FACILITY DESIGN

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1.0 INTRODUCTION

The United States, after a number of years of development based on plentiful and inexpensive oil and natural gas, is entering a period of time when it is essential to supplement these energy sources by the increased use of coal. Coal is the nation's most plentiful fossil fuel. Coal gasification is a means of accomplishing this. While utilization of coal through conversion to gaseous products is not new there is no industry within the US which might serve as a base for establishing cost, operational reliability and requirements, and design data for the large scale environmentally acceptable plants needed.

The Tennessee Valley Authority with systems engineering and analysis support from the George C. Marshall Space Flight Center has initiated a project which would establish the commercial base and demonstrate the requirements for gasifying coal in a large integrated facility. The project consists of gasifying 20,000 tens per day of Eastern coal in a four module plant-the construction or which is staggered to accommodate efficient use of construction manpower and product market development.

The purpose of this study is the systematic engineering and cost analysis of five selected gasifier technologies and the other associated systems required to produce medium BTU gas for delivery into a 600 psig distribution system in Northern Alabama. This effort in a series of design studies provides a reference basis for selection of the final plant configuration and choice of process technologies.

The methodology consists of selection of a systematic breakdown of the facility for purpose of study, identification of available technologies for each system, development of a definition level design and cost analysis for each system, conducting a series of trade studies using the definition design as a basis, and finally development of reference facility designs and cost analyses based on the previous work.

This report covers the work done in establishing a Reference Facility Design based on Babcock and Wilcox coal gasification technology. It is one of three such designs, the other two being based on Koppers-Totzek and Texaco coal gasification processes. The design covers a complete grass roots facility site at Murphy Hill, Alabama. The plant inputs are coal, water and electricity. The plant outputs are medium BTU gas (MBG) and solidified sulfur.

The design approach involved breaking each of four identical modules into process and support systems followed by analysis of each system through identification of available technologies and use of the comparative trade studies to select the preferred processes for each system.

2.0 SUMMARY

A four module, 20,000 TPD, based on B&W coal gasification technology has been designed. The plant processes Kentucky No. 9 coal with provisions for up to five percent North Alabama coal. Coal transportation is by river barge except for the Alabama coal which is trucked to the site. Medium BTU gas with heat content of 303 BTU/SCF and not more than 200 ppm sulfur is the primary plant product. Sulfur is recovered for sale as prilled sulfur. Ash disposal is on site. The plant is designed for zero water discharge. Trade studies provided the basis for not using boiler produced steam to drive prime movers. Thus process derived steam in excess of process requirements is superheated for power use in prime movers. Electricity from the TVA grid is used to supply the balance of the plant prime mover power requirements. A study of the effect of minemouth coal cleaning showed that coal cleaning is not an economically preferred route.

The plant design was arrived at by a systematic procedure based on published design work, process trade studies, team engineering experience, a NASA provided module level definition of some twenty systems, and the TVA document, "Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," March, 1980.

The overall plant configuration is shown schematically in Figure 2.1. As shown, General Facilities, Instrumentation and Control, and Coal Handling serve plant wide functions. All other plant components are contained in four identical process modules. The Instrumentation and Control System operates to monitor overall plant performance and to control intermodule relationships. The total list of systems is given in Table 2.1.

The design procedure involved defining available processes to meet the requirements of each system, technical/economic trade studies to select the preferred processes, and engineering design and flow sheet development for each module. Cost studies assumed a staggered construction schedule for the four modules beginning spring 1981 and a 90 percent on stream factor.

The basis and results of the design study are given in Tables 2.2 and 2.3.

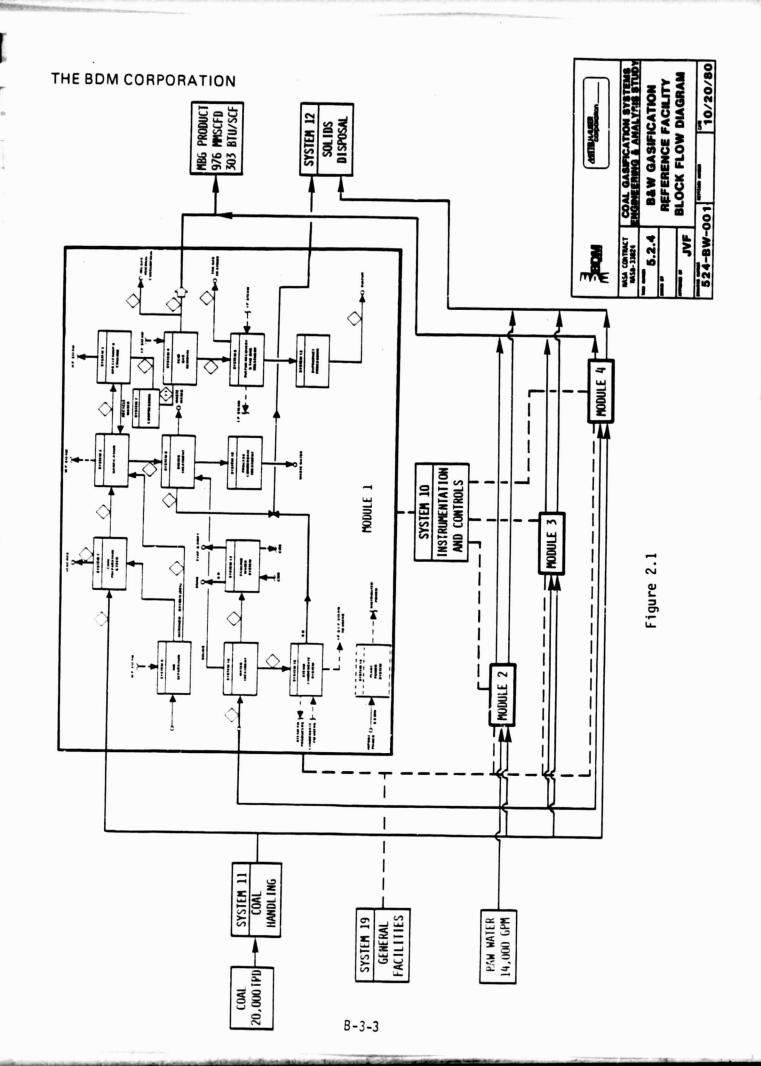


TABLE 2.1. LIST OF SYSTEMS

SYSTEM NO.		COST UNITS PER FACILITY	SYSTEM DESCRIPTION
1	1	4	COAL PREPARATION AND FEEDING
2	3	12	GASIFICATION
3	3	12	INITIAL GAS CLEANUP AND COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	2	8	AIR SEPARATION
7	1	4	COMPRESSION
8	1	4	PROCESS SOLIDS TREATMENT
10	0	1	INSTRUMENTATION AND CONTROL
11	0	1	COAL HANDLING
12	0	1	SOLIDS DISPOSAL
13	1	4	BY-PRODUCT PROCESSING
14	1	4	PLANT POWER SYSTEM
15	1	4	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKE-UP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	1	1	GENERAL FACILITIES

TABLE 2.2. DESIGN BASIS SUMMARY*

PLANT CAPACITY

PLANT TYPE

COAL TYPE

PLANT SERVICE FACTOR

PLANT LIFE

ELECTRICITY SOURCE

WATER

COAL RECEIPT

PRODUCT DELIVERY

PRODUCT QUALITY

SULFUR

COAL COST (1980)

LAND COST (1980)

CLEARING AND GRUBBING (1980)

ELECTRICITY COSTS (1980)

BY-PRODUCT CREDIT

ESCALATION

20,000 TPD

FOUR INDEPENDENT MODULES

KENTUCKY NO. 9

90 PER CENT

20 YEARS EACH MODULE

TVA GRID-4.16KV, 6.9KV, 13.8KV

TENNESSEE RIVER

BARGE +5% BY TRUCK

MBG AT 600 PSIG

MINIMUM 285 Btu/SCF

PRILLED

\$1.25/MM Btu

\$3,000/ACRE

\$2,000/ACRE

NONE

AS PER TVA SPECIFICATION

^{*}Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," Tennessee Valley Authority, March, 1980

TABLE 2.3. DESIGN STUDY RESULTS*

FEED COAL WATER PURCHASED ELECTRICITY MBG PRODUCT MBG PRODUCT MBG QUALITY SULFUR PRODUCT (PRILLED) SULFUR PRODUCT TOTAL CAPITAL REQUIREMENTS** OPERATING AND MAINTENANCE COSTS COAL, CATALYST, CHEMICALS	6,570,000 TPY 14,000 GPM 183 MM KWHY 976 M MSCFD 322 MM MSCFY 303 BTU/SCF 673 LTPD 222,000 LTPY \$3,347 MM (\$2,567 MM) \$138 MM/YR \$181 MM/YR
PLANT OPERATING STAFF	271 PERSONS
MAINTENANCE/YEAR	97 MM

^{*} COSTS ARE IN 1980 DOLLARS
**BASED ON INSTALLATION FACTOR OF 2.31 AND 1.5.
(SEE CHAPTER V.A.)

Designs in this study were based on published works, engineering experience and judgments, and consultations with appropriate process and equipment vendors followed by design calculations, flow process sheet development and documentation.

Cost estimates are based on normal conceptual level design estimating procedures as described in the "Coal Estimation and Economic Evaluation Methodology," BDM/W-80-258-TR-RV2, submitted as another part of this project work scope and reprinted as Appendix E. Schedule estimates are in accordance with overall guidelines furnished by NASA.

Details of the schedule are considered typical for projects of this type and while they do not represent a detailed analysis specific to this project they represent the complexity and illustrate schedule constraints associated with the subject project. Process flow diagrams are included for the main process systems, air separation system, cooling water system, and steam distribution system as aids in following written system descriptions.

Technical results of the study are reported in this appendix. A ccst analysis is included as a part of Appendix D.

3.0 OVERALL FACILITY DESCRIPTION

This section presents and discusses the process design of a plant designed to gasify 20,000 tons per day of coal in four 5000 TPD modules and to produce a medium-Btu product fuel gas (MBG) with a higher heating value of 303 Btu/SCF. The plant design is based upon using entrained flow gasifiers as offered by the Babcock & Wilcox Company, and upon the design criteria (coal characteristics, product specifications, site data, etc.) set forth by the Tennessee Valley Authority in March, 1980.

3.1 Overall Plant Processing

The overall plant configuration is shown in the block flow diagram, see Figure 2.1. Kentucky No. 9 coal is received by barge and either placed in a dead storage pile or processed in System 11, Coal Handling. Provisions have also been made to receive up to five per cent of plant capacity by truck. Dead storage provides for 90 days of supply in interruption.

Coal, either directly from barges or from dead storage, is processed in a Bradford breaker and conveyed to four process modules of 5000 TPD capacity each.

In each module, the raw coal fed to the plant from System 11 is crushed and dried in System 1 and fed to System 2, the Babcock & Wilcox gasifiers, along with unconverted char which is recovered and recycled from the gasifier outlet gas cyclones. The feed coal and recycle char are partially combusted (within the gasifier), with oxygen supplied from the air separation plant, to form a raw product gas which is rich in carbon monoxide and hydrogen. The raw gas leaves the gasifiers at 1800°F and 240 psia, passes through primary cyclones for partial recovery of char recycle, and is then cooled to 450°F in waste heat recovery boilers. Heat recovered in the waste heat boilers and from the gasifiers is used to generate steam for the plant. The raw gas from the waste heat boilers passes through secondary cyclones for additional recovery of char recycle. The gas is then further cooled to 100°F in Venturi scrubbers, which also function to dehumidify the gas and to remove residual char and ash particulates, in System 3. Initial Gas Cleanup and Cooling System.

The raw gas leaves the Venturi scrubbers at 100°F and about 200 psia. It is then compressed in System 7 to about 640-650 psia and sent to System 4, the Acid Gas Removal, where essentially all of the hydrogen sulfide and carbonyl sulfide, and about 45 per cent of the carbon dioxide contained in the raw gas is removed. The acid gas removal system uses the Selexol process (licensed by the Allied Chemical Corporation) in which a physical solvent (the dimethyl ether of polyethylene glycol) absorbs and removes hydrogen sulfide, carbonyl sulfide and the concomitant carbon dioxide from the raw gas.

The acid gas (hydrogen sulfide, carbonyl sulfide and carbon dioxide) is recovered from the Selexol solvent and processed in System 5, Sulfur Recovery and Tail Gas Treatment, wherein over 99 per cent of the hydrogen sulfide and carbonyl sulfide is converted into 188 short tons/day of by-product elemental sulfur. The sulfur recovery unit utilizes the conventional Claus process to convert about 90 to 95 per cent of the hydrogen sulfide (and some of the carbonyl sulfide) in the acid gas into by-product sulfur. The residual tail gas from the Claus unit is then processed in a Beavon-Stretford unit to recover additional by-product sulfur.

Oxygen utilized for the partial combustion of coal and recycle char in the gasifiers is produced in System 7, a conventional air separation plant. Atmospheric air is compressed to about 100 psia, cooled to 100 F and then separated cryogenically into oxygen and nitrogen at about 2 psig. The oxygen is compressed to about 290 psia for use in the gasifiers. A small portion of the nitrogen is also compressed, to about 295 psia, for use in transporting pulverized coal into the gasifiers.

In addition to the systems contained in each module and described in Section 4, the plant consists of System 10, the overall plant monitoring and control system; System 11, the coal handling and storage system serving all four modules; and System 12, the solid waste disposal system serving the total facility.

The purpose of System 10 is to provide operational monitoring and supervisory master control of module operations. The system includes instrumentation to measure key characteristics of all major streams (mass, temperature, pressure, composition, density, and/or others as appropriate) and to display these characteristics in a central location. A dedicated computer is provided to record data of interest at the facility management level. Telecommunications capabilities are provided to communicate with module and systems operations as required to control total plant operation to meet delivery requirements.

System 11, the Coal Handling System, provides for the unloading of coal delivered to the plant either by barge or truck, reclaiming the coal from storage, reducing the size of coal, and transporting the coal to Coal Preparation and Feeding, System 1.

Raw coal is unloaded from barges by the barge unloading subsystem which is designed to unload up to ten 1500-ton capacity barges per shift. The coal is unloaded at an average rate of 1200 tons per hour on a five-day week basis. Coal is transferred by conveyor to a radial stacker which then produces a kidney-shaped coal pile containing live and dead storage. Coal is reclaimed from live storage and conveyed to a Bradford breaker where it is reduced in size to 1.25x0. Coal from trucks is unloaded into a chute from which it is conveyed to the Bradford breaker. Crushed coal from the Bradford breaker is transported to day storage silos from which it is transferred via vibrating

belt conveyors to coal preparation and feeding, System 1, for further processing. Coal fines from the Bradford breaker are collected and sent to the gasification unit, System 2.

The purpose of System 12 is to store solid waste generated by facility operation during the facility life.

This system consists of a lined impounding pit sized to contain 20 years of solid waste. It includes the conveyor system to move the solids to the pit and a leachate recovery area to recover and pump leachate to the process condensate system.

Module descriptions are the subject of Section 4.

3.2 Facility Balances

OUTPUTS:

The key material inputs and product outputs for the overall plant may be summarized as:

INPUTS:	Raw Coal 98% Oxygen Raw Water Intake Imported Electric Power	20,000 short tons/day 15,400 short tons/day 14,000 gpm 68 MW

Product Gas
By-product Sulfur
Solids to Disposal (Wet)
Solids to Disposal (Dry)

Table 3.1 gives the thermal efficiency for the facility. Table 3.2 gives the final product characteristics.

TABLE 3.1. CONVERSION EFFICIENCY - BABCOCK & WILCOX PROCESS 20,000 TPD FACILITY

INPL	<u>JTS</u>	10° BTU/HR	PER CENT
8	COAL TO FACILITY ELECTRIC POWER TO FACILITY	18,300 80	
OUT	PUTS		
3	MBG FROM FACILITY COAL FINES FROM FACILITY	12,235 -0-	
EFF	ICIENCY		
COAL	L-TO MBG (③ + ①) x 100%		66.9
OVE	RALL PRODUCT EFFICIENCY 3 + (1 + 2) x	100%	66.6
OVE	RALL FACILITY EFFICIENCY 3 + 4 + (1) + 0	D) x 100%	66.6

TABLE 3.2. PRODUCT CHARACTERISTICS

	VOLUME PER CENT
CARBON MONOXIDE	63.3
HYDROGEN	30.7
NITROGEN AND ARGON	3.4
CARBON DIOXIDE	2.6
	100.0
HYDROGEN SULFIDE	4.4 PPM VOLUME
CARBONYL SULFIDE	19.9 PPM VOLUME
TOTAL SULFUR	24.3 PPM VOLUME
WATER VAPOR	7.0 LBS/MM SCF
HIGHER HEATING VALUE	303 BTU/SCF
TEMPERATURE	75 ^O F
PRESSURE	615 PSIA

4.0 MODULE DESCRIPTIONS

Each of the four modules in the plant consist of identical sequences of process systems as given in Table 2.1. Process flow diagrams and an equipment list are included in Section 5 for the principal process systems and such auxiliary systems, as cooling water and steam systems.

Each module receives 5000 TPD of coal from System 11, Coal Handling. The sequence of processing shown in Figure 2.1 is repeated in each of the four modules. Descriptions of each modular system are as follows.

4.1 System Descriptions

4.1.1 System 1 - Coal Preparation and Feeding

The coal preparation system receives 5,000 tons/day of crushed, raw coal ranging in size from 1.25" to 0". The raw coal is first pulverized (to 70% passing through a 200 mesh screeg) and dried. Drying is accomplished by sweeping hot flue gas, at 450°F, through the coal pulverizer. The hot flue gas is generated in a direct-fired air heater using MBG produced inplant as fuel. The dry, pulverized coal is separated from the flue gas in a cyclone recovering about 80 percent of the coal. The remaining 20 percent of the pulverized coal (fines) in the flue gas is recovered by venting the flue gas through a fabric filter baghouse. The pulverized coal from the cyclone and from the baghouse sent to a reservoir tank of about 8 hours storage, where it is stored at atmospheric pressure.

Since the gasifier operates at a pressure of 240 psia, it is necessary to pressurize the coal to about 290 psia for transport and injection into the gasifier. Two lock hoppers, each of 20 minutes storage, are pressured with compressed nitrogen. The nitrogen is obtained from the air separation plant. The pressurized coal from the lock hoppers is fed into a pressurized coal feed tank of about 40 minutes storage, from which the pulverized coal is transported to the gasifier using a portion of the compressed nitrogen as the transport medium.

Each of the lock hoppers operates in sequence with this cycle:

- The lock hopper receives coal from the reservoir tank at atmospheric pressure.
- The hopper inlet and outlet valves are closed, and the hopper is pressurized with compressed nitrogen.
- The hopper is held under pressure until feeding is required, as signalled by hydraulic load cells on the gasifier coal feed tank.

- The hopper outlet valve is opened and coal flows into the gasifier coal feed tank.
- The hopper outlet valve is closed and the hopper vent valve is opened. The hopper is de-pressured and compressed nitrogen is vented through a baghouse for recovery of coal fines.
- The de-pressured lock hopper is ready to again receive coal from the coal reservoir tank.

Coal from the gasifier coal feed tank is continuously transported, using compressed nitrogen, into the gasifier. Drawing No. 524-BW-Ol illustrates this system as well as Systems 2 and 3.

4.1.2 System 2 - Gasification

The gasification system includes coal, oxygen and recycle char injection nozzles, the gasifier vessel, slag removal equipment and a steam drum system.

The gasifier vessel consists of a vertical, cylindrical outer steel shell with an inner shell of water-cooled tubes (water wall) in which steam is generated by heat transferred into the tubes from the gasification zone. In the hot reaction zone (lower part of the gasifier), the tubes are covered with a dense refractory to protect the tubes from the molten, flowing slag and yet permit the high temperature required to maintain the gasification reactions and to maintain the slag as a molten fluid. Above the hot reaction zone, the gasification reactions are essentially completed and molten slag is not present. Thus, the inner shell of water-cooled tubes in the upper part of the gasifier is left bare (no refractory covering) to maximize heat transfer from the hot raw gas product into the tubes. The raw gas leaves the gasifier top at 1,800°F and 200 psia, and the molten slag leaves the gasifier bottom at 3,000°F.

Coal recycle char and oxygen are injected into the gasifier at two levels near the vessel bottom. Each level has a ring header with injection nozzles distributed around the periphery of the vessel.

The molten slag formed in the gasifier is a mixture of ash (from the feed coal) and unconverted coal char. The molten slag drains continuously from the gasifier bottom and is quenched and shattered in a water-filled, pressurized slag quench tank. At intermittent intervals, the quenched slag is drained from the quench tank into a slag lock hopper from which the slag is sluiced to dewatering facilities.

Water from an elevated steam drum is pumped into the bottom of the gasifier's water-cooled tubes, and a mixture of steam and water returns from the top of the tubes to the steam drum. The steam-water mixture is separated in the steam drum. The steam is made available for use within the plant and the water is recirculated to the gasifier tubes. Makeup boiler feedwater enters the steam drum and boiler blowdown water is discharged from the drum.

4.1.3 System 3 - Initial Gas Cleanup and Cooling

Complete gasification of the coal is not realized in one pass through the gasifier. Therefore, facilities are provided to recover char (coal ash and unconverted coal) from the raw product gas and to recycle it into the gasifier. Two cyclone stages are provided for that purpose. The hot gas from the gasifier flows through the primary char recovery cyclone stage at 1800°F and is cooled down to 450°F in the WHB (waste heat recovery boiler). The 450°F gas then flows through the secondary char recovery cyclone stage. The collected char from the primary and secondary cyclones is transported back into the gasifier injection nozzles by using steam as the transport medium. The 1800°F char from the primary cyclone passes through a char cooler and is cooled to 1200°F . The recycle char mixture of primary cyclone char (1200°F), secondary cyclone char (450°F) and steam has a temperature of about 900°F entering the gasifiers.

The WHB is designed to remove sensible heat from the hot raw gas as it is cooled from 1800°F down to 450°F and convert the recovered heat into steam. This is accomplished by using an inner shell of water-cooled tubes (water wall) as well as tubes configured cross-current to the gas flow.

The 450°F raw gas leaving the secondary char cyclone still contains a significant amount of unrecovered char and is well above its water dewpoint (calculated to be about 190°F). The gas passes through a Venturi scrubber and is cooled to 100°F, which is below its water dewpoint and therefore water is condensed out of the gas (i.e., the gas is dehumidified). At the same time, the Venturi scrubber removes essentially all of the residual char particulates from the gas. The clean, cool gas then flows to the gas compressors. Since the gas still contains sulfur compounds (hydrogen sulfide and carbonyl sulfide), it is referred to as 'sour' gas.

The gas cooling facilities in the gas cleaning and cooling train may be summarized as:

Waste Heat Boiler: Cools the gas from 1800° F to 450° F and recovers 340 million Btu/hr of heat which is converted into steam. The waste heat boiler includes a steam drum similar to the one described above for the gasifier.

Venturi Scrubber: Cools the gas from 450°F to 100°F and condenses out 22,500 lbs/hr (45 gpm) of water. The total heat removed is 104 million Btu/hr. 70 million Btu/hr is removed by water-cooling a circulating sidestream in the upper section of the Venturi scrubber, and 34 million Btu/hr is removed by makeup water fed to the scruber.

The performance of the char removal facilities in the gas cleaning and cooling train may be summarized as:

Primary Char Cyclone: Removes 85% of the char in the gasifier

outlet gas.

Secondary Char Cyclone: Removes 33% of the residual char (5% of

the original char content).

Venturi Scrubber: Removes essentially 100% of the residual

char (10% of the original char content).

Discussion of Process Design Elements

The raw coal feed rate of 5,000 T/D per gasification module was established in the Tennessee Valley Authority's design criteria. As tabulated later herein, some of the sub-systems within the 5,000 T/D module were designed as single trains and some as multiple trains. For example, the gasifier sub-system was designed as three trains (each of 2,500 T/D coal gasification) with two trains in operation and the third train in standby service. HOWEVER, ALL UF THE DESIGN DATA PRESENTED IN THIS DISCUSSION (AND ON THE GASIFICATION SECTION PROCESS DESIGN DIAGRAM) ARE FOR THE FULL 5,000 T/D COAL GASIFICATION MODULE.

During the work on this project, preliminary design information was made available by the Babcock & Wilcox Company. That information is referred to herein as the 'B&W design', which compares in feed rate with the design herein as follows:

	This Design	B&W Design
Raw coal feed, T/D	5,000	5,368
Raw coal moisture, %	9.6	6.9
Dry coal feed, T/D	4,824	5,000
Dry coal moisture, %	2.0	0.0

The coal drying equipment was designed to reduce the water content of the coal from 9.6 percent in the raw coal to 2.0 percent in the dry, pulverized coal, which requires removing 32,200 lbs/hr of water from the coal. Removing that amount of water, and heating the coal to 150° F, requires about 50 million Btu/hr of heat including an allowance for heat loss. The amount of flue gas, entering the pulverizer at 450° F and leaving at 150° F, must be about 695,000 lbs/hr to transfer 50 million Btu/hr of heat. To generate a

flue gas at 450° F, the direct-fired air heater must be supplied with about 10.5 times the stoichiometric combustion air needed to burn the MBG fuel gas. Thus, the required air rate to the heater is 142,000 SCFM and the fuel required is 70 million Btu/hr.

Pressurizing the coal lock hoppers and feed tank was taken to require about 17,000 lbs/hr. The amount of nitrogen needed to transport dry coal into the gasifiers was also taken to be 17,000 lbs/hr, and that amount of nitrogen was included in the hot product gas from the gasifier in addition to the nitrogen derived from the coal itself.

All of the flue gas from the pulverizer and all of the pressurizing nitrogen was assumed to be vented from the two baghouses in the coal preparation system.

Compressing 34,000 lbs/hr of nitrogen was calculated to require 2,431 HP of compression.

A material balance around the gasifier was developed by pro-rating from the B&W design. Based on that material balance, the carbon conversion in the gasifier was calculated to be 97.5 percent, which coincides with B&W's design information for the selected oxygen-to-coal ratio. The material balance was checked to see that each component (carbon, oxygen, nitrogen, hydrogen and ash) also balanced. After confirming the material balance, an enthalpy data base was developed and the heat balance around the gasifier was calculated. The heat balance indicated that about 500 million Btu/hr must be removed from the gasifier by generating steam, which coincides quite well with the B&W design.

Having confirmed the enthalpy data base, the slag quench duty of 42 million Btu/hr and the recycle char cooling duty of 25 million Btu/hr were calculated by using that enthalpy data base.

Using the same enthalpy data base, a heat curve was developed for the gasifier outlet gas, plotting incremental heat content versus temperature, f $^{\rm m}$ 1800°F down to 100°F. The heat curve indicated that about 340 million Btu/hr could be recovered by cooling the gas from 1800°F down to 450°F, which again coincides quite well with the B&W design and further validates the enthalpy data base which was used.

The heat curve also indicated that cooling of the gas from 450°F to 100°F in the Venturi scrubber would require a total heat removal of 104 million Btu/hr. It is not possible to compare this heat removal duty with that of the B&W design which does not indicate the exit temperatures from their Venturi scrubber. However, a water balance around their scrubber indicates that the B&W scrubber was only cooling the gas down to about 155°F. Since the gasifier heat balance and the waste heat boiler du'y had coincided well with the B&W design, and therefore validated the entherpy data being used,

it was decided that the heat curve developed from that data was correct. Thus, the total heat removal across the Venturi scrubber was taken as 104 million Btu/hr. The design of the Venturi scrubber, which is integrated with the cooling and recovery of the slag sluice water, is discussed in Section 5.05.

4.1.4 System 8 - Process Solids Treatment

This section describes and discusses the design of the gasification section cooling system, which supplies the integrated cooling requirements for quenching the gasifier slag and for supplying part of the Venturi scrubber cooling. As shown in the following diagram (at the end of this section), the integrated gasification water loop functions to recover and reuse the slag sluice water as well as to cool the water. The clarifier in the system serves to dewater the process solids byproduct before transporting to disposal.

4.1.4.1 Description of the Gasification Section Water Loop

The slag sluice water is heated by absorbing the slag quenching duty. Since the water is recycled for reuse, it must be cooled to remove the heat absorbed and the slag solids must be removed from the water. For those purposes, the slag sluice water is circulated through a clarifier (to remove the solids) and a cooling tower (to remove the heat).

The water content of the wet slag sludge removed in the clarifier functions, for all intents and purposes, as the blowdown from the cooling tower. However, the cooling tower requires makeup water to compensate for the blowdown plus the evaporation and drift losses to the atmosphere. The sum of the blowdown water plus the evaporation and drift losses (see gasification water loop diagram) determines the makeup rate to be 268,200 lbs/hr. The makeup water is introduced into the system by first passing through the Venturi scrubber where it functions to remove part of the total heat removed in the scrubber. Since the Venturi scrubber condenses about 22,500 lbs/hr of water from the raw gas, the amount of makeup water needed is reduced to 245,700 lbs/hr. The heat removed by the makeup water (and, in some part, by the condensed water) in passing through the Venturi scrubber amounts to about 34 million Btu/hr. Since the total heat removal duty across the Venturi scrubber is 104 million Btu/hr, the remaining duty of 70 million Btu/hr is supplied by the plant's utility cooling tower via a circulating sidestream cooler on the scrubber.

The flow stream temperatures and compositions, and an overall heat and material balance for the gasification water loop are given in the diagram at the end of this section.

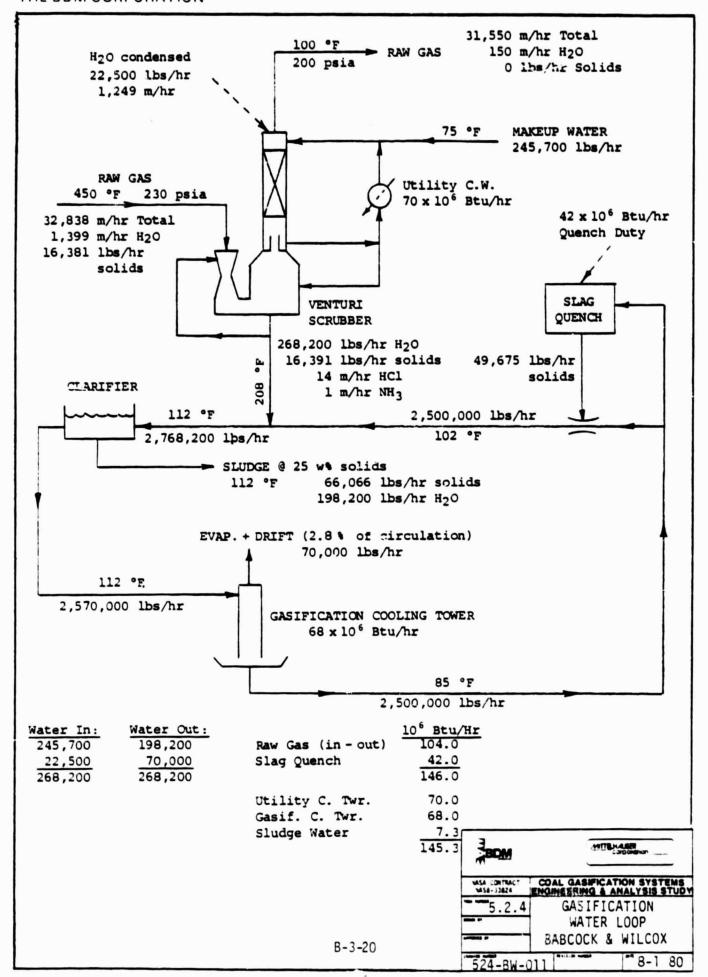
4.1.4.2 Discussion of Process Design Elements

The slag quenching and sluice water rate was taken to be 2,500,000 lbs/hr (5,000 gpm) from the B&W design. The water incurs a 17° F temperature rise in absorbing the 42 million Btu/hr duty required to quench and cool the gasifier slag from 3,000°F to 100° F.

The residual char removed from the raw gas in the Venturi scrubber amounts to 16,381 lbs/hr, which exits with the water leaving the bottom of the scrubber. That water joins the slag sluice water containing 49,675 lbs/hr of the quenched and shattered slag. Thus, the total solids to be removed in the clarifier amounts to 66,066 lbs/hr. Assuming that the sludge from the clarifier is 25 wt percent water, the sludge contains 198,200 lbs/hr of water (the cooling system blowdown) and 66,066 lbs/hr of solids.

The evaporation and drift loss from the cooling tower will be 1 percent of the circulating water rate for each 10°F of cooling across the tower. The drift loss may be taken as 0.1 percent of the circulating water rate. Thus, the evaporation and drift loss from the tower will be 2.8 wt percent of the circulating water rate of 2,500,000 lbs/hr for 27°F cooling across the cooling tower. The evaporation and drift, therefore, amounts to 70,000 lbs/hr. The total makeup required to the system is the sum of the blow-down plus the evaporation and drift, which amounts to 268,200 lbs/hr of water to be derived from the bottom of the Venturi scrubber. Since 22,500 lbs/hr of that amount is derived from condensing water in the Venturi scrubber, the net makeup to the Venturi scrubber must be 245,700 lbs/hr. Setting two conditions that (1) the bottoms water from the Venturi scrubber must not exceed its atmospheric boiling point of 212°F and (2) that the cooling duty not removed by the makeup water will be removed by a circulating sidestream cooler, a heat balance around the Venturi scrubber determined the bottoms temperature of 208°F and the sidestream cooling duty of 70 million Btu/hr. Finally, the mixture of slag sluice water at 102°F and the Venturi bottoms water at 208°F amounts to 2,768,200 lbs/hr of water entering the clarifier at 112°F. The rest of the material and heat balances on the water loop diagram simply involve the appropriate arithmetic.

One item of the above described design still requires some further study. The amount of hydrogen sulfide that will be absorbed in the Venturi scrubber (from the raw gas) and that will remain in the bottoms water at 208°F must be very carefully determined. Since the Venturi bottoms water contain more hydrochloric acid than ammonia, the ammonia will be tied up by the acid and will not be available for tieing up any hydrogen sulfide in the water. Thus, unless the ash and char solids in the water have enough basicity to tie up the hydrogen sulfide, the hydrogen sulfide will be present in the free gaseous state. When that free hydrogen sulfide enters the gasification cooling tower, it will be stripped to the atmosphere with the evaporation and will very probably constitute an environmental problem. If further study indicates that such a problem exists, it may be



necessary to provide a stripper on the bottoms water from the Venturi scrubber to remove any hydrogen sulfide and to route the hydrogen sulfide into the sulfur recovery plant.

4.1.5 System 6 - Air Separation

The air separation plant is designed to provide 3,850 short tons/day of 98 percent oxygen for use in the gasification section.

About 1,484,000 lbs/hr of atmospheric air is compressed, in two stages of compression, to a pressure of 110 psia and aftercooled to 100°F . The compressed air is then cryogenically separated into oxygen and nitrogen in a packaged 'cold box'. The separated oxygen from the cold box (at 2 psig and 70°F) is compressed, in four stages of compression, to 290 psia for use in the gasifiers.

A part of the separated nitrogen (about 34,600 lbs/hr at 2 psig and 70°F) is compressed, in four stages of compression, to 295 psia for use in pressurizing the gasifier coal feed system and in transporting coal into the gasifiers.

Interstage cooling down to 150°F is provided by air-cooling. After-cooling for the compressed air, oxygen and nitrogen is provided by air-cooling down to about 120°F and by water-cooling down to 100°F. The compressed oxygen and nitrogen inputs to the gasification section were aftercooled to 100°F because the B&W design so indicated. However, it would be worth investigating the feasibility of deleting the oxygen and nitrogen aftercoolers. That would provide an additional 20 million Btu/nr of heat input to the gasifiers, making it possible to generate about an additional 20,000 lbs/hr of steam. As discussed in the steam balance section herein, that would be desirable.

The air separation plant was designed as two trains, each in operating service and each providing 1,925 tons/day of oxygen. HOWEVER, ALL OF THE QUANTITIES REFERRED TO IN THIS DISCUSSION ARE FOR THE FULL 3,850 TONS/DAY OF OXYGEN OUTPUT.

4.1.6 System 7 - Raw Gas Compression

The raw gas from the Venturi scrubber $(666,000 \text{ lbs/hr} \text{ at } 100^{\text{O}}\text{F} \text{ and } 200 \text{ psia})$ is compressed, in two stages of compression, to 640 psia and aftercooled to 100^{O}F for processing through the acid gas removal system (Selexol unit).

Interstage cooling down to $130^{0}\mathrm{F}$ is provided by air-cooling. After-cooling is provided by air-cooling down to $120^{0}\mathrm{F}$ and by water-cooling down to $100^{0}\mathrm{F}$.

The raw gas compression train was designed as a single, operating train with no standby train.

SUMMARY OF COMPRESSION

Table 4.1 presents a detailed summary of the compression requirements for the 5,000 T/D coal gasification module, which may be summarized as:

Air compression	72,700 HP
Oxygen compression	19,200 HP
Nitrogen compression	2,430 HP
Raw Gas compression	23,000 HP

For the purposes of the steam balance later herein, it was assumed that (1) the air and oxygen total compression requirement of 91,900 HP could be reduced to 88,000 HP and (2) the raw gas compression could be reduced to 20,800 HP in a final, optimized plant design.

As shown in the steam balance section herein, all of the compressors are driven by steam turbines. The air, oxygen and raw gas compressors are also provided with standby motors for use during plant startup before steam becomes fully available.

4.1.7 System 4 - Acid Gas Removal System

The acid gas removal system utilizes the Selexol solvent process and receives 666,000 lbs/hr of sour gas at 100° F and 635 psia. The sweet (desulfurized) gas leaves the Selexol absorber at about 630 psia and a maximum temperature of 75° F. The component removals across the absorber are:

	<pre>Inlet Gas (mols/hr)</pre>	Outlet Gas _(mols/hr)	% Removal
Hydrogen	9,275.5	9,271.5	0.04
Carbon monoxide	19,115.1	19,110.1	0.03
Carbon dioxide	1,437.1	771.8	46.29
Hydrogen sulfide	451.5	0.2	99.96
Carbonyl sulfide	30.9	0.5	98.38
Nitrogen	1,030.2	1,030.2	0.00
Water	45.5	4.5	90.11
	31,385.8	30,188.8	

The Selexol unit includes an absorption refrigeration unit for chilling the lean solvent so as to achieve the above removals of hydrogen sulfide and carbonyl sulfide. The sour gas from the Selexol unit is rich in ${\rm CO}_2$ and at slightly more than atmospheric pressure.

TABLE 4.1. SUMMARY OF COMPRESSOR HORSEPOWERS

	GAS		Ps	$P_{\mathbf{d}}$	Ts	Td	_
	LBS/HR	MOL WT	psia	psia		°F	HORSEPOWER ^a
OXYGEN COMPRESSOR: 1st Stage 2nd Stage 3rd Stage 4th Stage	314,000 314,000 314,000 314,000	32 32 32 32 32	16.4 33.6 68.0 139.0	35.6 70.0 141.0 292.0	70 150 150 150	240 338 337 340 TOTAL	4,517 4,893 4,858 4,954 19,222 ^b
AIR COMPRESSOR: 1st Stage 2nd Stage	1,484.000 1,484.000	28.7 28.7	14.2 40.0	43.0 110.0	70 150	335 423 TOTAL	35,710 36,987 72,697 ^b
RAW GAS COMPRESSOR: 1st Stage 2nd Stage	666,000 666,000	21.2 21.2	200.0 354.0	356.0 640.0	100 130	232 273 TOTAL	11,025 11,956 22,981 ^b
INERT GAS COMPRESSOR: 1st Stage 2nd Stage 3rd Stage 4th Stage	34,608 34,608 34,608 34,608	28 28 28 28	16.4 34.0 70.0 144.0	36.0 72.0 146.0 297.0	70 150 150 150	247 342 338 335 TOTAL	577 631 617 606 2,431

NOTES:

$$k = 1.4$$
 (c_p/c_v)
 $z_m = 1.0$ (compressibility factor)
 $e_a = 0.78$ (adiabatic efficiency)

All calculated as adiabatic compression, using:

For use in steam balance, these were reduced by 5 to 10 percent on the assumption that an optimized design could achieve such reductions.

4.1.8 System 5 - Sulfur Recovery & Tail Gas Treatment

Acid gas from Acid Gas Removal, System 4, is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean $\rm H_2S$ acid gases. Typically in a Claus type sulfur plant, the acid gas is first passed through a knockout drum before entering the reaction furnace. The chemistry of the process involves converting $\rm H_2S$ to elemental sulfur according to the following equation:

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O$$

Any hydrocarbons in the acid gas are burned to ${\rm CO_2}$ and ${\rm H_2O}$. System 5 is shown schematically in Drawing Number 524-BW-O4.

The reactions are exothermic, and the heat liberated generates steam in the reaction furnace boiler and in the sulfur condenser. The sulfur from each condenser is drained to a recovery pit in Byproduct Processing. System 13 and the tail gas from the final condenser is fed to a Beavon tail gas treating unit where essentially complete removal of the remaining sulfur compounds is achieved before discharge to the atmosphere. The Beavon sulfur removal process is capable of reducing the sulfur content in the tail gas to less than 100 ppm. It is proprietary process. In this system hydrogenation and hydrolysis are used to convert essentially all sulfur compounds to hydrogen sulfide. This gas is then cooled, and passed into a contactor where the hydrogen sulfide is absorbed by the redox solution and oxidized to elemental sulfur. The reduced redox solution is reoxidized by contact with air and subsequently recirculated to the contactor. Elemental sulfur is removed in the air-blowing step as a froth which is pumped to a sulfur melter to be melted under pressure, separated from the redox solution, and transferred to Byproduct Processing, System 13. The decanted redox solution is returned to the system.

The chemical reactions are:

Hydrogenation and Hydrolysis

$$SO_2 + 3H_2 \longrightarrow H_2S + 2H_2O$$

 $S + H_2 \longrightarrow H_2S$
 $COS + H_2O \longrightarrow H_2S + CO_2$
 $CS_2 + 2H_2O \longrightarrow 2H_2S + CO_2$

The system receives 48,370 lbs/hr of acid gas from the Selexol solvent regenerator at about 7 psig and 120° F. The inlet gas composition is:

	<pre>Inlet Gas (mols/hr)</pre>
Hydrogen	4.0
Carbon monoxide	5.0
Carbon dioxide	665.3
Hydrogen sulfide:	451.3
Carbonyl sulfide	30.3
Nitrogen	0.0
Water	99.2
	1,255.1

The recovery of byproduct elemental sulfur from the sulfur recovery system is 185 short tons/day, which amounts to an overall sulfur recovery of about 99.9 percent.

4.1.9 System 13 - By-Product Processing

Sulfur is the only plant byproduct other than ash/char solids which are disposed of on site. Molten sulfur is pumped to a prilling tower in a continuously flowing circulation system. In the tower sulfur is dispensed in droplets through nozzles. Droplets fall counter current to a stream of cooling air and solidify prior to landing in the bottom prill collection section. From the prill tower sulfur is conveyed to a storage building from which it is transferred by truck or barge for sale. Drawing Number 524-BW-08 shows this schematically.

4.1.10 System 14 - Plant Power System

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

- Develop the necessary voltage stepdown arrangement for plant requirements
- Distribute the necessary power to the plant equipment.

This is non-proprietary equipment and is supplied by several U.S. vendors.

TVA's incoming substation transformers receive power from its prevalent distributed voltage switching station and step down this voltage to a medium voltage to supply the plant electrical power requirement for motors, heaters, lighting, and other miscellaneous loads.

The Medium Voltage Electrical Distribution System is a secondary selective system (double ended supply) with several medium voltage buses. Each medium voltage bus receives power from its respective incoming substation transformer through an incoming breaker and supplies power to the medium voltage distribution system through the feeder breakers.

The Low Voltage Electrical Distribution System typically consists of multiple 480 V double ended load centers and 480 V motor control centers (MCC's) supplying the power to 480 V loads throughout the plant. Two load centers are interconnected through a normally open tie breaker. In the event of loss of one load center transformer or its feeder, the 480 V loads of the affected load center are fed by the second load center through the tie breaker.

Each load center consists of an incoming line section, load center transformer, and low voltage section with metal enclosed draw out power circuit breakers.

Load center transformers are air cooled, dry type, 150° F temperature rise, with delta connected primaries and wye connected secondaries. All load center feeder circuit breakers are 1600A frame and 50,000A RMS symmetrical interrupting capacity. The 480 V motor feeder breakers are electrically operated with instantaneous and long time trip units.

480 V MCC's consist of starters, feeder circuit breakers and control devices, assembled in a common structure with horizontal and vertical buses.

A 125 Volt DC System supplies control power for medium voltage and 480 V volt plant switchgear control, protective relaying and annuciation. The system also supplies power for emergency lighting.

4.1.11 System 15 -Steam Generation/Distribution

A steam balance diagram is included in Section 6, Drawing Number 524-BW-10. As shown on that diagram, two of the largest steam demands within the plant are for the steam turbine drives of the air and oxygen compressors in the air separation plant. Since the coal gasifiers are the largest source of heat for generating steam within the plant, it was decided to produce 565 psia steam from the gasifiers and to drive the air and oxygen compressor turbines with that steam. In order to match the 565 psia steam supply (512,700 lbs/hr) with the horsepower demand of the air and oxygen compressors, it was necessary to superheat the steam to 1000 F and to condense the turbine exhausts at 2.5" Hg and 109 F.

The gasification section waste heat boilers were designed to produce 352,100 lbs/hr of 1500 psia steam to be used as follows:

Char recycle transport	32,700 lbs/hr
Sulfur plant (Claus gas reheaters)	20,000 lbs/hr
Raw Gas compressor turbine	299,400 lbs/hr
Total	352,100 lbs/hr

The 299,400 lbs/hr of 1500 psia steam for the raw gas compressor turbine is superheated to $940^{\circ}F$ so that the turbine exhaust would produce 165 psia steam for driving the irert gas (nitrogen) compressor turbine and for supplying other users at that pressure or lower.

The steam balance is very tight. The deaerator is 'out of balance' heatwise by about 13 million Btu/hr or 6.7 percent. That is, the heat entering the deaerator is 6.7 percent (13 million Btu/hr) less than the heat leaving the deaerator, which means that the boiler feedwater temperature will be somewhat lower than the 250° F shown on the steam balance diagram.

At the same time, from the viewpoint of protection against power failures, it would be desirable to drive the boiler feedwater pumps (1500 HP) and at least some of the cooling water circulation pumps (5700 HP) with steam turbines rather than electric motors.

Since the steam balance is about 13 million Btu/hr deficient, and since it would be desirable to drive the boiler feedwater and some of the cooling water pumps with steam turbines, it is apparent that more steam must be made available.

There are a number of options to be explored for making more steam available:

- 1 -- As mentioned earlier in this report, deleting the aftercooling of the compressed oxygen fed to the gasifiers would make about 20 million Btu/hr of more steam available from the gasifiers.
- 2 -- Designing the gasification section waste heat boilers to cool the raw gas down to 400°F (instead of 450°F) would make about 13 million But/hr of more steam available from the waste heat boilers.
- 3 -- Using boiler feedwater to cool the primary cyclone recycle char from 1800°F down to 1200°F (instead of cooling water) would provide another 25 million Btu/hr.

- 4 -- Re-designing the Claus unit to use internal gas-to-gas reheat, or gasified reheat, instead of steam reheat would make another 20,000 lbs/hr of steam available.
- 5 -- Superheating the 1500 psia steam to 1000°F (instead of 940°F) would reduce the consumption of steam required to drive the raw gas compressor.
- 6 -- It would be worth investigating if 1000°F steam temperature for turbine drive steam is the upper limit of current practice.
- 7 -- Although the exhausting of condensing turbines at 2.5" Hg and 109°F is probably as low as can be economically justified with 85°F cooling water, it would be worth investigating the feasibility of 2" Hg and 102°F.
- 8 -- As mentioned earlier in this report, the preliminary steam balance already assumes that the calculated compression horsepower for the air, oxygen and raw gas compressors could be reduced by 5-10% in a final optimized design. That remains to be confirmed or to determine if larger reductions can be achieved.

The above options should readily provide enough additional steam to close the deaerator heat balance and to drive the boiler feedwater pumps with steam turbines. Additional studies are required to determine if additional steam can be made available for driving some of the cooling water pumps with steam turbines.

4.1.12 System 16 - Water Supply

The Raw Water Treatment Unit is designed to provide treated and untreated water for the following facility water systems:

Fire Water Service Water Potable Water Cooling Water Boiler Feedwater

Raw water is pumped from the river to a Fire Water-Raw Water Storage tank.

The Raw Water-Firewater Storage Tank provides surge capacity for Water Treatment as well as storage capacity for firewater. During an emergency, firewater is pumped from the tank to the firewater header

system. The firewater pumps are motor driven and have a diesel engine driven spare. The spare pump is equipped with automatic start-up capability in case of power failure.

The raw water is pumped from the Raw Water-Firewater Storage Tank to the Softener-Clarifier. Lime, alum, and polyelectrolyte from the Clarifier Bulk Chemical Storage and Feed System are added to the Softener-Clarifier, which is equipped with an internal flocculation mechanism. The alum and polyelectrolyte aid in the removal of suspended solids from the raw water. Lime is added during the clarification step to "cold soften" the raw water. Chlorine is added to the raw water to inhibit algae growth in the clarifier and sand filters and reduce organic contamination.

The underflow from the clarifier is a one wt percent sludge and is pumped to solids treatment for further processing.

The clarified and softened raw water from the Softener-Clarifier flows to the Self-Backwashing Sandfilters where additional suspended solids are removed. A pressure differential across the filter bed initiates the backwash cycle. The backwash flows by gravity to the Sandfilter Backwash Sump and is recycled to the Softener-Clarifier. The filtered water flows to the Filtered Water Storage Tank and is utilized as cooling tower make-up for the Process Cooling Tower as service water for general plant use, as feed to the Demineralizer Package, and as feed to the potable water system.

Water intended for potable services is chlorinated and again filtered to meet American Water Works Association (AWWA) standards and stored in a tank sized to hold a day's potable water requirements. The chlorine residual is maintained at 0.5-1.0 PPM free chlorine in the tank.

Filtered water intended as feed to the Demineralizer Package is injected with Sodium Sulfide to remove trace amounts of chlorine which adversely affect the Demineralizer resins and after filtered through activated carbon to remove any remaining organic contaminants and dissolve iron.

In the demineralizer, the mineral sal's present in the water are removed by ion exchange. A two-step demineralization system, utilizing strong cation and strong anion exchangers in series, is provided. A degasifier following the strong cation and magnesium, which the anion exchangers remove anions such as chloride and sulfate. The strong anion exchanger also removes silica. The degasifier is provided to remove carbon dioxide and other dissolved gases.

A mixed bed polisher is provided to remove silica to 0.02 PPM and to polish returned turbine condensate for reuse.

The demineralized water and condensate is preheated to 220-225°F before being pumped into the boiler feedwater deaerators.

Acid, stored in the Acid Tank, is used to regenerate the cation exchangers. Caustic, stored in the Caustic Tank, is used to regenerate the anion exchangers.

Both the Acid and Caustic tanks have pumps and metering devices to control the flow of regenerants to the demineralizer package.

The Boiler Feedwater Deaerating heaters operate at 30 psig and 250° F. The Deaerators reduce the oxygen content of BFW to 0.005 cc/liter.

Hydrazine or sodium sulfite is injected into the storage compartment of the deaerators for chemical scavenging of any residual oxygen. Morpholine is injected into the suction of the boiler feed-water pumps to protect the condensate systems.

A tabulated water balance for the overall plant is given in Table 4.2 at the end of this section. The total raw water intake required, including an appropriate contingency, amounts to 3,500 gpm.

4.1.13 System 17 - Water Cooling

The purpose of this unit is to provide cooling water to the various process users in the facility.

This is a non-proprietary system and is supplied by several US vendors. System 17 is shown schematically in Drawing Number 524-BW-12.

The cooling tower system includes the tower and fans, sidestream filters, circulating water pumps, cold water basin, blowdown system, chemical addition equipment, and distribution system.

Cooling water is pumped from the cold water basin, through the distribution system to the process heat exchangers where low-level, sensible heat is picked up, and back to the cooling tower. The cooling tower rejects low-level heat by evaporative cooling to air drawn through the cooling tower by the cooling tower fans.

A portion of the circulating water is passed through side stream filters to reduce loading to suspended solids, dirt and scale.

The dissolved solids level of the cooling water is maintained by a continuous blowdown stream to the process condensate system. Water level in the cooling tower basis is maintained by continuous make up of the clean water from the raw water treatment system.

The blowdown stream is passed through a blowdown treatment system to recover chromate ions via ion exchange or by chemical reduction to chromium hydroxide and is sent to waste treatment for disposal.

TABLE 4.2. OVERALL WATER BALANCE

INPUTS:	GPM
Makeup to boiler feedwater (BFW) Makeup to gasification cooling tower Makeup to utility cooling tower Raw water treatment makeup TOTAL RAW WATER REQUIRED	114 491 2,357 148 3,110
OUTPUTS:	
Gasification cooling tower evaporation and drift Utility cooling tower evaporation and drift Gasification cooling tower blowdown (BD) Boiler blowdown (BD)	140 1,886 396 47 ^e
Deaerator vent Raw water treatment wastewater ^C	148 ^e
Steam into process: Into gasifiers via char eductor	65
Recovered by condensation in Venturi scrubber Utility cooling tower blowdown (BD) TOTAL OUTPUTS	-45 471 3,110

NOTES:

- a Excluding the water contained in the feed coal
- b Fed into Venturi scrubber (see diagram of gasification water loop)
- C Rinse waters and other wastage
- d Contained in sludge to disposal (see diagram of gasification water loop)
- e Wastewater streams for disposal
- The total raw water demand is estimated to be 3,500 gpm, including an allowance of 390 gpm for utility water and contingency.

Chlorine is added to the cooling water on a routine periodic basis to prevent algae growth. Chemical algicides are added periodically to further eliminate algae growth. Sulfuric acid is added to control pH and zinc and chromate inhibitors are added to the cooling water for corrosion control. Occasionally, a polyphosphate dispersant is added to enhance the action of the inhibitors.

4.1.14 System 18 - Waste Water Treatment

The purpose of this unit is to collect and treat all plant liquid effluent streams. The plant design is predicated on "zero discharge" and permits recycle and reuse of treated water. Streams treated include the following:

- Oily Water Sewers
- · Coal Pile Run Off
- Storm Water Run Off
- Demineralizer Regenerant Wastes and Rinse Water
- Cooling Tower Blowdown
- Sanitary Waste Water
- Gasifier Slag Quench Drains
- Separated Water From Solids Treatment
- Filtrate From Biological Treatment.

Process operations include:

OIL SEPARATOR - streams containing free and dissolved oil and treated in a gravity separator utilizing an emulsion breaking chemical and heat to separate the oil-water mixture.

SOUR WATER STRIPPER - water streams with appreciable $\rm H_2S$ or $\rm NH_3$ residuals are steam stripped to remove these contaminants.

EQUALIZATION BASIN - liquid streams with extremely high or low pH are mixed in an equalizing basin and treated with sulfuric acid or caustic to change the mixed pH to a value of 6.0 - 8.0.

GRAVITY SETTLING-THICKENER - liquid streams with high suspended or dissolved solids are treated in a gravity settler-thickener and mixed with lime, alum, coagulant aids, and polymers to facilitate separation and thickening.

MULTIPLE EFFECT EVAPORATION - neutralized wastes and brines are evaporated to recover water and concentrate the solids.

The recovered, treated water is used as make up to cooling towers or raw water supply. The resultant solids are conveyed to the Solids Disposal System.

4.1.15 System 19 - General Facilities

The purpose of this unit is to provide equipment or services to support the Gasification Facility at the facility level.

The equipment and Services provided in this unit are non-proprietary.

This unit is a general facility category and provides the following equipment and services:

- Administration Building
- Laboratories
- Change Rooms
- Warehouses
- Maintenance Buildings
- Operation Centers
- Security Offices
- Plant Air Facility
- Fire House
- Visitor Reception
- Plant Fencing
- Plant Lighting
- Roads, Bridges
- Docking Facilities
- Interconnection Pipe Ways
- Fire Protection Network
- Flare Stacks and Headers
- Plant Instrument Air Compressors
- Environmental Monitoring
- Site Preparation.

4.2 Module Balances

Material and energy balances have been determined for each module. Table 4.3 gives stream compositions and quantities for each intersystem module. Table 4.4 gives an energy balance at the modular level.

TABLE 4.3
MATERIAL BALANCE
MAN PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

		<	<	<
STREAM NUMBER		(1-10)	⟨o₁-1⟩	1-1
STREAM ID		HAN COAL FEED	TO NATER TO	TO GAS1
TOBUL	FORMULA WEIGHT	LB-MOLS LBS	LB-MOLS LBS	LB-MOLS LBS
CARBON	12.01	253637		253637
HYDROGEN	2.016	17925	_	23000
DXXCEN	32.0	23900		21261
NITROGEN & ARGON N	28.016	5761		05751
	33.06	15450		492
	15.451	766		
CARBON PLOVIDE	78.07			
COUNTY DIOVINE	10.00			
And Londa	750-07			
	30.052			
ana	10.00			
PROPANE	100,11			
N SULPIDE	34 076			
SULFIDE	. 920.09			
	76.13			
SULPUR DIOXIDE SO	64.06			
OXIDE	30.00			
	17.031			
HYDROGEN CYANIDE HCM	27.026			
HYDROGEN CHLORIDE HC1	36.461			
TAPHTHA -				
100 a NOT		05965		25965
ER SOLTOS				
SUBTOTAL. DRY		136817		194307
	18.016	39850		7693
TOTAL, WET		416667		402000
GAS MOLECULAR WEIGHT				
TEMPERATURE, OF				961
PRESSURE, PSIA				290
SCFII			3500	
			0000	

IABLE4.3 (Continued)
MATERIAL DALANCE
84W PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER	\(\frac{\frac{1}{2}}{2}	<-	<:
STREAM ID	VENT GAS PROM COAL		\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\
BYMBOL PONNILA METCHE	LB-MOLS LBS	LB-POLS LBS	TO SOLIDS TR
		NH H	HR
THE REAL PROPERTY.			
DENGER	213111	9275.5 18699	
6 ANGON N	18062 3 506040	19717	
32.06			
MONTON			
38		19115.1	
	364.9		
NE C			
SE CENT			
O S			
8,8			
		20153	
5		6701	
OXIDE			
2			
CYANTOE HCA		9.9	
N CHLORIDE			
ASH			
OTHER SOLIDS		112864	V3783
SUBTOTAL, DRY		\$1508	2777
			66002
	2649.6 44132	1398.2 25190	198606
GAS MOLECULAR WEIGHT	25179.0	32752.8	
TEMPERATURE, OF	9	11.11	1.60107
SCPH SCPH	51	1800	
GPM	9555431	12435113	

TABLE 4, 3 (Continued)
MATERIAL BALANCE
MAW PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER		1	\wedge	·	(<u>;</u>)	<u>`</u>	\ \(\frac{1}{2}\)
		LB-MOLS	LBS	MRG GAS INTERN	AL CONSIDERTION	ACID GAS TO SI	G-FUR BECOVERY
TORNIA	FORMULA WEIGHT	æ z	X X	H.R	H	HR HR	HR
HYDROGEN	19.51						
	3.0.0	6241.0	16614	10.00.0	707		
EN 6 ARGON	28.016	6310				7.6	•
	32.06		21118	114.5	3473		
	15.451						
PONOALDE	28.01	16986.0	622279	2194 1	70,00		
	19.01	0.984	30191	85.A	37,11	5.0	140
	16.042						08767
	20.052						
ENE	10.06						
PROPANE	1007						
	34.036	113					
	50 03			.023	-	451.3	15378
20		666	7	700	,	29.B	1790
SULFUR DIOXIDE SO	90.79						
OXIDE	30.00						
AMMONIA	17.031						
- 1	27.026						
NA BURGEN CHLORIDE HC1	36.461						
TAR & OTT							
ASH							
OTHER SOLIDS							
SUBTOTAL DAY							
WATER		26829.0	550401	3355.0	68827		116601
7	10.01	1.9	02	17	-	113.6	403%0
TOTAL, WET		26832.9	\$504.71			3.0	
TEMPERATURE ALEIGHT		20.51	1	7.6677	60912	1158.4	46650
PRESCRIBE PART		9		16.00		40.27	
SCPH POIN		6115		3		120	
Tag San		AMOUNTOI		4		589	
				2750	2	-	

IADLE 4.3(Continued)
MATERIAL BALANCE
BAW PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER	<u> </u>	<	<
STREAM ID		^ <u>;</u> }	\(\frac{7}{2}\)
	OH SULFUE	OXIDANT TO CASIFIER	Sour cas to Acia cas
CARRON CARRON	TR TR	LB-MOLS LBS	LB-MOLS LBS
T N			22
٦			
82 2	1330.0	2041.4	1869
CARBON MONOTES C1 15.453		600.	1030.2 31252
88			
7	509.0 26802		19115.1 535414
			0070
C. C. B. C. C.			
00.5			
SULFIDE COS	6200.		
55,			15305
200			87
NH.			
N CHLORIDE HC1 27			
4 01L			
ASK			
SUBTOTAL. DRY			
	1939.0		
TOTAL, WET	134 2414	10042.2 322926	31339.8 665837
TEMPERATION WEIGHT	2023 66477	10042.2	45.5 A20
PRESSURE DETA	2.1	32.16	31305.3 666657
SCFH	100		11.24
Mas	786837	240	655
		111771	11915920

IABLEG. 3 (Continued)
MATTERIAL BALANCE
EAN PHOCESS
INTEGRATED FACILITY/MODUSE DEFINITION

	\(\frac{1}{2}\)	(-	\(\frac{1}{1}\)
STREAM 1D	SOLIBS TO DISPOSAL	PREPARATION AND RANGE INC.	arrents (1 or las
EVADOL FORMIA MEIGHT	LB-MOLS LBS	TB-701.S LBS	'
CARBON C 12.01			
70		17925	
		23900	
CHIONINE C)		13430	\$61.1
20.01		727	
a i			
0			
Set 5			
Soo			
NITHOUS OUTGE NO.			
1.1			
HCM			
HC1			
TAR 6 OIL			
OTHER SOLIDS	29650	95495	
SUBTOTAL, DAY	6092		
NTER N20 18.016	22,013	376017	481.1 13424
GAS MOLECULAR METCHE	\$6323	ROCC	
ENPERATURE, OF			13474
SCPN SCPN			259
CPM			

IMBLE 4. 3 (Concluded)
MATERIAL BALANCE
MAN PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

TABLE 4.4

ENERG. BALANCE, 10⁶ BTU PER HOUR
BABCOCK & WILCOX PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

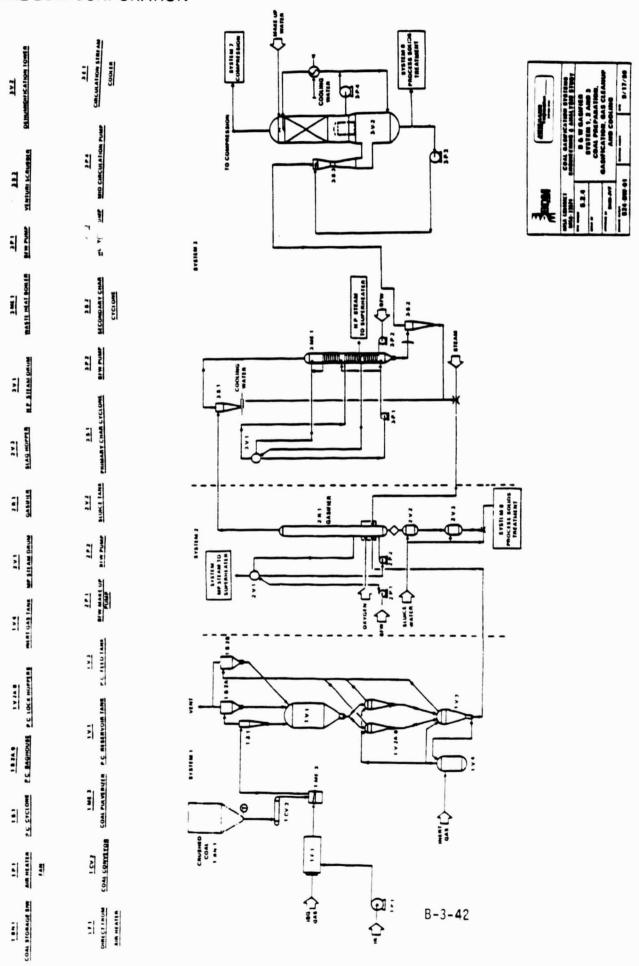
*	TOTAL	4575 13.3 0	4588.3		3081.4	70.7	62.1	69.6	3348.4	1239.9	19.8	-1196.9	-25.0 -1202.1 -1202.1
	SENSIBLE				,	- 0	0.5	16.1 45.8				-476.8	
	LATENT	13.3			•	- 2	·	53.5 15.5				-720.1	
	HII	4575			3061.4	70.7	61.6	,					
	PHASE	Solid Gas			Gas	Sol 1d Gas	Solid	Cas					
ENTITAL PY BALANCE	INLETS	STREAM GOAL AIT TO AIT SEP'N KAW WATER	SUBTOTAL IN	OUTLETS	STREAM MBG	Tall Gas	Sulfar	Air Sep'n Vent	SUBTOTAL OUT	NET INLET-OUTLET	HEAT AND SHAFT WORK ELECTRIC POWER SHAFT WORK	HEAT REJECTION	RADIATION/ECITIVE LOSS SUBTOTAL NET VALUE TOTAL ABSOLUTE VALUE

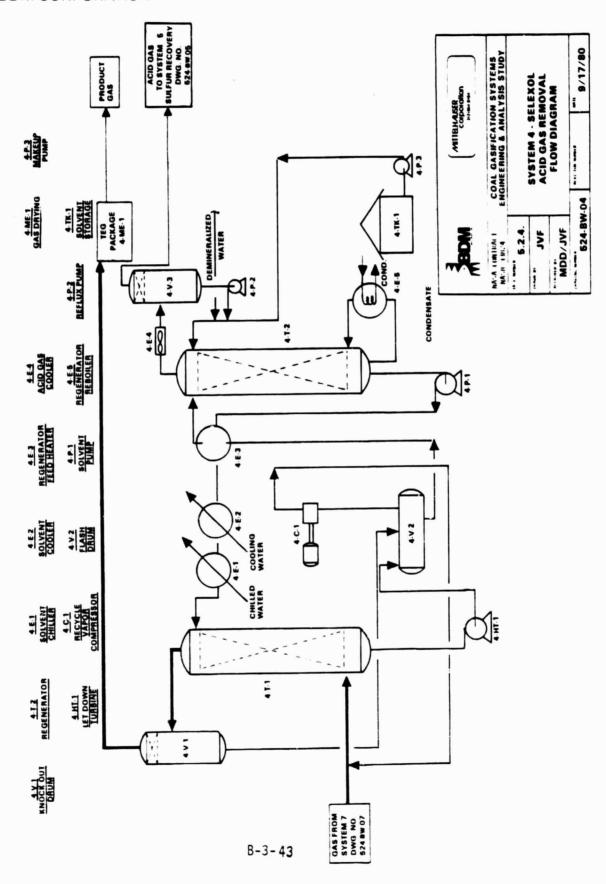
X 100% NET ENTHALPHY + NET HEAT & SHAFT WORK TOTAL ABSOLUTE VALUE OF HEAT & SHAFT WORK BALANCE:

1239.9 - 1202.1 1202.1

5.0 PROCESS FLOW DIAGRAM AND EQUIPMENT LIST

This section contains the process flow diagrams and the list of equipment for each module. The facility contains four times the quantity shown here except for the coal silos which are on a facility basis and System 5, which has two trains in Module 1 and only one train in subsequent modules.



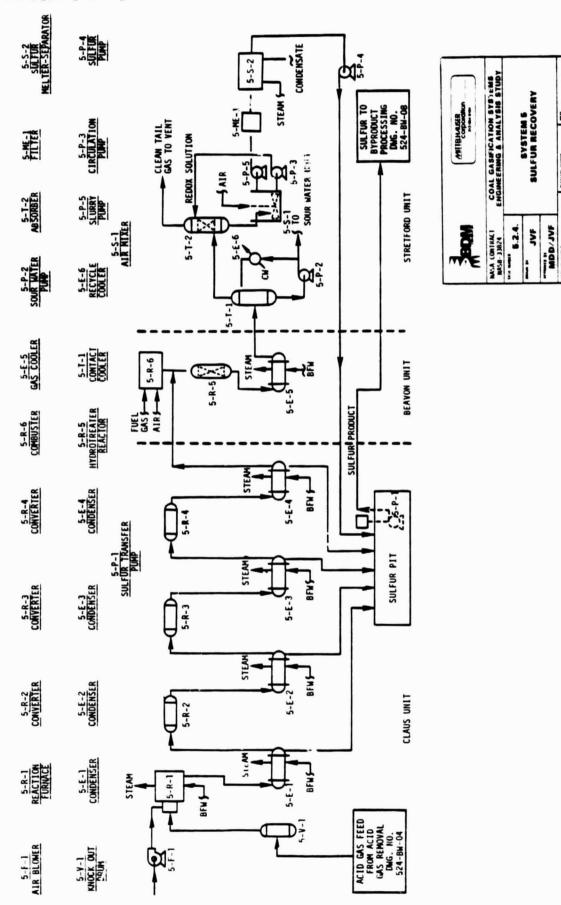


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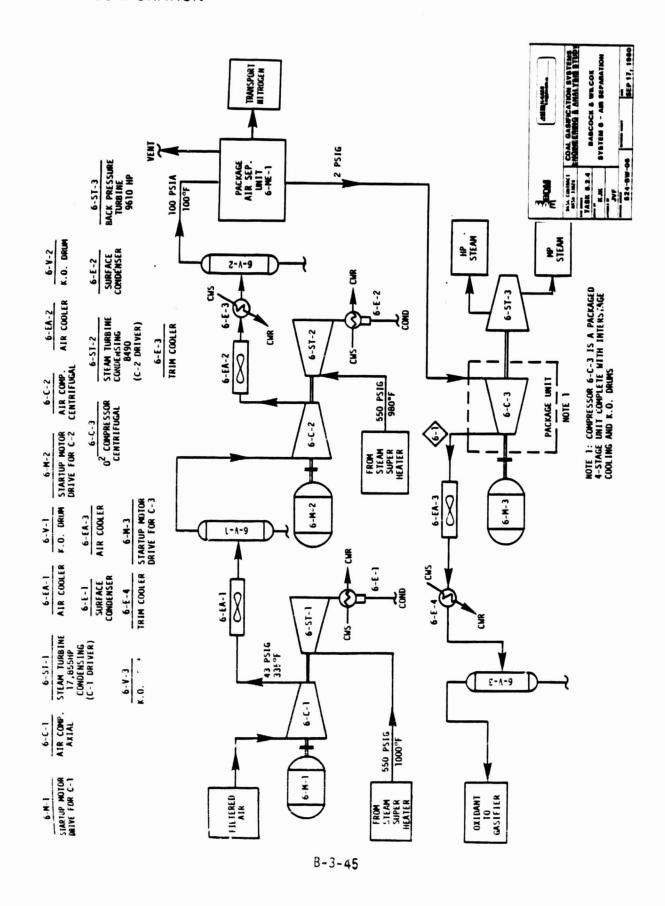
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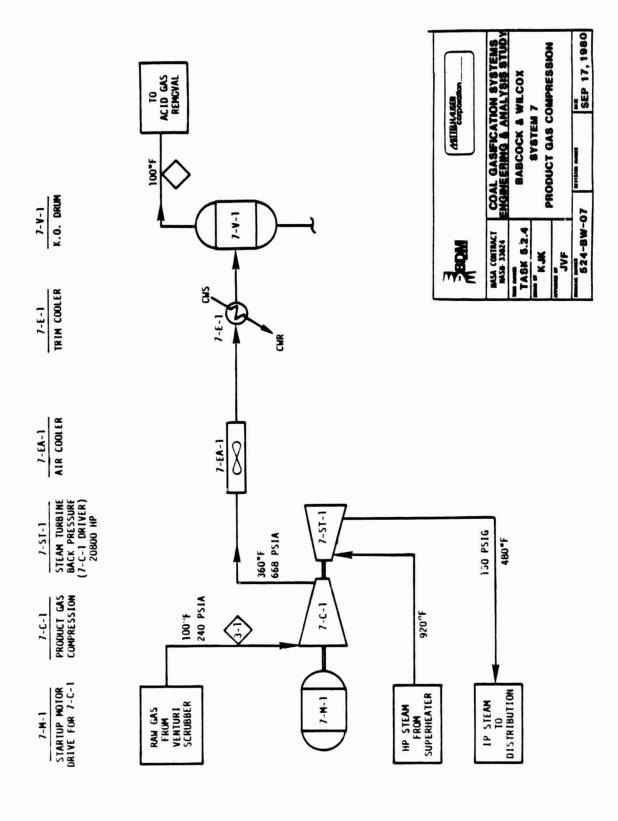
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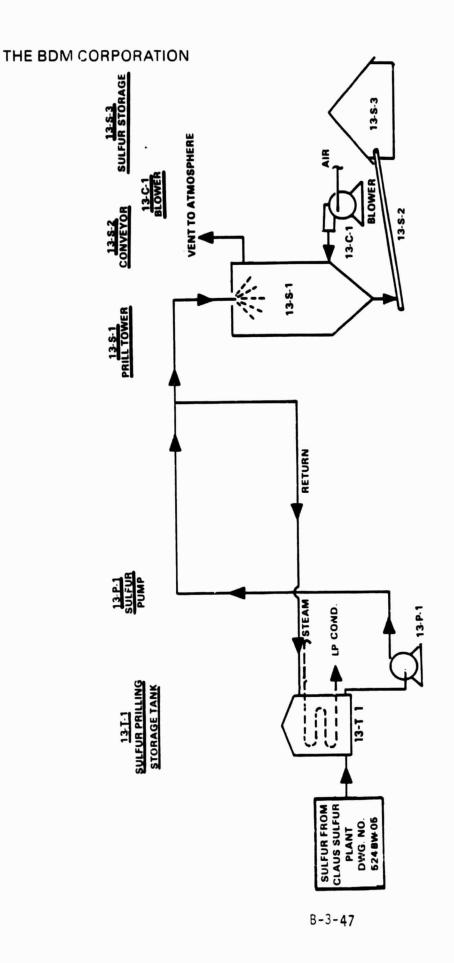


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COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY

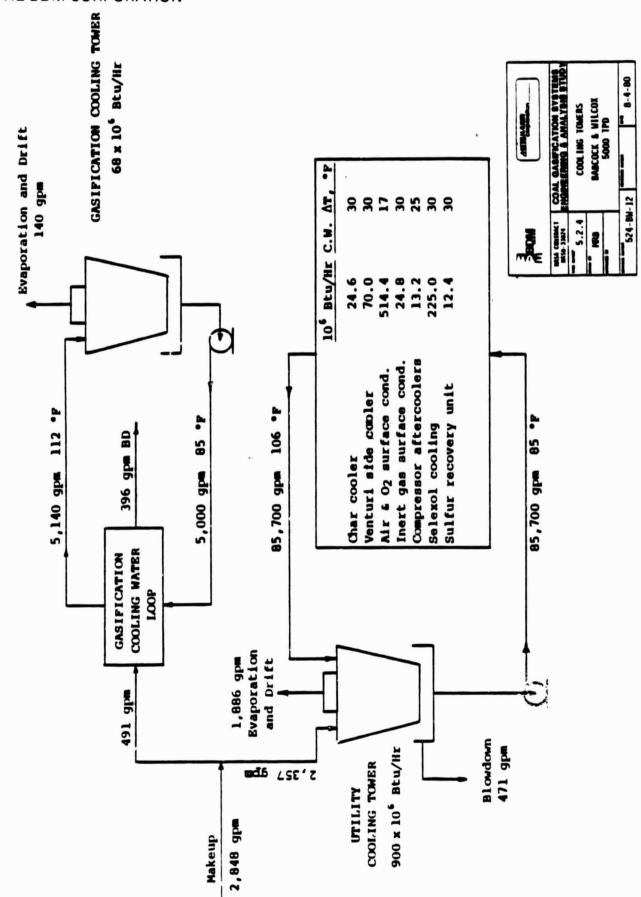
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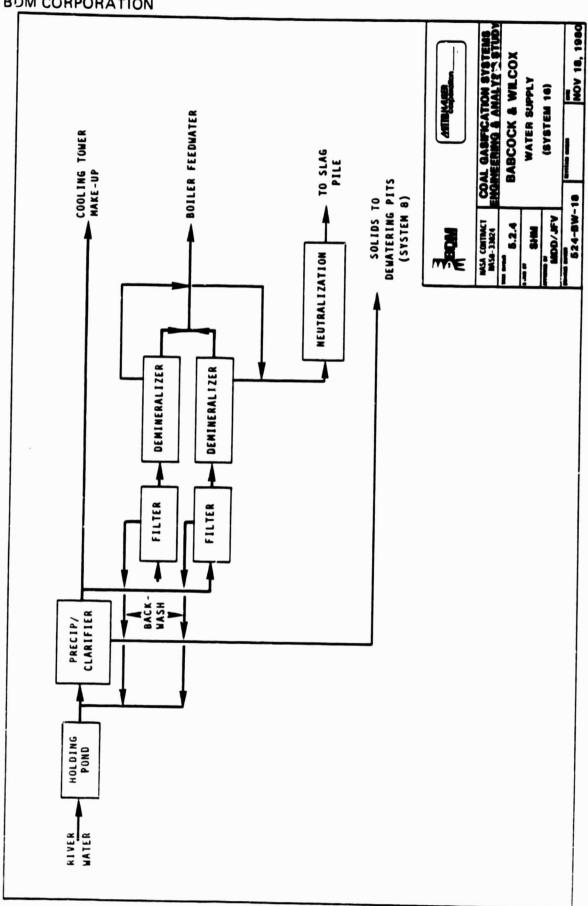
MASA CONTRACT

MIIBHAISER

MONE MONE SYSTEM 13 BY PRODUCT PROCESSING FLOW DIAGRAM



B-3-48



B-3-51

EQUIPMENT LIST
TVA MBG MODULE
B&W GASIFIER
SYSTEM NO: 1 - COAL PREPARATION & FEEDING

ITEM	T	QUA	NTITY	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
11-5-1	COAL STORAGE			COAL SILO
11-S-2	TRAMP IRON REMOVAL		3	MAGNETIC
1-CV-1	COAL TRANSPORT TO SYSTEM 1		3	COVERED CONVEYOR
1-BN-1	FEED SURGE	1	2	700 TON COAL BIN
1-CV-2	CONTROL FEED RATE	1	2	WEIGHT BELT CONVEYOR
1-ME-1	CRUSH COAL TO	1	?	CAGE MILL
1-ME-2	SEPARATE COAL AT 1/4"	1	2	VIBRATING SCREEN
1-5-3	SAMPLE COAL	1	7.	
1-ME-3	FINE GRIND COAL	1	2	PULVERIZER

EQUIPMENT LIST TVA MBG MODULE

B&W GASIFIER
SYSTEM NO: 1 - COAL PREPARATION & FEEDING

	APRIL C	QUA	NTITY	DESCRIPTION
ITEM Number	SERVICE	SPARE	OPERATION	DESCRIPTION
1-F-1	AIR HEAT DRYING	1	2	FIRED HEATER
1-P-1	TRANSPORT AIR SUPPLY	1	2	'AIR BLOWER
1-5-1	RECOVER PULVERIZE COAL			CYCLONES
1-S-2A,B	DUST CONTROL	1	2	BAG HOUSE
1-V-1	FEED SURGE	1	2	P. C. RESERVOIR TANK
1-V-2	PRESSURIZE COAL	1	2	POSSURIZED LOCK HOPPER
1-V-3	FEED CONTINUITY	1	2	FEED TANK
1-V-4	PRESSURIZED INERT CAS	1	2	PRESSURIZED TANK

EQUIPMENT LIST TVA MBG MODULE B&W GASIFIER SYSTEM NO: 2 - COAL GASIFICATION

ITEM	CERTICE	QUAN	ITITY	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
2-P-1	B.F.W MAKEUP PUMP	2	4	CENTRIFUGAL PUMP
2-P-2	BOILER FEED WATER PUMP	2	4	CENTRIFUGAL PUMP
2-V-1	HEAT RECOVERY SERVICE	1	2	STEAM DRUM
2-V-2	SLAG DISCHARGE	1	2	SLUICE TANK
2-1-3	SLAG DISCHARGE	1	2	SLAG LOCK HOPPER
2-R-1	GASIFICATION	1	2	B&W GASIFICATION REACTOR
				c-3
				В

EQUIPMENT LIST

TVA MBG MODULE

B&W GASIFIER

SYSTEM NO: 3 - INITIAL GAS CLEANUP & COOLING

ITEM	SERVICE	QUA	NTITY	DECCRIPTION
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
3-P-1	GAS DEHUMIDIFICATION	2	4	BOILER FEED WATER CENTRIFUGAL PUMP
3-P-2	GAS DEHUMIDIFICATION	2	4	BOILER FW MAKEUP CENTRIFUGAL PUMP
3-P-3	FINAL GAS DEHUMIDIFI- CATION	2	4	CENTRIFICAL WATER RECYCLE PUMP
3-P-4	GAS DEHUMIDIFICATION	2	4	MID TOWER CENTRIFUGAL CIRCULATION PUMP
3-V-1	HP STEAM PRODUCTION	1	2	STEAM DRUM
3-V-2	GAS DEHUMIDIFICATION	1	2	PACKED COLUMN GAS COOLER
3-ME-1	GAS COOLING	1	2	WASTE HEAT BOILER
3-E-1	GAS DEHUMIDIFICATION	2	4	CIRCULATION STREAM COOLER HEAT EXCHANGER
3-5-1	GAS CLEANUP			PRIMARY CYCLONE B-3-55

ON EQUIPMENT LIST
TVA MBG MODULE
B&W GASIFIER
SYSTEM NO: 3 - INITIAL GAS CLEANUP & COOLING

ITEM		OUA	NTITY	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
3~S-2	GAS CLEANUP			SECONDARY CYCLONE
3-5-3	GAS CLEANUP	1	2	VENTURI SCRUBBER
				·
				B-3-

EQUIPMENT LIST TVA MBG MODULE B&W GASIFIER SYSTEM NO: 4 - ACID GAS REMOVAL

ITEM	4500.05	QUA	TITY	DECCRIPTION
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
4-V-1	SOLVENT KNOCK-OUT DRUM	0	1	
4-V-2	SOLVENT FLASH DRUM	0	1	
4-V-3	REGENERATOR KNOCK-OUT	0	1	
4-T-1	ACID GAS ABSORBER	0	1	
4-T-2	SOLVENT REGENERATOR	0	1	
4-E-1	SOLVENT CHILLER	0	1	SHELL & TUBE EXCHANG
4-E-2	SOLVENT COOLER	0	1	SHELL & TUBE EXCHANG
4-E-3	REGENERATOR FEED PRE- HEAT EXCHANGER	0	1	SHELL & TUBE EXCHANG
4-E-4	ACID GAS COOLER	0	1	AIR FAN EXCHANGER B-3-5

EQUIPMENT LIST TVA MBG MODULE B&W GASIFIER SYSTEM NO: 4 - ACID GAS REMOVAL

ITEM	SERVICE		NTITY	DESCRIPTION	
NUMBER	JENTIGE	SPARE	OPERATION	DESCRIPTION	
4-E-5	REGENERATOR REBOILER	0	1	SHELL & TUBE EXCHANGE	
4-P-1	SOLVENT CIRCULATION PUMP	1	1		
4-P-2	REGENERATOR REFLEX PUMP	1	1		
4-P-3	SOLVENT TRANSFER PUMP	1	1	·	
4-TK-1	SOLVENT STORAGE TANK	0	1		
4-ME-1	TEG GAS DRYING PACKAGE	0	1	INCLUDES TEG CIRCULATI PUMP, TEG CONTRACTOR AND TEG REGENERATOR	
4-C-1	RECYCLE VAPOR COM- PRESSOR	0	1		
4-HT-1	RICH SOLVENT LET DOWN TURBINE	0	1		
				B-3-58	

ON EQUIPMENT LIST
TVA MBG MODULE
B&W GASIFIER
SYSTEM NO: 5 - SULFUR RECOVER & TAIL GAS CLEAN-UP

ITEM	CERTICE	QUA	YTITY	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
5-P-1	SULFUR PRODUCT PUMP	1	1	
5-P-2	STRETFORD SOUR WATER PUMP	1	1	
5-P-3	REDOX SOLUTION CIRCULATION PUMP	1	1	
5-P-4	STRETFORD SULFUR TRANSFER	1	1	· · · · · · · · · · · · · · · · · · ·
5-P-5	REACTION FURNACE AIR BLOWER	1	1	
5-S-1	AIR OXIDATION MIX TANK	0	1	
5-S-2	STRETFORD SULFUR SEPARATOR MELTOR	0	1	
5-S-3	MOLTER SULFUR STORAGE	0	1 .	
5-V-1	ACID GAS KNOCK-OUT DRUM	0	1	В

EQUIPMENT LIST
TVA MBG MODULE
BAW GASIFIER
SYSTEM NO: 5 - SULFUR RECOVER & TAIL GAS CLEAN-UP

ITEM	CERVICE	QUA	YTITY		
NUMBER	SERVICE	SPARE OPERATION		DESCRIPTION	
5-R-1	REACTION FURNACE	0	1	ACID GAS BURNER AND STEAM GENERATION	
5-R-2,3,4	SULFUR CONVERTERS	0	2	HORIZONTAL REACTORS	
5-R-S	BEAVON REACTORS	0	1	HORIZONTAL GAS HYDROLYSIS REACTOR	
5-T-1	BEAVON UNIT CONDENSOR	0	1		
5-T-2	STRETFORD ABSORBER	0	1		
5-E-1,2,3,4	SULFUR CONDENSORS	0	4	SHELL & TUBE EXCHANGE	
5-E-5	BEAVON UNIT GAS COOLER	0	1	SHELL & TUBE EXCHANGE	
5-E-6	STRETFORD RECYCLE COOLER	0	1		
				B-3-60	

EQUIPMENT LIST TVA MBG MODULE B&W GASIFIER SYSTEM NO: 6 - AIR SEPARATION

,				
ITEM	CERVICE	QUA	NTITY	T
NUMBER	SERVICE	SPARE	OPERATION	. DESCRIPTION
6-C-1	AIR COMPRESSION	0	2	AXIAL COMPRESSOR
6-C-2	SECOND AIR COMPRESSOR	0	2	CENTRIFUGAL COMPRESSOR
6-C-3	OXYGEN COMPRESSOR	0	2	CENTRIFUGAL COMPRESSOR
6-M-1	START UP MOTOR FOR 6-C-1		2	
6-M-2	START UP MOTOR FOR 6-C-2		2	
6-M-3	START UP MOTOR FOR 6-C-3		, 2	
6-ST-1	STEAM TURBINE DRIVE FOR 6-C-1		2	
6-ST-2	STEAM TURBINE DRIVE FOR 6-C-2		2	
6-ST-3	STEAM TURBINE DRIVE FOR 6-C-3		2	B-3-61

EQUIPMENT LIST TVA MBG MODULE B&W GASIFIER SYSTEM NO: 6 - AIR SEPARATION

ITEM	SERVICE	QUANTITY		DESCRIPTION
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION
6-V-1	FIRST AIR K.O. DRUM		2	
6-V-2	SECOND AIR K.O. DRUM		2	
6-V-3	OXYGEN K.O. DRUM		2	
6-EA-1	FIRST AIR COOLER		2	AIR FIN
6-EA-2	SECOND AIR COOLER		2	AIR FIN
6-EA-3	OXYGEN COOLER		2	AIR FIN
6-E-1	TURBINE EXHAUST CONDENSOR		2	SURFACE CONDENSOR
6-E-2	TURBINE EXHAUST CONDENSOR		2	SURFACE CONDENSOR
6-E-3	AIR TRIM COOLER		2	COOLING WATER EXCHA

EQUIPMENT LIST TVA MBG MODULE B&W GASIFIER SYSTEM NO: 6 - AIR SEPARATION

NUMBER SERVICE SPARE OPERATION DESCRIPTION			QUANTITY		QUANTITY			ITEM
6-E-4 AIR TRIM COOLER 2 COOLING WATER EX	ON	DESCRIPT			SERVICE	ITEM Number		
	XCHANGE	COOLING WATER	2		AIR TRIM COOLER	6-E-4		
						•		
						-		

EQUIPMENT LIST TVA MBG MODULE B&W GASIFIER SYSTEM NO: 7 - GAS COMPRESSION

ITEM		QUANTITY		QUANTITY		
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION		
7-C-1	PRODUCT GAS COMPRESSION	0	1	CENTRIFUGAL COMPRESSOR		
7-M-1	START UP MOTOR FOR 7-C-1	O	1			
7-ST-1	BACKPRESSURE TURBINE DRIVE FOR 7-C-1	0	1			
7-EA-1	COMPRESSOR DISCHARGE COOLER	0 .		AIR FIN		
7-E-1	PRODUCT GAS TRIM COOLING	0	1	COOLING WATER EXCHANGE		
7-V-1	COMPRESSED GAS CON- DENSATE REMOVE	0	1	K.O. DRUM		

EQUIPMENT LIST TVA MBG MODULE B&W GASIFIER SYSTEM NO: 13 - BY PRODUCT PROCESSING

ITEM		QUANTITY	QUAN	NTITY	
NUMBER	SERVICE	SPARE	OPERATION	DESCRIPTION	
13-T-1	MOLTEN SULFUR STORAGE STORAGE TANK	0	2	STEAM HEATED 500 TONS EACH	
13-P-1	PRILL TOWER FEED PUMP	1	1	MOLTEN SULFUR PUMP	
13-5-1	PRILLING TOWER	0	1		
13-5-2	PRILLED SULFUR CON- VEYOR	0	1		
13-C-1	AIR BOILER	1	1		
13-5-3	PRILLED SULFUR STORAGE	0	1	5000 TONS STORAGE BUILDING	
				B-3-	

APPENDIX B-4

LURGI AND BGC/LURGI COAL GASIFICATION PLANTS

DEFINITION LEVEL DESIGNS

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1.0 INTRODUCTION

The United States, after a number of years of development based on plentiful and inexpensive oil and natural gas, is entering a period of time when it is essential to supplement these energy sources by the use of coal. Coal is the nation's most plentiful fossil fuel. Coal gasification is a means of accomplishing this. While utilization of coal through conversion to gaseous products is not new, there is no industry within the US which might serve as a base for establishing cost, operational reliability and requirements, and design data for the large scale environmentally acceptable plants needed.

The Tennessee Valley Authority with systems engineering and analysis support from the George C. Marshall Space Flight Center has initiated a project which would establish the commercial base and demonstrate the requirements for gasifying coal in a large integrated facility. The project consists of gasifying 20,000 tons per day of Eastern coal in a four module plant, the construction of which is staggered to accommodate efficient use of construction manpower and product market development.

The purpose of this study is the systematic engineering and cost analysis of five selected gasifier technologies and the other associated systems required to produce medium Btu gas for delivery into a 600 psig distribution system in Northern Alabama. This effort in a series of design studies provides a reference basis for selection of the final plant configuration and choice of technologies.

The methodology consists of selection of a systematic breakdown of the facility for purpose of study, identification of available technologies for each system, development of a definition level design and cost analysis, conducting series of trade studies using the definition design as a basis, and finally development of reference facility designs and cost analyses based on the previous work. This report covers the definition design efforts for two of the five designs—the Lurgi and BGC/Lurgi gasification systems. The designs for the other gasification systems—the Koppers—Totzek, the Texaco and the Babcock and Wilcox—appear in Appendices B-1, B-2 and B-3, respectively.

The trade studies were done primarily based on the Texaco and K-T cases and previous to this work with the other three gasifiers. Therefore, the designs for the Lurgi and BGC/Lurgi that are presented here, though they are of the first level effort, have the results of the trade studies built in.

This report contains a description of the plant configuration chosen, a description of the technology chosen for each system, the material and energy balances. The cost analysis for the Lurgi and BGC/Lurgi is presented in Appendix D.

2.0 SUMMARY

Definition level plant designs have been prepared based on Lurgi and BGC/Lurgi coal gasification processes. Each of these plants process 20,000 TPD total raw coal to the gasifiers.

The Lurgi coal gasification process consists of reacting one inch by twenty-eight mesh coal with oxygen and steam in a fixed bed reactor to produce medium BTU gas. Coal is fed into the top of the reactor from where it moves downward by gravity countercurrent to the oxygen steam and gasification products. Ash and unreacted carbon are removed from the bottom by a mechanical grate which is kept relatively cool by the flow of steam and oxygen into the bottom of the reactor. A portion of the steam requirement is generated in the reactor jacket. Products, partially cooled by contact with coal feed are further cooled by a quench and waste heat boiler. Heavy products, ammonia and some H₂S are recovered from the quench liquor in the waste water treating section.

The BGC/Lurgi is similar to the Lurgi in description except that ash is removed from the bottom of the reactor as molten slag. Molten slag is removed from the reactor by quenching in a water bath to form ash granules which are removed as a slurry. Since there is no grate cooling requirement the steam requirement is much less than that of the Lurgi. The result is that lower quantities of vapor state materials per unit of coal allow increased coal throughout per unit reactor cross section, higher maximum temperatures with less methane production, and less hydrogen production.

Coal fines produced in the feed handling preparation steps of the Lurgi and BGC/Lurgi designs are either burned as boiler fuel or exported as plant byproducts. The plants are each designed to be built in four distinct stand alone modules processing 5000 TPD each. Exceptions are such facility wide sections as coal handling and general facilities. Module construction is staggered to allow efficient use of construction manpower and the development of product markets. Tars, oils and phenolic compounds are burned as boiler fuel along with coal fines to produce steam requirements for the process units. No additional steam for prime mover power is generated. Purchased electricity is utilized to fill all power requirements not covered by process generated steam. Because of the decision to burn tars, oils and phenolic compounds the plants products are medium BTU gas, sulfur, and ammonia in the Lurgi and BGC/Lurgi cases. All plant solids are disposed of on site. The plants are designed for recycle of treated waste water with zero plant effluent.

Tables 2.1 and 2.2 give a summary of the process and cost results obtained in this study.

TABLE 2.1
FACILITY PROCESS RESULT SUMMARY*

ITEM	LURGI	BGC/LURGI
COAL, T/YR FINES SOLD, T/YR ELECTRICITY, 1000 Kwh/YR	10,000,000	10,000,000 3,918,000
YIELD MMSCFD	1,160	960
HHV BTU/SCF COMPOSITION 1b/hr	308	384
Hydrogen	29,968	15,183
Ni trogen	3,928	3,889
Carbon Monoxide	152,127	438,365
Carbon Dioxide	363,892	20,591
Methane	45,720	36,664
Ethane	2,371	1,901
Light HC	3,892	2,998
Hydrogen Sulfide	61	70
Carbonyl Sulfide	300	192
Water	84	70
Total	602,343	519,923
SULFUR 1000 TONS/YR	58	59
AMMONIA 1000 TONS/YR	89	89
EFFICIENCY - PERCENT**	68.0	76.4

^{*} PER YEAR EXCEPT WHERE NOTED

^{**} EFFICIENCY BASED ON COAL CONSUMED, PURCHASED ELECTRICITY AS ELECTRICITY AND MBG PRODUCT.

TABLE 2.2 FACILITY COST SUMMARY

ITEM	LURGI	BGC/LURGI
CAPITAL REQUIREMENTS (1980 DOLLARS)	2,747,000	2,061,000
ANNUAL OPERATING COSTS COAL @ 1.25/MMBTU OTHER	\$274 MM 93	\$274 MM 35
PRODUCT COSTS (CONSTANT 1980 DOLLARS PER MMBTU)	5.26	4.06

3.0 GENERAL FACILITY DESCRIPTION

The TVA coal gasification facility is to be designed to process 20,000 TPD of Kentucky Number 9 coal into medium BTU gas having not more than 200 ppmv sulfur. The coal specifications are given in Table 3.1. Design specifications were given in a TVA document, "Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant", March 1980, provided by NASA. In order that the most efficient use of construction labor might be accomplished and in order to allow time for development of a user network for the MBG product and provide for maximum operating flexability the plant is to be constructed in four 5000 TPD modules.

The plant has been divided into a group of interrelated systems according to the guidelines given in Table 3.2 for purposes of systematically studying the available processes for inclusion in the plant design. A summary of the processes available for each system has been published earlier in "Coal Gasification Catalog" as a part of Task 5.1 of this study. For each of the five gasification technologies to be considered for the plant a combination of systems was selected to provide a definition level design. The purpose of definition level designs is to provide the starting point for a series of engineering analyses and trade studies which result in a more definitive and optimin plant configuration and process selection in the next level of design. The two designs presented in this report, while catagorized as definition level designs, have the benefit of earlier trade studies performed based on definition level Texaco and Koppers-Totzek designs. Therefore they represent an advanced version of definition level design.

3.1 <u>Methodology of Design</u>

The selection of systems defined for the plant have been arranged for each of the two gasification processes as shown in the block flow diagrams in Sections 4.0 and 5.0. For each system an analysis of available processes and technologies was accomplished based on:

- A system level approach
 - select applicable system design from design data base
 - data availability and relevance
 - adjustments for current application
- No secrecy agreements required
- Production of system level block flow diagrams plus documentation of I/O, interfaces, costs and other requirements as the basis for trades and second level reference facility designs.

For each process selected available published designs were studied. Those designs most nearly meeting the requirements of this effort were chosen as a basis for defining the requirements of the definition level design. These designs were scaled and adjusted as required to meet the

TABLE 3.1 COAL CHARACTERISTICS

	Mean 1	Standard Deviation ²
H.H.V., as Received, Btu/lb	10.980	547.6
Total Moisture, wt %	9.564	1.878
Inherent Moisture, wt %	3.25	0.75
Bulk Density, as Received, 1b/ft3		•
Free Swelling Index ³	(3.0-6.5)	-
Grindability Index	59.	5.117
2.1.2 Proximate Analysis, Dry	7, wt %	
Volatile Matter	37.54	1.878
Fixed Carbon	46.63	1.604
Ash	15.83	3.086
TOTAL	100.00	
2.1.3 Ultimate Analysis, Dry,	, wt %	
Carbon	67.31	2.794
Hydrogen	4.757	.2409
Nitrogen	1.529	.001326
Oxygen	6.343	1.085
Sulfur	4.100	. 4858
Ash	15.83	3.086
Chlorine	.1310	. 05965
TOTAL	100.000	

¹Mean = $\bar{x} = \frac{n}{2}x_1/n$, based on analysis of 14 samples of Kentucky No. 9 coal received by TVA from various mines during the period 1972-1977; results are rounded to four significant figures.

²Standard Deviation = 1/(n-1) $\{\frac{n}{\Sigma}(x_1-x_2)^2\}^{1/2}$, based on analysis of the 14 sample coals; results are rounded to four significant figures.

 $^{^3}$ Data indicates range of values; for design purposes, assume FSI = 6.5.

TABLE 3.2

SYSTEM DESCRIPTION	COAL PREPARATION & FEEDING	GASIFICATION	INITIAL GAS CLEANUP & COOLING	ACID GAS REMOVAL	SULFUR RECOVERY	AIR SEPARATION	COMPRESSION	PROCESS SOLIDS TREATMENT	INCINERATOR	INSTRUMENTATION & CONTROL	COAL MANDLING	SOLIDS DISPOSAL	BYPRODUCT PROCESSING	PLANT POWER SYSTEM	STEAM GENERATION/DISTRIBUTION	RAW WATER MAKEUP	COOLING WATER SYSTEM	WASTE WATER TREATMENT	GENERAL FACILITIES
SYSTEM NO.	-	2	3	4	2	9	7	80	6	9	=	12	13	14	15	91	17	81	19

TABLE 3.3 SYSTEM TECHNOLOGY ASSESSMENT

SYS	SYSTEM	LURGI	BGC-SLAGGER
-	COAL PREPARATION & FEEDING	d C	83
2	GASIFICATION	S	2
٣	INITIAL GAS CLEANUP & COOLING	ಕ್ರಿ	ಕ್ರಿ
4	ACID GAS REMOVAL	5	క
2	SULFUR RECOVERY & TAIL GAS TREATMENT	5	5
9	AIR SEPARATION	3	ಕ್ರಿ
7	COMPRESSION	ಕ್ರಿ	ಕ್ರಿ
8	PROCESS SOLIDS TREATMENT	3	3
6	INCINERATOR		
10	INSTRUMENTATION & CONTROL	G C	3
=	COAL HANDLING	3	ಕ್ರಿ
12	SOLIDS WASTE RECYCLING/DISPOSAL	3	3
13	BYPRODUCT PROCESSING	CP/CA	CP/CA
14	PLANT POWER SYSTEM	ಕ್ಷಿ	ಕ್ರಿ
15	STEAM GENERATION/DISTRIBUTION	CP/CA	CP/CA
91	WATER SUPPLY	3	ಕ್ರಿ
11	COOLING WATER SYSTEM	ಕ್ರಿ	ಕ್ರಿ
18	WASTE WATER TREATMENT	CA/RD	CA/30
19	GENERAL FACILITIES		•

CP = COMMERCIALLY PROVEN, CA = COMMERCIALLY AVAILABLE, RD = READY FOR DEMONSTRATION KEY:

I/O and interface requirements of the present module designs. These adjusted designs have been placed in the block flow diagram sequence and used as a basis for material balances, energy balances, and cost analyses. Table 3.3 presents the status of systems chosen.

3.2 System Descriptions

Many of the systems shown in the block flow diagrams are either the same for all designs or similar in description for each of the two designs except for size and cost considerations. This section defines each of these systems. More detailed descriptions of those systems specific to Lurgi and BGC/Lurgi are given in Sections 4.0 and 5.0, respectively.

3.2.1 System 11 - Coal Handling

The Coal Handling System provides for the unloading of coal delivered to the plant either by barge or truck, reclaiming the coal from storage, reducing the size of coal, and transporting the coal to Coal Preparation and Feeding, System 1.

Raw coal is unloaded from barges by the barge unloading subsystem which is designed to unload up to ten 1500 ton capacity barges per shift. The coal is unloaded at an average rate of 1200 tons per hour on a 5 day week basis. Coal is transferred by conveyor to a radial stacker which then produces a kidney shaped coal pile containing live and dead storage. Coal is reclaimed from live storage and conveyed to a Bradford breaker where it is reduced in size to 2"x0. Coal from trucks is unloaded into a chute from which it is conveyed to the Bradford breaker. Crushed coal from the Bradford breaker is transported to day storage silos from which it is transfered via vibrating belt conveyors to coal preparation and feeding, System 1, for further processing. Coal fines from the Bradford breaker are collected and sent to the gasification unit, System 2.

3.2.2 System 1 - Coal Preparation and Feeding

The Coal Preparation and Feeding System takes primary crushed coal from the coal handling system, grinds and dries the coal to its final size requirement, and stores the prepared coal in intermediate surge bins for transfer to the coal gasification system. Since each of the three gasifiers has different feed requirements this system is further described in the appropriate section.

3.2.3 System 2 - Coal Gasification

This system takes the prepared coal from System 2 and converts it into gaseous products and slag byproduct. The specific gasifier descriptions are given in the appropriate section.

3.2.4 System 3 - Initial Gas Cooling and Cleanup

This system takes hot gases from System 2 and reduces the temperature as appropriate for downstream processing, by quenching, steam generation and hear exchange. Products are sent to compression and condensate treating. Each design has different requirements which are described in the appropriate section.

3.2.5 System 7 - Compression

This system receives raw gas from the gas cooling section and compresses it prior to acid gas removal sufficient to meet a 600 psig final plant discharge pressure. Since each gas ifier has a different discharge pressure and gas composition this system is further described in the appropriate later section.

3.2.6 System 4 - Acid Gas Removal

The purpose of this system is to receive raw gas and remove sufficient acid gas to meet the plant product specifications of 200 ppmv sulfur and 285 BTU/SCF HHV.

Though the material and energy balances and total system cost for this system vary between the two designs presented here the system descriptions are otherwise identical

The Selexol Sulfur Removal Unit is a proprietary process licensed by Allied Chemical Co. The primary purpose of this system is to remove H_2S and COS from the product gas stream in order to obtain a sulfur concentration of not more than 200 ppmv. Secondarily, the process removes CO_2 , which increases the gas heating value.

The Selexol process removes sulfur compounds and carbon dioxide by means of countercurrent contact in an absorber. The physical absorption of the compounds to be removed is done using the dimethyl ether of polyethylene glycol as a solvent. Chilled solvent is used. Solvent chilling is accomplished by means of an absorption refrigeration system utilizing low pressure gasifier jacket steam.

CO₂ removal from the rich solvent leaving the absorber is accomplished by a series of successively lower pressure flashes. Flashed gas is recycled to the absorber inlet. Sulfur compounds are removed in the stripping column. The stripping column utilizes a steam heated reboiler. The stripper overheat gas, after cooling to about 120°F, flows to the Sulfur Recovery and Tail Gas Treatment Unit - System 5.

The sour gas from the Selexol unit is rich in ${\rm CO}_2$, and at slightly above atmospheric pressure.

3.2.7 System 5 - Sulfur Recovery and Tail Gas Treatment

The Sulfur Recovery and Tail Gas Treatment system receives H₂S and some COS and CO₂ from System 4 and converts the sulfur bearing gases into molten sulfur. The system is similar in both designs.

Acid gas from Acid Gas Removal, System 4, is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean H_2S acid gases. Typically in a Claus type sulfur plant, the acid gas is first passed through a knockout drum before entering the reaction furnace. The chemistry of the process involves converting the H_2S to elemental sulfur according to the following equation;

Any hydrocarbons in the acid gas are burned to ${\rm CO_2}$ and ${\rm H_2O}$.

The reactions are exothermic, and the heat liberated generates steam in the reaction furnace boiler and in the sulfur condenser. The sulfur from each condenser is drained to a recovery pit in Byproduct Processing, System 13, and the tail gas from the final condenser is fed to a Beavon tail gas treating unit where essentially complete removal of the remaining sulfur compounds is achieved before discharge to the atmosphere. The Beavon sulfur removal process is capable of reducing the sulfur content in the tail gas to less than 100 ppm. It is proprietary process. In this system hydrogenation and hydrolysis are used to convert essentially all sulfur compounds to hydrogen sulfide. This gas is then cooled, and passed into a contactor where the hydrogen sulfide is absorbed by the redox solution and oxidized to elemental sulfur. The reduced redox solution is reoxidized by contact with air and subsequently recirculated to the contactor. Elemental sulfur is removed in the air-blowing step as a froth which is pumped to a sulfur melter to be melted under pressure, separated from the redox solution, and transferred to Byproduct Processing, System The decanted redox solution is returned to the system.

The chemical reactions are:

Hydrogenation and Hydrolysis

3.2.8 System 8 - Process Solids Treatment

The purpose of this unit is to collect and dewater the various solids slurries, or sludges resultant from the Facility Operation for economical, environmentally acceptable disposal.

The processes involved in this system are nonproprietary and are supplied by various US vendors.

Ash and slag from the gasifier and gas cooling system, biological sludges, and solid wastes, from process condensate treatment are treated in this unit. The methods below are used to accommodate the solids treatment.

Gravity settlers separate dense solids from the waste streams.

Flot thickeners-clarifiers treat slurry from the gravity settlers. Thickener and coagulent aids are added to facilitate solid-liquid separation.

Rotary Drum Filters filter sludges from the process condensate treating systems and the float thickeners.

Recovered water is sent to the Process Condensate Treating System, and solids are conveyed to Final Solids Disposal.

3.2.9 System 9 - Incinerator

An analysis of the plant requirements led to the conclusion that no incinerator was required. Miscellaneous waste oils and tars from the Lurgi or BGC/Lurgi plants are disposed of in the tar burning boilers. Combustable solid waste from such systems as coal handling and feed preparation are disposed of via the solids disposal system.

3.2.10 System 10 - Instrumentation and Control

At the module, system, and subsystem level instrumentation and control functions are included as part of the system.

The purpose of this unit is to provide operational monitoring and control of the plant and supervisory master control of the module operations.

The system includes instrumentation to measure key characteristics of all major streams (mass flow rate, temperature, pressure, composition, density, viscosity, and others as appropriate); and to display these characteristics in a central location. A dedicated computer and alarm systems are provided to record stream characteristics and identify conditions that should be brought to the unit operators' attention. Telecommunications capabilities are provided to communicate with unit operators and general facilities (System 19). All major alarms will be redundant, one on the data acquisition system and one hard-wired alarm.

3.2.11 System 12 - Solid Waste Recycling/Disposal

The purpose of this unit is to store solid waste generated by facility operation during the facility life.

This is a non-proprietary system.

This system consists of a lined impounding pit sized to contain 20 years of solid waste. It includes the conveyor system to move the solids to the pit and a leachate recovery area to recover the pump leachate to the process condensate system.

3.2.12 System 13 - Byproduct Processing

This system receives byproduct sulfur from System 5 and ammonia from System 18 and provides products to sales. More specific descriptions are given with later sections of the report.

3.2.13 System 14 - Plant Power System

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

- Develop the necessary voltage stepdown arrangement for plant requirements
- · Distribute the necessary power to the plant equipment.

This is non-proprietary equipment and is supplied by several U.S. vendors.

TVA's incoming substation transformers receive power from its prevalent distributed voltage switching station and step down this voltage to a medium voltage to supply the plant electrical power requirement for motors, heaters, lighting, and other miscellaneous loads.

The Medium Voltage Electrical Distribution System is a secondary selective system (double ended supply) with several medium voltage buses. Each medium voltage bus receives power from its respective incoming substation transformer through an incoming breaker and supplies power to the medium voltage distribution system through the feeder breakers.

The Low Voltage Electrical Distribution System typically consists of multiple 480 V double ended load centers and 480 V motor control centers (MCC's) supplying the power to 480 V loads throughout the plant. Two load centers are interconnected through a normally open tie breaker. In the event of loss of one load center transformer or its feeder, the 480 V loads of the affected load center are fed by the second load center through the tie breaker.

Each load center consists of an incoming line section, load center transformer, and low voltage section with metal enclosed draw out power circuit breakers.

Load center transformers are air cooled, dry type, 150°F temperature rise, with delta connected primaries and wye connected secondaries. All load center feeder circuit breakers are 1600A frame and 50,000A RMS symmetrical interrupting capacity. The 480 V motor feeder breakers are electrically operated with instantaneous and long time trip units.

480 V MCC's consist of starters, feeder circuit breakers and control devices, assembled in a common structure with horizontal and vertical buses.

A 125 Volt DC System supplies control power for medium voltage and 480 V volt plant switchgear control, protective relaying and annuciation. The system also supplies power for emergency lighting.

3.2.14 System 15 - Steam Generation/Distribution

Steam Generation/Distribution provides a means for recovering waste heat from the plant process systems while providing process steam requirements. The system consists of steam generation (process and boilers), steam distribution, condensate recovery, and boiler feed water distribution. More specific systems are described in other sections.

3.2.15 System 16 - Water Supply

The Raw Water Treatment Unit is designed to provide treated and untreated water for the following facility water systems:

Fire Water Service Water Potable Water Cooling Water Boiler Feedwater

Raw water is pumped from the river to a Fire Water-Raw Water Storage tank.

The Raw Water-Firewater Storage Tank provides surge capacity for Water Treatment as well as storage capacity for firewater. During an emergency, firewater is pumped from the tank to the firewater header system. The firewater pumps are motor driven and have a diesel engine driven spare. The spare pump is equipped with automatic start-up capability in case of power failure.

The raw water is pumped from the Raw Water-Firewater Storage Tank to the Softener-Clarifier. Lime, alum, and polyelectrolyte from the Clarifier Bulk Chemical Storage and Feed System are added to the Softener-Clarifier, which is equipped with an internal flocculation mechanism. The alum and polyelectrolyte aid in the removal of suspended solids from the raw water. Lime is added during the clarification step to "cold soften" the raw water. Chlorine is added to the raw water to inhibit algae growth in the clarifier and sand filters and reduce organic contamination.

The underflow from the clarifier is a one wt percent sludge and is pumped to solids treatment for further processing.

The clarified and softened raw water from the Softener-Clarifier flows to the Self-Backwashing Sandfilters where additional suspended solids are removed. A pressure differential across the filter bed initiates the backwash cycle. The backwash flows by gravity to the Sandfilter Backwash Sump and is recycled to the Softener-Clarifier. The filtered water flows to the Filtered Water Storage Tank and is utilized as cooling tower make-up for the Process Cooling Tower as service water for general plant use, as feed to the Demineralizer Package, and as feed to the potable water system.

Water intended for potable services is chlorinated and again filtered to meet American Water Works Association (AWWA) standards and stored in a tank sized to hold a day's potable water requirements. The chlorine residual is maintained at 0.5-1.0 PPM free chlorine in the tank.

Filtered water intended as feed to the Demineralizer Package is injected with Sodium Sulfide to remove trace amounts of chlorine which adversely affect the Demineralizer resins and after filtered through activated carbon to remove any remaining organic contaminants and dissolve iron.

In the demineralizer, the mineral salts present in the water are removed by ion exchange. A two-step demineralization system, utilizing strong cation and strong anion exchangers in series, is provided. A degasifier following the strong cation and magnesium, which the anion exchangers remove anions such as chloride and sulfate. The strong anion exchanger also removes silica. The degasifier is provided to remove carbon dioxide and other dissolved gases.

A mixed bed polisher is provided to remove silica to 0.02 PPM and to polish returned turbine condensate for reuse.

The demineralized water and condensate is preheated to 220-225°F before being pumped into the boiler feedwater deaerators.

Acid, stored in the Acid Tank, is used to regenerate the cation exchangers. Caustic, stored in the Caustic Tank, is used to regenerate the anion exchangers.

Both the Acid and Caustic tanks have pumps and metering devices to control the flow of regenerants to the demineralizer package.

The Boiler Feedwater Deaerating heaters operate at 30 psig and 250° F. The Deaerators reduce the oxygen content of BFW to 0.005 cc/lites.

Hydrazine or sodium sulfite is injected into the storage compartment of the deaerators for chemical scavenging of any residual oxygen. Morpholine is injected into the suction of the boiler feed-water pumps to protect the condensate systems.

3.2.16 System 17 - Water Cooling

The purpose of this unit is to provide cooling water to the various process users in the facility.

This is a non-proprietary system and is supplied by several US vendors.

The cooling tower system includes the tower and fans, sidestream filters, circulating water pumps, cold water basis, blowdown system, chemical addition equipment, and distribution system.

Cooling water is pumped from the cold water basis, through the distribution system to the process heat exchangers where low-level, sensible heat is picked up, and back to the cooling tower. The cooling tower rejects low-level heat by evaporative cooling to air drawn through the cooling tower by the cooling tower fans.

 λ portion of the circulating water is passed through side stream filters to reduce loading to suspended solids, dirt and scale.

The dissolved solids level of the cooling water is maintained by a continuous blowdown stream to the process condensate system. Water level in the cooling tower basis is maintained by continuous make up of the clean water from the raw water treatment system.

The blowdown stream is passed through a blowdown treatment system to recover chromate ions via ion exchange or by chemical reduction to chromium hydroxide and is sent to waste treatment for disposal.

Chlorine is added to the cooling water on a routine periodic basis to prevent algae growth. Chemical algicides are added periodically to further eliminate algae growth. Sulfuric acid is added to control pH and zinc and chromate inhibitors are added to the cooling water for corrosion control. Occasionally a polyphosphate dispersant is added to enhance the action of the inhibitors.

3.2.17 System 18 - Waste Water Treatment

Waste Water Treatment receives all plant liquid effluent streams, recovers ammonia and hydrocarbon byproducts and treats the water for recycle. Zero aqueous discharge is designed into the plant. More specific descriptions are given in other sections.

3.2.18 System 19 - General Facilities

The purpose of this unit is to provide equipment or services to support the Gasification Facility at the facility level.

The equipment and Services provided in this unit are non-proprietary.

This unit is a general facility category and provides the following equipment and services:

- Administration Building
- Laboratories
- Change Rooms
- Warehouses
- Maintenance Buildings
- Operation Centers
- Security Offices
- Plant Air Facility
- Fire House
- Visitor Reception
- Plant Fencing
- · Plant Lighting
- Roads, Bridges
- Docking Facilities
- Interconnection Pipe Ways
 Fire Protection Network
- Flare Stacks and Headers
- Plant Instrument Air Compressors
- Environmental Monitoring
- Site Preparation.

4.0 LURGI FACILITY DEFINITION DESIGN

The Lurgi Facility Definition Design processes 20,000 TPD of Kentucky No. 9 coal in four identical modules. The basis of the design is the Lurgi moving bed coal gasifier. The configuration of each module is divided for purposes of engineering studies and cost analysis into a number of interconnected systems. Drawing Number 521-L-001 is a block flow diagram of the design. These systems are listed in Table 4.1. The systematic analysis which led to the choice of systems given in Section 3.0 applies to this design. In addition a selected number of trade studies specific to Lurgi based gasification were conducted in order to complete the choice of technologies for each system. Table 4.5 gives the material balance corresponding to the block flow diagram. Tables 4.6 and 4.7 give the plant energy balance. Table 4.8 presents the operating requirements.

4.1 Lurgi System Trade Studies

The system technology distinctly associated with the Lurgi or BGC Lurgi as compared to the other gasifier processes of concern in this study are tar/oil disposition, phenol recovery, and ammonia recovery. These process requirements arise directly from the operation of a fixed bed countercurrent reactor with its relatively cool coal feed entry section wherein coal is devolatized under conditions not sufficiently severe to cause gasification of the vaporized volatile material. The low temperature also promotes the formation of ammonia as compared to the molecular nitrogen formed in hotter gasification reactors.

4.1.1 Tar/Oil Disposition Trade

Alternatives considered for the disposition of recovered tars and oil are:

- a) burn in fired equipment and
- b) sell as a product.

Table 4.2 presents the considerations relative to each alternative. Based or these considerations, lack of dependable data on product quality/ quantity to be expected as a basis for sales and previous team experience with difficulty in finding suitable markets, the decision is to burn the tar/oil product in the boilers of the Steam Generation and Distribution System.

4.1.2 Phenol Recovery Trades

The alternatives considered for phenol recovery include:

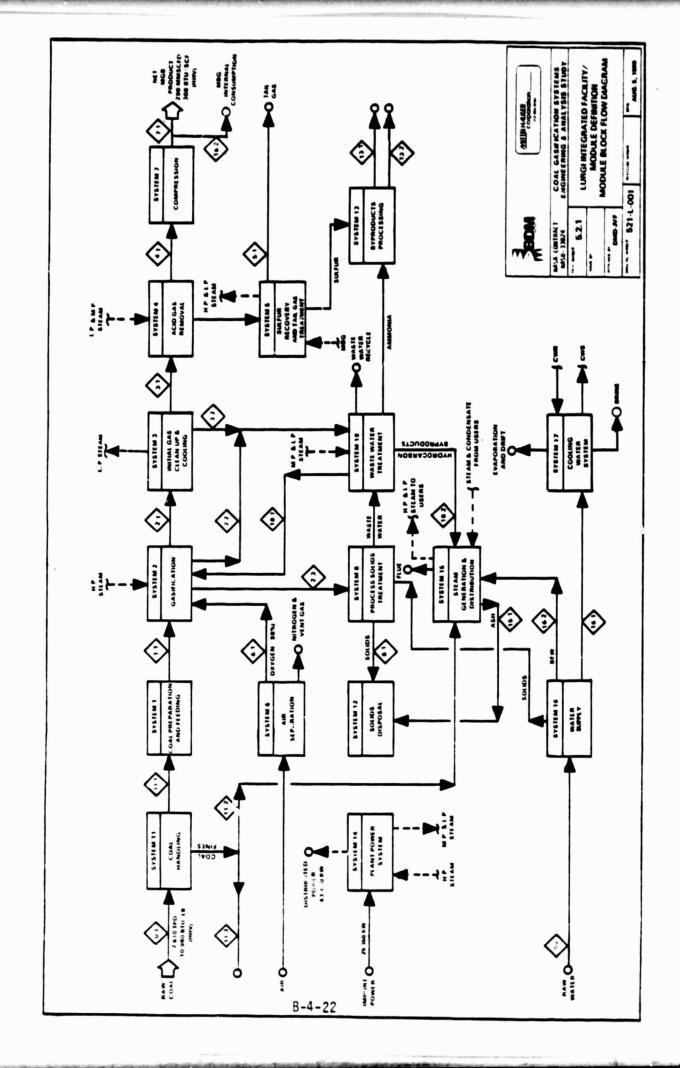
- a) do no recover (bio-oxidation)
- b) Phenosolvan process
- c) Chempro process.

TABLE 4.1. LIST OF SYSTEMS, LURGI PLANT

SYSTEM NO.	NUMBER OF O	COST UNITS PER FACILITY	SYSTEM DESCRIPTION
1	1	4	COAL PREPARATION & FEEDING
2	7	28	GASIFICATION
3	1	4	INITIAL GAS CLEANUP & COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	1	5	AIR SEPARATION
7	1	4	COMPRESSION
8	1	4	PROCESS SOLIDS TREATMENT
10	-	1	INSTRUMENTATION AND CONTROL
11	-	1	COAL HANDLING
12	-	1	SOLIDS DISPOSAL
13	1	4	BYPRODUCT PROCESSING
14	1	4	PLANT POWER SYSTEM
15	2	8	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKEUP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	-	1	GENERAL FACILITIES

TABLE 4.2 TAR/OIL DISPOSITION TRADES LURGI AND BGC/SLAGGER

ALTERNATIVE NO. DESCRIPTION	SELL AS BYPRODUCT	USE AS FUEL
MEETS PERFORMANCE CRITERIA	Yes	Yes
RELATIVE CAPITAL COST RELATIVE ENERGY CONSUMPTION PROPRIETARY COMMERCIALLY PROVEN	Low Lowest No	Higher Higher No Yes
APPLIED DEVELOPMENT NEEDS	_	-
COULD DELAY IMPLEMENTATION DATA AVAILABILITY	Yes	No
SYSTEM-LEVEL DESIGN	Yes	Yes
COST	Yes	Yes
SUBSYSTEM LEVEL DESIGN	Yes	No
COST	Yes	No
RELATIVE COMPLEXITY	Low	Moderate
RELATIVE OPERATING COST	-	-
POTENTIAL ENVIRONMENTAL PROBLEM		SO _x , NO _x
BYPRODUCT MARKETABILITY	Questionable	^- ^



The considerations applicable to these alternatives are presented in Table 4.3. Based on these considerations and the fact that:

- Phenol extraction is only needed with Lurgi based facilities and Lurgi generally prefers or insists on use of Phenosolvan
- Direct biological oxidation is very risky due to very high BOD concentration
- Phenosolvan and Chempro are very similar in overall process approach.

The decision is to use Phenosolvan to recover phenols and to burn them in the Steam Generation and Distribution System.

4.1.3 Ammonia Recovery Trades

The process alternatives considered for ammonia recovery include:

- a) Chevron WWT
- b) Phosam-W
- c) stripping only

The considerations applicable to these alternatives are presented in Table 4.1. Based on these considerations and the fact that

- Phosam-W has been found to be the economic choice from past direct comparisons when an ammonia producing gasifier is used
- Power costs are higher for the WWT process
- If ammonia is not recovered severe problems can occur in the sulfur recovery plant
- Phosam-W is proven on dirty gas similar to coal gasification products.

The decision is to use Phosam-W to recover ammonia from sour water.

4.2 System Descriptions

The systems shown in the block flow diagram are catagorized as similar in description to those given in Section 3.0 or as specific to the Lurgi based plant. Systems in the later catagory are:

- System 1 Coal Preparation and Feed
- System 2 Coal Gasification
- System 3 Initial Gas Cleaning and Cooling

TABLE 4.3
PHENOL RECOVERY TRADES
LURGI AND BGC/SLAGGER

ALTERNATIVE NO. DESCRIPTION	DO NOT RECOVER (BIO-OXIDATION)	PHENO- SOL VAN	CHEM- PRO
MEETS PERFORMANCE CRITERIA	Yes	Yes	Yes
RELATIVE CAPITAL COST RELATIVE ENERGY CONSUMPTION PROPRIETARY COMMERCIALLY PROVEN	Higher Higher No Yes	Lower Lower Yes Yes	Lower Lower Yes Yes
APPLIED DEVELOPMENT NEEDS	Some	No	No
COULD DELAY IMPLEMENTATION DATA AVAILABILITY	Yes	Yes	Yes
SYSTEM-LEVEL DESIGN	Yes	Yes	Yes
COST	Yes	Yes	Yes
SUBSYSTEM LEVEL DESIGN	Yes	No	No
COST	Yes	No	No
RELATIVE COMPLEXITY	Lower	Higher	Higher
BELATIVE OPERATING COST	Higher	Lower	Lower
POTENTIAL ENVIRONMENTAL PROBLEM	Higher	Lower	Lower
BYPRODUCT MARKETABILITY	No	Possible*	Possible*

^{*} At least will have fuel value

TABLE 4.4
AMMONIA RECOVERY TRADES

ALTERNATIVE NO. DESCRIPTION	CHEVRON WWT	PHOSAM -W	3 STRIPPING ONLY
MEETS PERFORMANCE CRITERIA	Yes	Yes	Yes
RELATIVE CAPITAL COST RELATIVE ENERGY CONSUMPTION PROPRIETARY COMMERCIALLY PROVEN	Highest Highest Yes Yes	Mid Mid Yes Yes	Lowest Lowest No Yes
APPLIED DEVELOPMENT NEEDS	Slight	Slight	Slight
COULD DELAY IMPLEMENTATION DATA AVAILABILITY	No	No	No
SYSTEM-LEVEL DESIGN	Yes	Yes	Yes
COST	Yes	Yes	Yes
SUBSYSTEM LEVEL DESIGN	No	No	Yes
COST	No	No	Yes
RELATIVE COMPLEXITY	Highest	Mid	Lowest
ANNUALIZED COST	-	Lowest	•
POTENTIAL ENVIRONMENTAL PROBLEM	None	None	High
BYPRODUCT MARKETABILITY	Yes	Yes	No

- System 6 Air Separation
- System 7 Compression
- System 15 Steam Generation/Distribution
- System 18 Waste Water Treatment

4.2.1 System 1 - Coal Preparation and Feeding

This system receives 2"x0 coal from System 11, Coal Handling and crushes it to maximum 1 inch in size. Coal in the size range 1"x28 mesh is conveyed to the Coal Gasification Section. Minus 28 mesh coal is recovered and used as supplemental fuel in the Steam Generation and Distribution System. Excess coal fines are sold as a plant byproduct.

4.2.2 Coal Gasification

The Lurgi gasifier, a dry ash, gravitating bed type, is commercially available from Lurgi Kohle and Mineraloetechnik.

The gasifier is essentially a refractory-lined, water-jacketed cylindrical shell operating at 30 atmospheres pressure. Coal is received from Coal Preparation into lock hoppers situated above the gasification reactors. Coal is fed to the gasifier by gravitational feed from the lock hoppers and spread over the top of the bed of coal. The bed of coal gravitates from top to bottom. The coal flows countercurrent to the gasification medium (oxygen and steam). The steam generated in the water jacket is used in gasification. Dry ash is removed continuously by a rotating grate into a semi-automatic ash lock.

The gas leaves the gasifier at a temperature of 650 degrees F. The gas is washed in a scrubbing cooler where it is cooled and water saturated. Traces of coal dust contained in the gas are removed via the action of heavy tar condensation on the particles in the scrubber. This mixture of tar-dust can be recycled to the gasifier. The scrubber is an integral part of the waste heat boiler system and the gas leaves the waste heat boiler at 399 degrees F. At this point, the gas consists primarily of CO_2 , CO, H_2 and CH_4 , and some H_2S . The proportion of these components depend on the type of coal and the operating conditions. The gas also contains tar, oil, light naphtha, other hydrocarbons, and sulfur compounds.

4.2.3 System 3 - Initial Gas Cleaning and Cooling

Raw gas from the Coal Gasification System enters the Initial Gas Cleaning and Cooling System from the waste heat boiler in the Coal Gasification Systems. Since this stream still contains ammonia, carbon dioxide, hydrogen sulfide, and oil and water vapor, special design considerations are required to prevent plugging and excessive fouling of cooling surfaces. This design including vertical tubes with tube walls washed with reinjected gas liquor is based on proprietary technology owned by Lurgi Kohle and Mineraloeltechnik, GmbH. Gas leaves this system at 100°F. Water effluent containing tar, oil, ammonia and acid gases is sent to Waste Water Treating.

4.2.4 Air Separation

This system is described in Section 3 with final compression of the oxygen to 515 psig.

4.2.5 System 7 - Compression

This system receives gas at 412 psig from the Acid Gas Removal System and boosts the pressure to 600 spig for plant discharge.

4.2.6 System 15 - Steam Generation and Distribution

The Steam Generation and Distribution System produces process steam from the gasifier jackets and high pressure steam from the waste heat boiler in the Coal Gasification System and medium and low pressure steam in the sulfur recovery system. Additional steam requirements are produced by high pressure steam boilers fired with tar, oil and phenolic liquids supplemented with coal fines from the Coal Preparation and Feed System. Process steam requirements are primarily Coal Gasification, Acid Gas Removal (regenerator reboiler and solvent chilling unit) and Waste Water Treating Systems. High pressure steam is let down through back pressure turbines to supply additional low and medium pressure steam for processing and miscellaneous plant heat requirements. A condensate recovery, deaerator, and boiler feed water distribution network complete this system.

4.2.7 System 18 - Waste Water Treating

The Waste Water Treatment System is designed for Tar/Oil separation, process condensate treatment, phenol recovery, ammonia recovery, and return of treated water to the plant. Solids removed in this system are put in the Solids Disposal System.

Tar/oil separation is accomplished in gravity settling tanks. The tar oil product is used as fuel in the Steam Generation and Distribution System. Phenolic compounds are recovered in a Phenosolvan unit. Phenolic compounds are to also be burned as fuel pending development of a makret. Ammonia is recovered in a Phosom-W unit and sent to the Byproduct Process System for sale.

Streams containing miscellaneous free and dissolved oil are treated in a gravity separator utilizing an emolsion breaker and heat to separate the oil water mixture.

Streams containing a high as pH are treated with sulfuric acid or lime as appropriate in equalization basins. Neutralized brines are evaporated with the concentrated solids goild solids disposal. Recovered treated water is used as makeup for the cooling tower and raw water supply.

TABLE 4.5
MATERIAL BALANCE
LURGI PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER Stream 1D				
,		\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\	\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\	77
	-	RAW COAL	EAN WATER	TER PEED
SYMBOL	FORMULA WEIGHT	LB-MOT.S LBS	LB-FOLS 1.19S	LB-MOLS LBS
ú	12.01	386.033.	•	251,635.
Ш,	2.016	27, 282.		17.925.
0,	32.0	8,769.		3,762.
ARGON N2	28.016	36.378.		21,902
S	32.06	73.514	*	767
-	10.00	110		
2	78.01			•
CARBON DIOXIDE CO2	44.01			
CE	16.042	-		
6.9	28.052	•		
Call	30.068			
CZH	42.07B	•		
CHE	44.094	•		
HYDROGEN SULFIDE HIS	34.076	•		
	60.075		-	4
	76.13	•	•	
1	90.09	•		
	30.008	•	-	-
N. T.	17.031	-		
CVANTOE HCK	27.026	•	•	
-	36.461		•	
		1	•	
'		, oot 00		059 05
		90, 788.		42000
		•		
		573,515.	•	3/8,817.
H,0	18.016	60.652.		00000
		634,167.		. 799, 914
GAS MOLECULAR WEIGHT				
TEMPERATURE, OF				
PSIA				
		-	2.083.	1

TABLE 4.5 (Continued)

MATERIAL BALANCE
LURGI PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

ARCON N2 ARCON N2 ARCON N2 C1 C2 C2 C2 C3 C4 C2 C4 C4 C5 C5 C6 C6 C7 C7 C7 C7 C7 C7 C7 C7	12.01 12.01 2.016 2.016 28.016 28.016 35.453 28.016 35.453 28.016 44.01 16.042 44.01 16.042 42.078 44.094 42.078 44.094 42.078 44.094 42.078 44.094 42.076 60.075 76.13 17.031 17.031 17.031	EDILER COAL LB-WOLLS LB-	29.518. 1.148	1-3 COAL FINES EXPORT LB-FOLS HIR 13	81,880. 5,787. 1,859. 7,715. 4,988. 4,988. 4,988. 4,988. 4,988. 1,218.	SULPIR RYPRODUCT LB-MOLS LB-	12 13 13 13 13 14 15 15 15 15 15 15 15 15 15 15 15 15 15
TUTAL, WET GAS MOLECULAR MEIGHT TEMPERATURE, OF PRESSURE, PSIA SCPH							

TABLE 4.5 (Continued)

MATERIAL BALANCE
LURGI PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER	2-41	\triangle	1-91	^	2-91	^
STREAM ID SYMBOL PORMULA WEIGHT	ANITOROUS APPONTA	APPONTA LPS	COOLING TOWER HAKE-UP	HAKE-UP	MOLLER FEED	FEED WATER
		,		•		,
EN H						-
0,		1				
EN & ARGON NZ		-				•
						-
BOLXCNO						
8						
2.5						
		,				
NE C'B'						
C 30°						'
B.55	-					•
soo						•
or cs,		•				
2						
OXIDE NO						
· III		2,604.				
CYANIDE HCM						
A CHLORIDE IIC1						,
NAPHTHA						•
TAR & OIL		•				
ASH .						-
OTHER SOLIDS		•				•
TAL, DRY		5,604.				
WATER H ₂ 0 18.016		,		•		.029,669
TOTAL, WET		•				•
GAS MOLECULAR WEIGHT						-
PRESSURE. PSIA	1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1	1				
SCFII		1	1 m m m			
				-		

. TABLE 4.5 (Continued)

MATERIAL BALANCE LURGI PROCESS INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER		(15-1)	⟨1-81 ⟩	(18-2)
STREAM 1D SYMBUL	FORMULA WEIGHT	BOLLER ASH TO DISPOSAL	RECYCLE GAS LIQUOR LB-POLS FR	HYDROCARBON BY-PRODUCTS TO BUTLE 14-HOLE IN THE
CARBON	12.01	•	•	
Z.	2.016			
DAYGEN CAPON NO	32.0		•	
	32.06			1
ar.	35.453			
w	28.01	•	1	
	14.01	• [
	16.042			
ETHYLENE C2B4	28.052		•	
	30.06	-		
PROPERE	44.094			•
N SULFIDE	34.076			
İ	60.075			
DE	76.13			
3	64.06	•	•	
OXIDE	30.008	•	1	
	17.031	•	5,738.	
HYDROGEN CYANTDE HCM	27.026	•		
N CHLORIDE	36.461		520.	
			4,124.	4,028.
TAR & OIL		- 11 701		21,631.
ER SOLIDS				
SUBTOTAL, DRY		11.491.	10 182	25, 550
MATER H ₂ O	18.016		669,711.	1,139.
TOTAL, WET		•	680.093.	26. 789
GAS MOLECULAR WEIGHT				
TEMPERATURE, OF			205.	
PRESSURE, PSIA				•
SCFII				-
G. H.			1,415.	5

TABLE 4.5 (Continued)

MATERIAL BALANCE
LURGI PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

			~
STREAM NUMBER	\(\frac{1-11}{1-11}\)	\(\frac{1}{2}\)	\(\sigma_{1}\)
		\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\	MARKY TAKEN VALUE
STREAM ID	FIER COAL	RAH CAS	1.P-MOLE 1.86
SYMBOL FORMULA WEIGHT	10-401	THE PERSON NAMED IN COLUMN 1	
CA.BON C 12.01	253,615.		•
EN NS	17,925.	30,030.	
0,	5,762.	3.937.	
EN 6 ARGON N2	15, 449		
	. 464	•	•
CHIOKING CO 28.01		152,441.	
9		436,816.	
Car.		7.376.	
		3,899.	
, n. 5			•
E C'B			•
c in		15,540.	
N SULFIDE BS	-	.099	
cos			•
CARBON DISULPIDE CS2 76.13			,
SO2		•	
OXIDE NO		5,604.	5,738.
AMMONTA LILETA	-	•	
HCM.		•	1707
HLORIDE		4.028.	4,124
TAR & OIL		•	17.00
ASH	.054.92		
OTHER SOLIDS	110 311	721.146	
	370,017	869.881.	117,009
WATER H20 18.016	.0.000		110 017
TOTAL. WET	416.667.	1,541,027.	
		366	199.
TEMPERATURE, OF		077	
PRESSURE, PSIA		31,518,044,	
SCFH	1		
3d5			

TABLE 4.5 (Continued)
MATERIAL BALANCE
LURGI PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER		<	<	
		\(\frac{6.3}{2}\)	\frac{1}{2}	<
STREAM ID			Sep Sends	(25)
TORNAS	PORMULA WEIGHT	128-1014 Las	1.0-101.6 Las	12-NOL6 1.86
CARBON	12.01	3 686		
2	2.016	. 904.7		
	32.0		.00,030	
EN 6 ARGON	28.016		1 011	
SULPUR	32.06		3,937.	
	15.451			
	28.01	1	177 651	
DIOXIDE	44.01		716 757	
	16.042		0101000	
200	28.052		2 376	
	30.068		1 800	
32	42.078			
PROPANE C'HE	44.094			
	34.076	1	075 51	
CARBONYL SULFIDE CÓS	60.075		. 099	
	76.13			
2	64.06			
OXIDE	30.008			
AMMONIA	17.031	•		707 5
CYANIDE	27.026			-
N CHLORIDE	36.461			
MAPHTHA.				4,028.
- Control		.059,850.		
SURTOTAL DEV		-		
		62,136.	111,514.	9.632
H20	18.016		1, 342.	924.588.
TOTAL, WET		62,136.	712.856.	000 710
GAS MOLECULAR WEIGHT			20.68	77777
TEMPERATURE, OF		750.	.001	277
PRESSURE, PSIA			430.	
SCFH			13,081,479.	
CPM				The same of the sa



TABLE 4.5 (Continued)

MATERIAL BALANCE LURGI PROCESS INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER		\rightarrow{\frac{1}{2}}	\$	\frac{1}{2}
SYMBOL FORM	FORMULA WEIGHT	COMPRESSION AND INTERNAL USE LB-MOLS HR HR	ACID GAS LB-MOLS HR HR	SCL. IDS TO DISPOSAL LE - MCL.S ER HR
CARBON C HYDROGEN B2	12.01	30,030,		2,486
NITROGEN 6 ARGON N2	28.016	7.917.		
CHLORINE	35.453			•
2	28.01	152,441		,
METHANE CH.	16.042	45,815.	92,170.	
	28.052	2,376.		
	30.068	7,077		
PROPYLENE C386	44 094		•	•
N SULPIDE	34.076	. 19	07.7.31	
	60.075	. Orox	360.	
3	76.13	•	•	
SULFUR DIOXIDE SO	30.00	,		
	17.031	•	-	
HYDROGEN CYANIDE HCM	27.026	•		
	36.461		-	
NAPITHA				
TAR 6 01L				059,85
ASH			-	120.
OTHER SOLIDS		603.505.	108.009.	62, 256.
TAL, DRY	210 01	312.	1,841.	6,917
WATER 120	877.81	603.817.	111,850.	69.173.
TOTAL, MET		18.95	40.41	
TEMPERATURE OF		57.	120	
PRESSURE, PSIA		427.	1.050.380.	
SCFII		12,030,333.		

TABLE 4.5 (Continued)
MATERIAL BALANCE
LURGI PROCESS
INTEGRATED FACILITY/MODULE DEFINITIOA

Note	STYMOL FORWILA WEIGHT HIS PRODUCT LIST HE PROTESS ONWELLS HE HE HE HE PROTESS ONWELLS HE HE HE HE HE PROTESS ONWELLS HE HE HE HE HE HE HE PROTESS ONWELLS HE HE HE HE HE HE HE PROTESS ONWELLS HE HE HE HE HE HE HE PROTESS ONWELLS HE HE HE HE HE HE PROTESS ONWELLS HE HE HE HE PROTESS ONWELLS HE HE HE PROTESS ON THE HE P	STREAM NUMBER		\rightarrow{\frac{7}{2}}	\(\frac{1-3}{2}\)	\(\frac{1}{2}\)	
Activity Activity	Name			NBG PRODUCT	SS OXIG	TAIL GAS LB-NOLS HR	E3
1.2 1.2	12.014 1		12.01	•].
1.00 1.00	1.00 1.00	EN	2.016	29.968.	111 111		11
Name 10 10 10 10 10 10 10 1	NAME State	FN & ABCOM	12.0	9 0 0 1	3.068		m
Name Columb Col	Name Col. 15, 15.2 15, 15		12.04	1,250			
NAME COLUMN COL	No. No.		35.453				
HERE CONTROL TO THE C	No. No.		28.01	152,121			93.5.5.
The control of contr	Mark Cold 16.042 45.22. 45.22. 46.042 45.22. 46.042 45.042 45.042 45.042 45.042 46.04		14.01	163.892.			
C	Color Colo		16.042	45.720.	-		
The color of the	Name		20.02	2,371.			
No. 1.0	Mark		30.06	1,892.			
See Sulfide 150 14.026 14.026 14.026 14.026 14.026 14.026 14.026 14.026 14.026 14.026 14.026 12.026	SCHIEF 150 14.016 150 14.016 150 14.016 150		110.25				-
National Color	NYL, SULFIDE CÓS 60.075 NH LISTALLIDE CÓS 61.05 NH DISULPIDE CS 76.13 NH DISULPIDE CS 76.13 NH DISULPIDE CS 76.13 NH LISTAL DISULPIDE HCT 1.01 NH LISTAL DEV CANIDE HCT 1.026 NGEN CYANIDE N	SULFIDE	14 076	.19			
NU DISULPIDE CS 76.13 NU DISULPIDE CS 64.06 NU DISULPIDE CS 10.00 NU DISULPIDE CS 10.00 NU DISULPIDE CS 10.00 NU DISULPIDE CONTROL CONTRO	NUM DISJULPIDE CS 76.13 NUM DISJULPIDE CS 64.06 NUM O 20.000 NUM DISJULPIDE CS 64.06 NUM DISJULPIDE CS 50.2 64.06 NUM DIS		60.075	300.		-	
N DIOXIDE SO	No continue So		76.13				
104 OXTOR NO 10.000 105 OXTOR NO 11.001 11.001 11.01	UNS OXIDE NO. 20.000 UNS OXIDE NO. 21.021 CEN CYMIDE HIT 21.021 CEN CYMIDE HIT 21.021 CEN CYMIDE HIT 21.021 SOLIDS SOLID		64.06				
12.031 12.031 12.031 12.031 12.032 12.035 1	17.031 17.031 17.031 17.031 17.031 17.031 17.031 17.035 1		30.00	•			
SOLIDS	SOLIDS		17.931	•			
121.6.3 122.729 122.729 122.729 122.729 122.5.53 122.5.53 122.5.53 123	1.24 1.24		27.026	•			
# SOLIDS # SOLIDS # SOLIDS # SOLIDS # SOLIDS # March PRY H20 18.016 602.259 174.229 125.55 2	# SOLIDS # SOLIDS # SOLIDS # AD 18.016	N CHLORIDE	36.461	•			
R SOLIDS NELL DRY H2O 124, 219 NET NET NET OLECULAR MEIGHT STATURE, OF SATURE, OF SATURE, PSIA 1,261,952, 1,264,591 1,264,591	R SOLIDS TYAL, DRY H2O 10.016 BA BA BA BA BA BA BA BA BA BA			•			
R SOLIDS TTAL, DRY H 20	SOLIDS 124,219 124,2	210		•	-		
TYAL, DRY H2O 10.016 BA ABL OURCULAR WEIGHT 174,729 17.57 18	TYAL, DRY R R R R R R R R R R R R R R R R R R						-
H ₂ O 18.016 602.354 31.92 37.57 3	R H ₂ O 18.016 602,259. WET 84. SOLECULAR WEIGHT 10.02 SRATURE, OF 515. SURE, PSIA 1.268,591.	SUBTOTAL. DRY			176 279		121.6.3
NET 11.92 17.51	OLECULAR MEIGHT 11.92 17.57 OLECULAR MEIGHT 100 SATURE, OF 515. SIS. L.268.591		18.016	602,259 84		1	1.872
OLECULAR MEIGHT 31.92 SRATURE, OF 51A 2.071.952.	OLECULAR MEIGHT 11.92 SATURE, OF 515. SURE, PSIA 2.077.952.			176 107			125,5 3.
SATURE, OF 302. SURE, PSIA 2.07.932.	SATURE, OF 102. SURE, PSIA 2,071,952.	SAS MOLECULAR WEIGHT		7767 700	31.92	15.11	
SURE, PSIA 2,071,952.	SURE, PSIA 315.	FEMPERATURE, OF			302.	80	
2,011,922	2,011,952	PRESSURE, PSIA			515.	Amblent 1	
The same of the sa	R-15	SCFII			2,011,952.	1.688.2	

TABLE 4.6

EMERCY BALANCE, 10⁶ BTU PER HOUR

LUNG! HOBOLLE

TASK 5.2.1 INTEGRATED FACILITY/MOBULE DEFINITION

	ž	SK 5.2.1 INTEGE	ATED FACILITY/P	TASK 5.2.1 INTEGRATED PACILITY/MODULE DEFINITION	
ENTHALPY BALANCE INLETS	PHASE	МИ	LATENT	SENSIBLE	TOTA
STREAM COAL ALT TO ALT SEP'N RAY MATER	Solld Gas Liquid	2486	. • .		X
SUBTUTAL IN					3
	e				
STREAM MBG	Solid	82.K	. 1	. ~	7278
Tail Gas	•	, 5		~	• ;
Air Sep'n Vent	5. 14. 5. 14.	٠,		.~;	;~;
Flue Gases	•	, 3	9	2	<u> </u>
MADIENALE	Liquid	۲,	ı ış	*	2
SUBTOTAL OUT					1604
NET IMLET-OUTLET					140
HEAT AND SHAFT WORK FIRETHIC POMER					901
SHAFT WORK			576-	-328	-130
HEAT REJECTION HEAT INPUT (STEAN)			•		7
RADIATION/FUGITIVE LOSS SUBTOTAL NET VALUE					-1236
TOTAL ABSOLUTE VALUE					
BALANCE: MET ENTIALPHY + NET HEAT & SHAFT WORK	AFT WORK	1403 - 1236	E 100E - 13.5E	3.51	

BASIS: 60°F; LIQUID WATER

TABLE 4.7 CONVERSION EFFICIENCY LURGI PROCESS 20,000 TPD FACILITY

PERCENT		53.5%	53.4% 74.48%	20.89
10 ⁶ BTU/HR 27,852 100	14,912 5,908		%00	
INPUTS 1. COAL TO FACILITY 2. FLECTRIC POWER TO FACILITY	OUTPUTS 3. MBG FROM FACILITY 4. COAL FINES FROM FACILITY	$\frac{\text{EFFICIENCY}}{\text{COAL-TO-MBG}} (3 + 1) \times 100\%$	OVERALL PRODUCT EFFICIENCY $(3) + (1) + (2) \times 100\%$ OVERALL FACILITY EFFICIENCY $(3) + (4) + (1) + (2) \times 100\%$	GASIFICATION EFFICIENCY (3) \div (1) \div (2) $-$ (4)

TABLE 4.8. OPERATING REQUIREMENTS FOR EXPECTED OPERATIONS LURGI PROCESS - PER MODULE

TPY at 100% Operation TPY at 100% Operation \$/YR at 100% Operation kWh/HR at 100% Operation
Man-Hours/YR
Man-Hours/YR
Factored as 15% of Operating Labor Cost
Factored at 1.6% of Total Depreciable Direct Investment
Factored at 2.4% of Total Depreciable Direct Investment

5.0 BGC/LURGI FACILITY DEFINITION DESIGN

The BGC/Lurgi Facility Definition Design processes 20,000 TPD of Kentucky No. 9 coal in four identical modules. The basis of the design is the BGC/Lurgi moving bed coal gasifier with slagging bottom. The configuration of each module is divided for purposes of engineering studies and cost analysis into a number of interconnected systems. Drawing Number 521-BGC-001 is a block flow diagram of the design. These systems are listed in Table 5.1. The systematic analysis which led to the choice of systems given in Section 3.0 applies to this design. In addition a selected number of trade studies specific to Lurgi based gasification were conducted in order to complete the choice of technologies for each system. Table 5.2 gives the material balance corresponding to the block flow diagram. Tables 5.3 and 5.4 give the plant energy balance. Table 5.5 contains an operation requirements summary.

5.1 BGC/Lurgi System Trade Studies

These studies are identical to those described in Section 4.1 for the Lurgi based facility.

5.2 System Descriptions

The systems shown in the block flow diagram are catagorized as similar in description to those given in Section 3.0 or as specific to the BGC/Lurgi based plant. Systems in the later catagory are:

- System 1 Coal Preparation and Feed
- System 2 Coal Gasification
- System 3 Initial Gas Cleaning and Cooling
- System 6 Air Separation
- System 7 Compression
- System 15 Steam Generation/Distribution
- System 18 Waste Water Treatment

5.2.1 System 1 - Coal Preparation and Feeding

This system receives 2"x0 coal from System 11, coal handling and crushes it to maximum 1 inch in size. Coal in the size range 1"x28 mesh is conveyed to the Coal Gasification Section. Minus 28 mesh coal is recovered and used as supplemental fuel in the Steam Generation and Distribution System. Excess coal fines are sold as a plant byproduct.

There is reason to believe that the fines cut off point for the BGC/Lurgi gasifier could be less than 28 mesh due to the lower gas velocity at a given coal feed rate (less steam). However due to lack of any other data this design is based on a minus 28 mesh minimum coal size.

TABLE 5.1 SYSTEM DEFINITION BGC/LURGI PROCESS

	SYSTEM DESCRIPTION	COAL PREPARATION & FEEDING	GASIFICATION	INITIAL GAS CLEANUP & COOLING	ACID GAS REMOVAL	SULFUR RECOVERY	AIR SEPARATION	COMPRESSION	PROCESS SOLIDS TREATMENT	COAL HANDLING	SOLIDS DISPOSAL	BYPRODUCT PROCESSING	PLANT POWER SYSTEM	STEAM GENERATION/DISTRIBUTION	RAW WATER MAKEUP	COOLING WATER SYSTEM	WASTE WATER TREATMENT	GENERAL FACILITIES
NUMBER OF COST UNITS	PER FACILITY	4	12	4	4	2	2	4	4	_	٧	4	4	80	4	4	4	_
NUMBER OF	PER MODULE	-	3	_	_	_	_	_	_	-	_	-	_	2	_	_	_	-
	SYSTEM NO.	-	2	က	4	2	9	7	80	=	12	13	14	15	91	17	91	19

5.2.2 Coal Gasification

This unit converts coal into medium heating value crude synthesis gas by partial oxidation in the presence of steam. The reaction takes place during countercurrent flow in a moving bed. The crude gas leaving the gasifier is scrubbed, quenched and saturated by gas liquor to remove coal dust and heavy tar. This multi-phase steam enters a waste heat exchanger for cooling and further condensation of heavier hydrocarbons prior to further processing.

This design is based on proprietary technology held by Lurgi Kohle and Mineraloeltechnik, GmbH and British Gas Corporation.

In the process the coal and flux, entering the top of the gasifier, descends in a moving bed in countercurrent flow to the steam, oxygen and produced gas. While traveling from the top to the bottom of the gasifier, the coal is dried, devolatilized, and gasified. The heat required for these three steps is supplied by the exothermic reaction between the carbon in the coal and the oxygen in the bottom of the gasifier. Flux is added to form a low melting temperature eutectic.

As the produced gas passes through the coal bed, its final composition is determined by the following:

- Exothermic and endothermic reactions occurring simultaneously in the gasification zone,
- Formation of hydrocarbons, phenols, fatty acids, and minor organic compounds in the devolatilization zone, and
- Evaporation of coal moisture in the drying zone.

After leaving the gasifier the raw gas is scrubbed and cooled in the wash cooler scrubbers and the waste heat exchanger.

The coal and flux are fed to the coal bunkers by a belt conveyer system. The feed chutes at the bottom of the coal bunkers control the flow of coal into the coal locks. Each gasifier has two coal locks that operate automatically on a cyclic basis. There coal locks are pressurized with an inert gas and feed alternately the coal surge vessel.

The gas released during depressurization and the gas displaced during filling of the coal locks contains a small amount of coal dust. The dust is removed prior to disposal of the streams.

In the bottom of the gasifier the coal ash melts as a eutectic with the added flux to form slag. The molten slag collects at the bottom and is tapped intermittently through a tap hole into the quench vessel. In the quench vessel the slag granulates immediately upon contact with the quench water. The granulated slag falls into the slag hopper and is dumped intermittently to the Solids Treatment System.

The raw gas leaving the Gasifier requires immediate treatment to remove as many impurities as possible. The initial treatment is provided in wash coolers operating in parallel. In the wash coolers gas liquor is injected to quench and saturate the raw gas. A multi-phase stream containing raw gas, condensed hydrocarbons, dust, stream, and water flows from the wash coolers to the waste heat exchanger for additional treatment. The multi-phase flow enters the waste heat exchanger above the gas liquor level. The gas and liquor are separated, and the gas is cooled by producing low pressure (35 psig) steam. The gas liquor condensed from the raw gas is collected in a sump. A portion of the dusty gas liquor is utilized as quench water for the wash coolers with the remaining liquor being sent to the Waste Water Treating System.

5.2.3 System 3 - Initial Gas Cleaning and Cooling

Raw gas from the Coal Gasification System enters the Initial Gas Cleaning and Cooling System from the waste heat boiler in the Coal Gasification Systems. Since this stream still contains ammonia, carbon dioxide, hydrogen sulfide, and oil and water vapor, special design considerations are required to prevent plugging and excessive fouling of cooling surfaces. This design including vertical tubes with tube walls washed with reinjected gas liquor is based on proprietary technology owned by Lurgi Kohle and Mineraloeltechnik, GmbH. Gas leaves this system at 100°F. Water effluents containing tar, oil, ammonia and acid gases is sent to Waste Water Treating.

5.2.4 Air Separation

This system is described in Section 3 with final compression of the oxygen to 490 psig.

5.2.5 System 7 - Compression

This system receives gas at 412 psig from the Acid Gas Removal System and boosts the pressure to 600 spig for plant discharge.

5.2.6 System 15 - Steam Generation and Distribution

The Steam Generation and Distribution System produces process steam from the gasifier jackets and high pressure steam from the waste heat boiler in the Coal Gasification System and medium and low pressure steam in the sulfur recovery system. Additional steam requirements are produced by high pressure steam boilers fired with tar, oil and phenolic liquids supplemented with coal fines from the Coal Preparation and Feed System. Process steam requirements are primarily Coal Gasification, Acid Gas Removal (regenerator reboiler and solvent chilling unit) and Waste Water Treating Systems. High pressure steam is let down through back pressure turbines to supply additional low and medium pressure steam for processing and miscellaneous plant heat requirements. A condensate recovery, deaerator, and boiler feed water distribution network complete this system.

5.2.7 System 18 - Waste Water Treating

The Waste Water Treatment System is designed for Tar/Oil separation, process condensate treatment, phenol recovery, ammonia recovery, and return of treated water to the plant. Solids removed in this system are put in the Solids Disposal System.

Tar/oil separation is accomplished in gravity settling tanks. The tar oil product is used as fuel in the Steam Generation and Distribution System. Phenolic compounds are recovered in a Phenosolvan unit. Phenolic compounds are to also be burned as fuel pending development of a makret. Ammonia is recovered in a Phosom-W unit and sent to the Byproduct Process System for sale.

Streams containing miscellaneous free and dissolved oil are treated in a gravity separator utilizing an emolsion breaker and heat to separate the oil water mixture.

Streams containing a high or low pH are treated with sulfuric acid or lime as appropriate in equalization basins. Neutralized brines are evaporated with the concentrated solids going to solids disposal. Recovered treated water is used as makeup for the cooling tower and raw water supply.

TABLE 5.2
MATERIAL BALANCE
LURGI/BGC PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

		<	<	<
STREAM NUMBER		1-0	0-1	(II-1)
STREAM ID		RAW COAL FEED	O WATER TRE	PREPARAT
HAS.	SYMBOL FORMULA WEIGHT	LB-MOLS LBS	LB-MOLS LBS	LB-MOLS LBS
CARBON	12.01	284 013		253 635
EN	2.016	27.282.		17,925
	32.0	971.91		23,902.
NITROGEN & ARGON N	28.016	8.770.	•	5,762
SULFUR	32.06	21.515.	•	15,450
CARBON MONOXIDE	28.01	192		
				•
METHANE CH.			•	
			1	
	30.068			
22		,		
	60.075	•		-
DE.				
OXIDE				
		•		
	27,026	,		
HYDROGEN CHLORIDE HC1		,	•	-
TAR & OIL				
		90,787.		.059,650.
OTHER SOLIDS			•	
SUBTOTAL, DRY		513,517.	t	376,818.
WATER H ₂ O	18.016	.60,652.	243,316.	39,850.
TOTAL, WET		634,169.	,	416,668.
GAS MOLECULAR WEIGHT		•	•	
TEMPERATURE, OF		į	•	
PRESSURE, PSIA				
SCFII			•	-
Erm.		•	486.	

TABLE 5.2 (Continued)

MATERIAL BALANCE
LURGI/BGC PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER			\(\frac{\sigma_{-11}}{\sigma_{-11}}\)		
STREAM ID	SYMBOL	FORMULA WEIGHT	COAL FINES TO STEAM CENERATION LB-MOLS HR HR	COAL FINES EXPORT LB-110LS LRS HR	MAKE-UP MATER TO COOLINE WATER STSTEM
CARBON	J	12.01	15.873.	765 911	
HYDROGEN	no no	2.016	1,122.	6,235.	
NITROGEN & ARGON	22	28.016	1,496.	10,981.	
	25	32.06		7,047	
CHLORINE	3	35.453	31.	227	
CARBON MONOXIDE	00	28.01			
CARBON DIOXIDE	603	14.01	•		
METHANE	City	16.042	,		
ETHYLENE	6.1	28.052			
ETHANE .	Call	30.068		,	
PROPYLENE	Cille	42.078			
PROPANE	CHES	14.094			
_	B.55	34.076			
CARBONY SULFIDE	cos	60.035		•	
CARBON DISULFIDE	CES	76.13		•	
SULFUR DIOXIDE	505	90.199		•	
NITROUS OXIDE	NO	30.00		•	
AMMONIA	Ī	17.031			-
	IICN	27.026			
HYDROGEN CHLORIDE	<u>=</u>	36.461	•		
TAP COLL					
ASH			1 233		
OTHER SOLIDS	-		1777	17.404.	
SUBTOTAL, DRY			23,583.	173.116.	
WATER	0 H	18.016	2,494.	8,308.	204 234
TOTAL, WET			26,077.	191,424.	
GAS MOLECULAR WEIG	=		1		
TEMPERATURE, OF					09
PRESSURE, PSIA					
SCFH					
The state of the s					

TABLE 5.2 (Continued)

MATERIAL BALANCE LURGI/BGC PROCESS INTEGRATED FACILITY/MODULE DEFINITION

18-1 18						3,611.		919.
MAKEUP WATER TO STEAM GENERALION BOILER ASH TO SOLIDS DISPOSAN LB-MOLS LB-MOLS HR HR HR							68,171.	.09
STREAM NUMBER STREAM 1D SYMBOL FORMULA WEIGHT	CARBON C 2.016 HYDROGEN B2 2.016 DAYGEN D2 32.0	PNOXIDE	2 - Z	SULF IDE COS SULF IDE CS SULF IDE CS	NO RING	92	SUBTOTAL, DRY WATER H20 18.016	GAS MULECULAR WEIGHT TEMPERATURE, OF PRESSURE, PSIA

TABLE 5.2 (Continued)
MATERIAL BALANCE
LURGI/BGC PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

	-					The second secon
2000			<		<	<
SINCAN NUMBER			(18-2)		\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\	\\-\\\\
STREAM ID			EAN CENE	RATION	RYPZODUCE ANAXBOUS ADSONEA	SIZED COM TO GASIFIER
	SYMBOL	PORMULA WEIGHT	LB-MOLS L	LBS	LB-10LS LBS	12-HOLE 138
CARBON	u	12.01				
HYDROGEN	B,	2.016		,		17 925
	0,0	12.0				23.902
NITROGEN & ARGON	N	28.016		,		5.762.
SULPUR	25	32.06			-	15.450.
CHLORINE	13	35.453		,		767
CARBON MONOXIDE	00	28.01		,	•	•
CARBON DIOXIDE	ó	44.01		-	1	
METHANE	CH.	16.042		ı		
ETHYLENE	C.B.	28.052				
ETHANE .	CÉB	30.068			-	
PROPYLENE	CHE	42.078				•
PROPANE	C 386	44.094			-	•
HYDROGEN SULFIDE	B .5°	34.076			-	
CARBONYL SULFIDE	sos	60.075		-		•
CARBON DISULFIDE	CS,	76.13		1		-
SULFUR DIOXIDE	505	64.06			-	-
NITROUS OXIDE	NO	30.008			-	•
AMMONIA	, HM	17.031		1	5.604.	•
	HCH	27.026		,	•	•
HYDROGEN CHLORIDE	IIC]	36.461		ı		
NAPIITIIA	,		7	4,028.		
TAR 6 OIL			21	21,631.		•
ASH				-		59 650
OTHER SOLIDS				-	•	
SUBTOTAL, DRY			25	25.659.	5.604.	376. AIA.
WATER	H ₂ 0	18.016		1,139.		19,850
TOTAL, WET			56	26, 798.		416.668.
GAS MOLECULAR WEIGHT	GHT					•
TEMPERATURE, OF						
PRESSURE, PSIA					-	
SCFH						
EP.				51.		

TABLE 5.2 (Continued)
MATERIAL BALANCE
LURGI/BGC PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

				<
		<	<	\(\frac{1}{2}\)
STREAM NUMBER			TARRY GAS 1-1	١
STREAM ID		RAW CAS TO COOLING	ATER TRE	PROCESS SOLIDS TREATMENT
SYMBOL	FORMULA WEIGHT	LB-MOLS LBS	LB-NOLS LBS	
			1	1,217.
u	2 016	15,220.	•	
D2	12.0	,		
	28.016	1,898.		
57	32.06	. .		
T5 C1	15.451	1.00 077		
8	28.01	38.403.		
CARBON DIOXIDE CO2	10.81	36.738.		•
100	16.042	1,904.		
5.00	20.07	3.005		
C2H6	10.00		•	
S C S	10.77		•	
-	14.036	15,377.		
HYDROGEN SULPIDE H25	34.076	943.	,	•
-	60.00			-
- 1	19.13		•	
2	64.00			
OXIDE	30.00	3,604.		
	17.031		•	•
CYANIDE	870.17		503.	
HYDROGEN CHLORIDE HCL	78. 387	4,028.	•	
			21,631.	
TAR 6 OIL		,	•	27.920.
		•		278 07
OTHER BOLLDS		565,143.	17,139	
IAL, DRI	18 016	95,062.	403,763	
WATER		660,205.	427,90	. 798,09
TOTAL, WET		30.1	,	
3		299	200	7,800.
TEMPERATURE, OF		977		
PRESSURE, PSIA		12 464, 136.		
SCFH			937.	
NG5				

TABLE 5.2 (Continued)
MATERIAL BALANCE
LURG!/BGC PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

STREAM NUMBER		000 th	(1) 6.45	MG. PRODUCT 4-1 645 10
STREAM ID		ID CAS		COMPRESSION AND INTERNAL USE
3 TOBMAS	FORMULA WEIGHT	TB-XOLS LaS	970 970M-97	11 TON-61
CARBON	12.01		•	•
HYDROGEN	2.016	15.220.	•	13.214.
	32.0	•		•
EN 6 ARGON	28.016	3,498.	-	3,696.
	32.06	•		
	15.453	•		
	28.61	440.023.	•	439.242.
CARBON DIOXIDE CO.	44.01	30,403.	•	20.632
TETHANE CB.	16.042	36,738.	•	36,730.
	28.052	1,904.		1.904.
	30.068	3,005.	•	3,005.
NE	42.07B		•	•
	44.094		•	
	34.076	11.311.	•	. 20.
	60.075	943.	The second secon	.193.
CARBON DISULPIDE CS.	76.13	•	•	•
SULFUR DIOXIDE SO.	64.06	•		
	30.00	•	•	
A.MONTA NH.	17.031	•	3,604.	•
CYANIDE	27.026	•	•	
HYDROGEN CHLORIDE HC1	36.461	•	•	•
NAPHTHA		•	4,028.	
TAR & OIL				•
ASH		•	•	•
-		-		•
TAL, DRY		555.511.	9,632.	530,896.
MATER	18.016	1.050.	145,902.	25.
TOTAL, WET		356,561.	155,534.	521,153.
GAS MOLECULAR WEIGHT		20.42		19.61
-		100.	•	- 23
PRESSURE, PSIA		430.	•	425.
100		10, 342,893.		9,965,973.
- Ara				

TABLE 5.2 (Continued)
MATERIAL BALANCE
LURGI/BGC PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

		AGR (4-2) OFF-GAS	\(\frac{1}{2}\)	\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\\
STREAM ID	FORMULA WEIGHT		SQL IDS TO FINAL DISPOSAL	10 1000 10 10 10 10 10 10 10 10 10 10 10
CARBON	19 61	•	1313	•
I.Y DROGEN	2.016	•		15.181.
	12.0	•	•	•
TTROGEN & ARGON MA	28.016	•	•	3,889.
	32.06	,		-
CHLORINE	15.453	•		
CARBON MONOXIDE CO	28.01	782.		
DIOXIDE	44.01	11.711.	-	•
CHIANE	16.042	•	-	
NE	20.052	•		•
	30.068	•	•	
NE.	42.078		•	
	14.094	•		•
	34.076	15,307.		
CARBONYL SULTIDE CÓS	60.075	750.	•	
	76.13	•	•	
	64.06	•		•
OXIDE	30.00	•		
	17.031	•	•	•
HYDROGEN CYANIDE NCM	27.026	•		•
N CHLORIDE	36.461	•	•	•
NAPRTHA		•	•	
TAR & OTL		•		•
ASH		•	.069,650	•
		•	Š	
SUBTOTAL, DRY		. 919. M	. 160, 897.	519.053.
SATER H30	18.016	1,230.	6,766.	70.
TOTAL, WET		35,646.	67,663.	319.921.
GAS MOLECULAR WEIGHT		37.15	•	19.61
		130.	-	120. (MAX.)
PRESSURE, PSIA		24.	-	615. (MIR.)
SCPH		366,142.	•	9.962.103.

TABLE 5.2 (Continued)
MATERIAL BALANCE
LURGI/BGC PROCESS
INTEGRATED FACILITY/MODULE DEFINITION

SYMBOL FORMULA WEIGHT 18-10			<	<	<
SYMBOL FORMULA WEIGHT TO GASHIRE TAIL GAS THE SALEN SALEN LESS LESS LESS LESS LESS LESS LESS L			1-9	(1)	(<u>r</u>)
SYMBOL FORMULA WEIGHT LIB-				Sut Tile	IV-PRODUCT SULFUR
12.01 12.0	SYMBO		11	11	19-407-61
1	t.i	12.01			
12.0 169.519, 169.519, 169.519, 169.519, 160.51, 160	, i	2.016			
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B. C.	8. CAS
HEAT INPUT (SIEAR) RADIATION/POTITVE L'SS	UT (STEAM) W/PUGITIVE LIYSS
RADIATION/FUGITIVE LPSS	N/FUGITIVE LYSS
Silver and a second	
SUBIDIAL NEI VALUE	SUBIDIAL NET VALUE
TUFAL ABSOLUTE VALUE	SOUTHE VALUE

TABLE 5.4
CONVERSION EFFICIENCY
BGC/SLAGGING LURGI PROCESS
20,000 TPD FACILITY

PERCENT		54.8%	53.8%	83.4%	76.4%
10 ⁶ вти/нв 27,852 525	15,268 8,404) × 100%	(3.00) + (2.0) × 100%) × 100%
INPUTS 1. COAL TO FACILITY 2. ELECTRIC POWER TO FACILITY	OUTPUTS 3. MBG FROM FACILITY 4. COAL FINES FROM FACILITY	$\frac{\text{EFFICIENCY}}{\text{COAL-TO-MBG}} (3 \div (1)) \times 100\%$	OVERALL PRODUCT EFFICIENCY (3) + ((1) + (2)	OVERALL FACILITY EFFICIENCY $(3) + (4) \div (1) + (2) \times 100\%$	GASIFICATION EFFICIENCY $(3)+(1)+(2)-4)\times 100$

TABLE 5.5. OPERATING REQUIREMENTS FOR EXPECTED OPERATIONS LURGI/BGC PROCESS - PER MODULE

	BASIS	UNITS
Raw Materials Coal Import	TPY at 100% Operation	2,777,650 TPY
Coal Fines Export	TPY at 100% Operation	589,167 TPY
Catalyst and Chemical Makeup	100% Operation	1,204,500/YR
Utility Requirements Import Power	kWh/HR at 100% Operation	259,558,800 kWh/YR
Operating Requirements Labor		
Supervisors	Man-Hours/YR	.30,000 mh/YR
Operators	Man-Hours/YR	145,720 mh/YR
Supplies	Factored as 15% of Operating Labor Cost	
Maintenance Requirements Labor	Factored at 1.6% of Total Depreciable Direct Investment	
Supplies	Factored at 2.4% of Total Depreciable Direct Investment	

6.0 DESIGN INFORMATION SOURCES

- (1) Economic Studies of Coal Gasification Combined Cycle Systems for Electric Power Generation, (January, 1978), EPRI Report AF-642.
- (2) Effects of Sulfur Emission Controls on the Cost of Gasification Combined Cycle Power Systems, (October, 1978), EPRI Report AF-916.
- (3) Evaluation of Intermediate-Btu Coal Gasification Systems for Retrofitting Power Plants, (August, 1977), EPRI Report AF-531.
- (4) Economics of Current and Advanced Gasification Processes for Fuel Gas Production, (July, 1976), EPRI Report AF-244.

APPENDIX B-5

COAI GASIFICATION FACILITY SCHEDULE ANALYSIS

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PROJECT SCHEDULES

This appendix contains the master and project schedules needed to support the evaluation of the reference facility designs, economic studies and project planning for the TVA coal gasification facility. Enclosed are five such schedules: one Coal Gasification Facility Project - Master Schedule and four project schedules for each of the four TVA Medium-Btu Gasification (MBG) Modules. The schedules are time-phased by project activity (such as engineering, procurement, construction and testing) and by module or by the major systems (such as air separation, gasification, etc.). These schedules developed at this early stage of the project should be considered as anticipatory rather than definitive and should be useful for overall project comprehension.

1.0 LEVELS OF SCHEDULES

There are three distinct types of schedules differing in the level of detail they provide. These schedules are:

1.1 Management Schedules

Management level schedules identify the basic milestones and parameters of a project. These schedules are normally originated during the planning and preliminary engineering phase of a project. The types of management schedules and a brief description of their use are as follows:

- Proposal Schedule Assists business development in its efforts to gain new contracts.
- Front End Schedule Depicts the key activities to be accomplished during the first sixty to ninety days after project award in order to successfully start the project.
- Master Schedule Defines major milestone events for engineering, procurement and construction, and provides the basic parameters for more detailed project and control level schedules.

1.2 Project Schedules

Project schedules provide a greater quantity of detail. They are designed for the use of those directly involved on a day-to-day basis in the management of the gasification project.

Project schedules depict the project plan and are developed after management level schedules have been approved and issued. The types of project schedules and a brief description of their use are as follows:

• Engineering Network - A critical path method diagram which details the activities and targets of engineering and procurement as established by the master schedule.

- Engineering Schedule A bar chart time-scaled format which details the activities and targets of engineering and procurement as established by the master schedule.
- <u>Construction Network</u> A critical path method diagram which details the activities and targets of construction as established by the master schedule.
- <u>Construction Schedule</u> A bar chart time-scaled format which details the activities and targets of construction as established by the master schedule.

1.3 Control Schedules

Control schedules provide the greatest amount of detail. Such schedules, when integrated with control reports, provide the detailed information necessary to effectively and accurately evaluate project performance. Control level schedules and reports should be maintained and updated by the person responsible for the work, in conjunction with the schedule engineer. Control level schedules and reports are the basis for project and management level schedules and reports. Control level schedules and reports are divided into three basic types: engineering, procurement, and construction. Each type of control schedule relates directly to a particular phase of the project.

2.0 CONTENT OF REPORT

This report first presents a "Coal Gasification Facility Project - Master Schedule." The master schedule is a management level schedule and depicts only major milestone events and major phases of the project such as engineering, procurement, and construction. It provides module rather than system level of detail. The purpose of the master schedule is to depict in the most general terms the overall plan for the execution of the project with regard to time and resources. The master schedule serves as a basis for the preparation of more detailed schedules and logic networks. It is important that the master schedule be agreed upon and fixed as early in the project as possible since so much of the other planning and scheduling relies upon it. It is not intended to be revised. Only major changes in scope of work or major interruptions due to factors such as labor strikes should justify revision of the master schedule.

Four project schedules are presented to accompany the master schedule. Each one depicts the time-phasing for the development of a particular module at the system level of detail. Utilizing a bar chart time-scaled format, they detail major milestones for engineering, procurement, construction, and testing.

Control schedules by their nature are too detailed for inclusion in this report. For examples of their proper use refer to: K. M. Guthrie, <u>Managing Capital Expenditure for Construction Projects</u>, Craftsman Book Company of America, Solana Beach, California (1977).

All schedules contained in this report have been made consistent with the <u>Coal Gasification Facility Work Breakdown Structure Matrix</u> as given in Appendix H.

It is recognized that the final allocation of systems within this Work Breakdown Structure may be changed in the future. For example, the category of Off Sites Systems may cease to exist, and the systems assigned to it (e.g., Coal Handling, Solids, Disposal, etc.) may be reallocated to a different category. The section which follows will describe the methodology used in preparing the schedules described above.

3.0 METHODOLOGY

The following material in this chapter outlines the four basic steps necessary in the development of a master schedule and the required project schedules. These four steps are:

- Collection of Information
- Determination of Controlling Factors
- Schedule Preparation and Analysis
- Review and Issue

3.1 Collection of Information

This initial step is the most critical since all assumptions, suppositions, and determinations are made from the data collected. The information was divided into major categories as follows:

3.1.1 Project Information

The master schedule as conceived by the Tennessee Valley Authority (TVA) gasification project team and the NASA gasification project team was used as a starting point for the schedule development. All milestones and essential schedule project schedule dates were used as the basis for schedule development.

3.1.2 Engineering, Procurement, and Construction Information

Information was collected by system type for engineering, procurement, and construction. Engineering was assumed to include activities occuring between initiation of home office engineering and design to approval for construction process implementation. Procurement was assumed to begin with the initiation of bid requests to hardware suppliers to the actual delivery of the systems on site. Construction was assumed to begin with manufacture of the field unit and implementation on site to the activities which commenced systems test. The TVA/NASA best estimates for each segment of engineering,

procurement, and construction were extracted from the master schedule and compared with the best estimates made by personnel on the BDM/Mittelhauser project team. The results of both estimates have been used to develop the project schedule enclosed.

In most cases, the BDM/Mittelhauser team's estimates were more conservative than the original TVA/NASA estimate. In any case, the least conservative estimate became the low estimate (indicated by the solid line on the project schedules) and the more conservative estimate became the high estimate (indicated by the broken line extension of the project schedule). All timelines were presented as typical for construction projects of this type and do not represent a detailed risk analysis of this project.

3.1.3 Systems and Module Test and Evaluation Information

The start-up and test cycle as defined by TVA was divided into two segments: (1) Systems Test and Evaluation set at sixteen (16) months during which time each system as it is implemented is tested independently of all other systems in the module, and (2) Module Test and Evaluation set at six (6) months during which time all systems are integrated and tested as a complete module. In all cases, systems test and evaluation overlap the construction activities and the module test and evaluation commences with completion of systems test and evaluation and ends with the start of commercial operations.

3.1.4 Cost Engineering Information

The schedule development was made to be consistent with the Work Break-down Structure as defined in the April 30 interim report. It is recognized that the subsequent development of reference designs and cost estimations might cause a reallocation of systems within the structure. At a future date, the schedules should be made consistent with any changes made to the Work Breakdown Structure which result from experience in the cost estimation and engineering design process.

3.2 Determination of Controlling Factors

The TVA Coal Gasification Facility proposed schedule as defined in Table 4.1 is the primary controlling factor for all major milestones established in the master schedule. The proposed schedule provides all the major milestone events necessary to effectively outline the execution and completion of the project and to commence commercial operation of the modules 1-4 in years 1985-87. The determination of major equipment deliveries is the prime controlling factor in establishing whether such a schedule can be met. Best estimates by TVA/NASA and BDM/Mittelhauser indicate that such a schedule is achievable, but much depends on the future procurement schedules agreed to by the hardware suppliers. No attempt has been made in this study effort to identify and contact the hardware vendors to estimate future procurement schedules for each system.

3.3 Schedule Preparation and Analysis

The information obtained in the data collection process was organized by system (e.g., gasification, process solids, etc.) and by activity (engineering, procurement, and construction). For each system, the best estimates of time requirements for the activities of engineering, procurement and construction were analyzed to determine the values to be scheduled. For example, the time estimates for TVA/NASA and BDM/Mittelhauser are given for the activities of engineering, procurement, and construction as well as the estimations for time overlap between engineering procurement and procurement-construction. In the example, the BDM/Mittelhauser estimates for engineering and procurement were lower than the TVA/NASA estimates and therefore used as the resultant low estimates for those same activities. The TVA/NASA estimate for construction was lower than the BDM/Mittelhauser estimate and therefore used as the low estimate for construction. In all cases, the overlap estimates have been used as flexible categories to assure that the combined totals can be satisfied within the project major milestones.

After completing the above process for each system, the combined totals were assigned short, medium, or long time requirement ranges according to the following definitions:

long time ≥ 40 months
medium time < 40 months
short time < 34 months

For example, the gasification system combined total is 46-51 months to engineer, procure and construct the system, and is a system having by definition, a long time requirement. All systems with similar long time requirements were scheduled first; systems with medium time requirements were scheduled second; and finally the systems with the short time requirements, and therefore the greatest flexibility, were scheduled last.

The resultant schedules are shown in Section 4.0.

3.4 Review and Issue

The submission of this document to TVA and NASA commences the final step in the schedule development. The schedule may be finalized and approved by TVA and NASA or revised as necessary to permit latest changes in plans by those organizations. In any case, project and control schedules must be developed in greater detail to permit careful monitoring of the gasification project.

4.0 REFERENCE FACILITY SCHEDULES

4.1 Milestones

The program development methodology encompasses the establishment of a specific set of time-structured elements scheduled for completion at predesignated dates. To facilitate effective program management of system development, and to ensure management review of program status, a set of objective-oriented milestones has been established. These milestones include:

- Program Requirements Review (PRR)
- Preliminary Design Review (PDR)
- Critical Design Review (CDR)
- Operational Readiness Review (ORR)
- Start of Commercial Operations (SCO)

4.1.1 Program Requirements Review (PRR)

The PRR will be a vehicle for review and approval of the complete systems requirements for all functions to be performed by the coal gasification facility. It will occur four months from the start date and will present for program management approval a complete Functional Description, a Test Plan and a list of system deliverables related to both the total systems and individual module development.

4.1.2 Preliminary Design Review (PDR)

The PDR will occur twelve months after the start date and at this time program management will review the complete system and subsystem designs. All system and subsystem specifications will be completed in draft form for review. The Test Requirements will be approved at this review. Construction of well defined systems such as coal handling, solids disposal, plant power, general facilities may begin shortly after the PDR and prior to the critical design review.

4.1.3 Critical Design Review (CDR)

Twenty months from the start date, a CDR will be held to approve all specifications. The drafts presented at the PDR will be revised as necessary to meet program development requirements, and specifications will be defined to the subsystem level. The final version of system, subsystem, and test specifications will be approved at the CDR. Approval of the CDR will mark the initiation of major construction activity for all systems not already

started. The final designs and specifications provide the necessary guidance and instructions for remaining program development activities.

4.1.4 Operational Readiness Review (ORR)

This milestone is the fourth to be reached and occurs approximately 51 months from the program start date. The objective of the ORR is to review completed system acceptance test results to determine operational readiness of each module. Complete program documentation review is also performed during this review. Following the ORR, a six-month period of module testing will commence.

4.1.5 Start of Commercial Operation (SCO)

The SCO constitutes the final phase of program development. The results of module testing and evaluation will be reviewed and commercial operation of each module will commence. Total facility management, operation, maintenance, and logistic support will proceed in accordance with the conceptualized standard operating procedures, facility operating instruction, system safety plans, and quality assurance requirements.

4.2 Schedule Dates

Table 4.1 provides schedule dates encompassing the complete life cycle of the MBG facility. The phases for module development and operation include:

- Design and Construction
- Start-up/Test (System and Module)
- Start Commercial Operation
- Module Retirement

4.3 Master Schedule

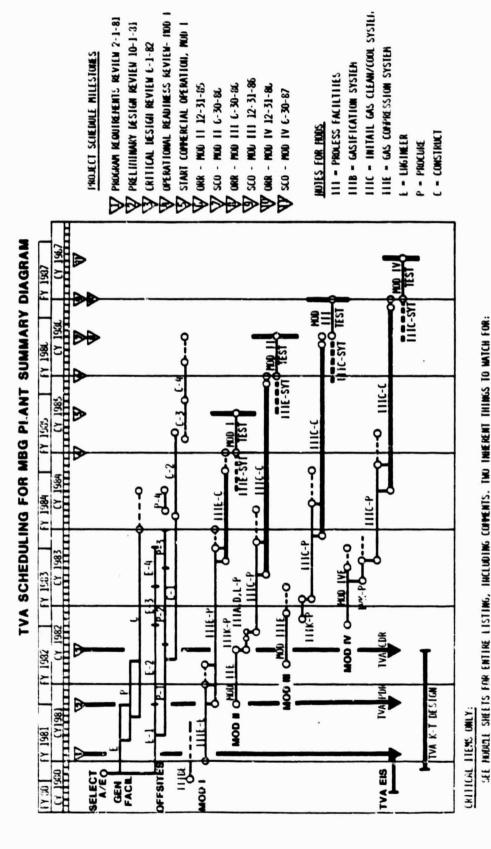
The major program development activities and their time-phased relation-ship to each of the four system modules is shown in the Coal Gasification Facility Project Master Schedule (Figure 4.1). Specific major activities include engineering procurement, construction, and testing. Also included are the program milestones and their associated dates.

4.4 Project Schedules

Figures 4.2 to 4.5 depict major systems development activities and milestones defined in Section 4.1 for each individual coal gasification module to be developed. Both low and high time estimates of each major development activity have been established according to the methodology defined in Section 4.3.

TABLE 4.1. THE TVA COAL GASIFICATION FACILITY PROPOSED SCHEDULE

	MODULE I	MODULE II	MODULE III	MODULE IV
DESIGN AND CONSTRUCT	5	[a [o ot	4-01-82	10-01-82
BEGIN	08-10-01	10-10-01		
COMPLETE	12-31-84	12-31-85	98-08-9	12-31-86
START-UP/TEST				
SYSTEM TEST - BEGIN	9-01-83	9-01-84	3-01-85	9-01-85
- COMPLETE	12-31-84	12-31-85	98-06-9	12-31-86
MODULE TEST - BEGIN	1-01-85	1-01-86	7-01-86	1-01-87
- COMPLETE	6-30-85	6-30-86	12-31-86	6-30-87
START CC'IMERCIAL OPN.	7-01-85	7-01-86	1-01-87	7-01-87
RETIRE MODULE	6-30-05	90-08-9	12-31-06	6-30-07



(1) INSURE THAT ALL POSSIBLE FARLY CONSTRUCTION IS COMPLETED BEFORE LAST EQUIPMENT
ARRIVES, 1.E., MAINIAIN OVERLAP.

(2) LIEMS B (GASIFICATION), C (GAS CLEANUP), D (ACID GAS REMOVAL) ARE TIME-CONSUMING
TO TEST, WILL REQUIRE RELATIVELY LONG TIME BEFORE ATTAINING DESIGN-SCALE EQUIL:BRIUM.

Figure 4.1. Master Schedule

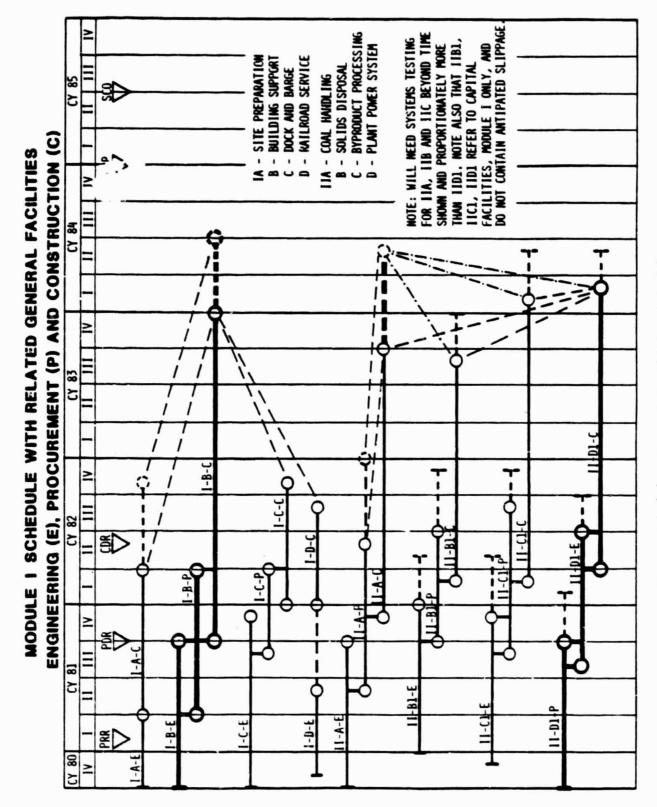
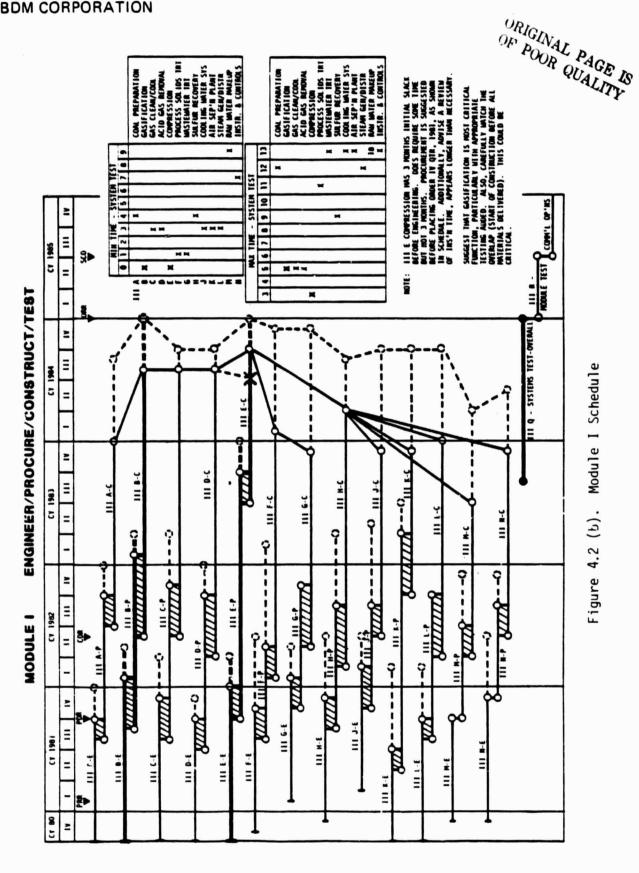
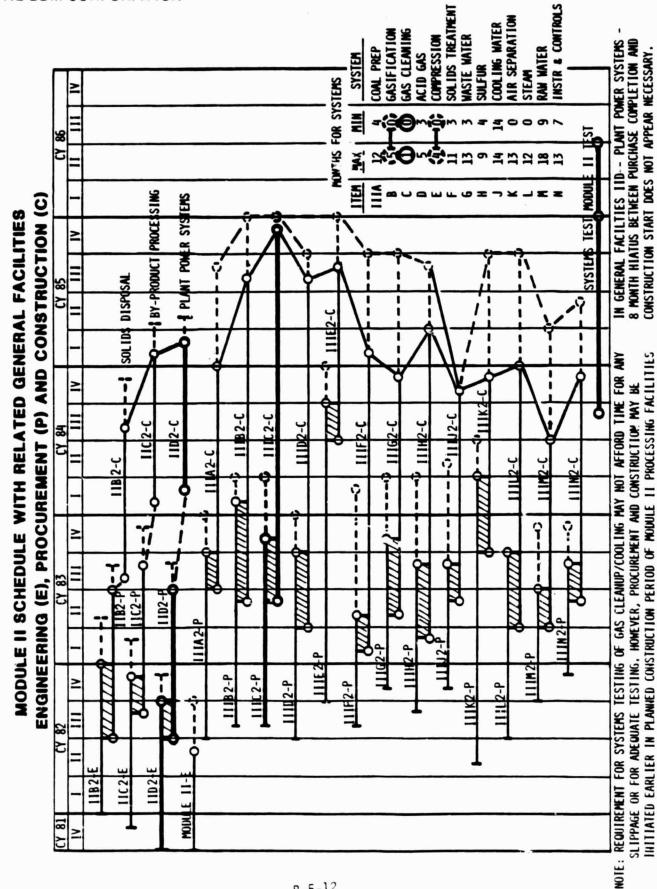


Figure 4.2 (a). General Facilities Schedule



B-5-11



Module II Schedule Figure 4.3.

HITTATED EARLIER IN PLANNED CONSTRUCTION PERIOD OF MODULE 11 PROCESSING FACILITIES

Module III Schedule

Figure 4.4.

B-5-13

COMPRESSION SOLIDS TRIMIT COOLING WATER GAS CLEANING WASTE WATER AIR SEPAR'H RAW WATER ACID GAS SYSTEM STEAM MONTHS FOR SYSTEMS 8 2 CARY-PRODUCT PROCESSING ¥ 28222 - - PLANT POWER SYSTEMS MODILE ENGINEERING (E), PROCUREMENT (P) AND CONSTRUCTION (C) MODULE IV SCHEDULE WITH RELATED GENERAL FACILITIES SYSTEMS TESTING 2 SOLIDS DISPOSAL Ξ 8 ರ 17 P ¢ ID4-C 2 ¢ þ C√ 82 9 Ç Ŷ 2 Summing S 1 d a a a a 81118 7 Ξ C 84 184-P ç IIIE4-P 11184-P 111F4-P 1184FE SITES 10FF ENGINEERING MODULE IV GENERAL

Figure 4.5. Module IV Schedule

SAME ELEMENTS APPLY TO MODULE IV AS FOR PRIOR FACILITIES.