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PITTSBURGH ENERGY TECHNOLOGY CENTER

DIRECT COAL LIQUEFACTION BASELINE DESIGN AND SYSTEM ANALYSIS

CONTRACT NO. DEAC22 90PC89857

TASK II TOPICAL REPORT - BASELINE

VOL. I OF III

DECEMBER 1991
PITTSBURGH, PENNSYLVANIA

The information and data contained in this report are the result of an ecomonic evaluation and a preliminary design effort and because of the nature of this work no guarantees or warranties of performance, workmanship, or otherwise are made, either expressed or by implication.

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December 16, 1991

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Project Manager

Subject: D.O.E. Coal Liquefaction
Base Line Design and System Analysis
Contract No. DE-AC22 90PC89857
Bechtel Job No. 20952
TASK II, VOL I. FINAL REPORT
Letter No. BLD-076

Dear Mr. Lee:

As you know the Task II report consists of three volumes which have been labelled as Volume I, Volume II and Volume III. Attached for your files are three copies of the subject report. As requested, one of these copies is bound in a loose leaf 3-ring binder for your desk use.

Copies to other members of DOE, as required by the contract are separately and directly transmitted.

This final Task II, Volume I report incorporates all the DOE comments which appeared on the "draft" issue and dated August 1991. The report does not contain any confidential information which, as agreed, would have been segregated in the appendix of this report.

On behalf of the Amoco/Bechtel project team we thank DOE/PETC for the opportunity to conduct this important direct coal liquefaction study.

Sincerely yours,

Sam N. Habash

Sam N. Habash
Project Manager

Attachment

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1. INTRODUCTION

The U.S. Department of Energy (DOE) has established a program to "foster an adequate supply of energy at a reasonable cost," in accordance with the National Energy Policy Plan IV (NEPP IV). A cost effective direct coal liquefaction program sponsored by the Pittsburgh Energy Technology Center (PETC) is an integral part of NEPP IV.

The overall goal of the coal liquefaction program is "to develop the scientific and engineering knowledge base with which industry can bring economically competitive and environmentally acceptable advanced technology for the manufacture of synthetic liquid fuels from coal.

The present assignment from PETC is undertaken by Bechtel (in collaboration with Amoco as the main subcontractor) to develop a computer model for a baseline direct coal liquefaction design based on two stage direct coupled catalytic reactors. Specifically, the scope of work calls for the development of:

1) a baseline design based on previous DOE/PETC results from Wilsonville pilot plant and other engineering evaluations,

2) a cost estimate and economic analysis, and

3) a computer model incorporating the above two steps over a wide range of capacities and select process alternatives.

In this study, the Topical Reports are also the Task reports. This Topical report addresses the baseline design development (Task II) of the direct coal liquefaction study which is based on a scale-up of the Wilsonville Pilot Plant with certain other processing alternates. The overall number of topical/task reports for this study are given below as follows:

INTRODUCTION - continued

Task No.	Title
1	Management Plan
2	Baseline and Options (Alternates) Design Development
3	Cost Estimate and Economics of the Baseline and Alternates
4	Development of Mathematical Algorithms and Models for Equipment Sizing, Scale-up, Costing and Train Duplication for Incorporation into the Aspen Simulation Program
5	Development of an Aspen Process Simulation Model of the Baseline Design and the Alternates
6	Development of a Training Manual for the Simulation Model
7	Final Report

INTRODUCTION - continued

This topical/task report for Task II is divided into three (3) volumes as follows:

Volume No.	Table of Contents Sections Covered	Planned Date of Issue
I	1 - 23	November 1991
II	24 - 42	November 1991
III	43 - 50	December 1991

Note that Volume III covers the alternate processing options which by definition lag the baseline design (Volumes I and II) by about a month.

The *Table of Contents, Introduction* (Section 1) and the Executive Summary (Section 2) are included in their entirety in all three volumes for the readers' reference.

2. EXECUTIVE SUMMARY

Introduction:

This study is an assignment of Bechtel from the U.S. Department of Energy (DOE)'s Pittsburgh Energy Technology Center (PETC) to develop a computer model for a baseline direct coal liquefaction design based on two stage direct coupled catalytic reactors,

Scope and Technical approach:

The scope of the study and the technical approach to accomplish the overall objective of the study include:

- a baseline design based on previous DOE/PETC results from Wilsonville pilot plant and other engineering evaluations,
- a cost estimate and economic analysis,
- a computer model incorporating the above two steps over a wide range of capabilities and selected process alternatives,
- a comprehensive training program for USDOE/PETC staff to understand and use the computer model,
- a thorough documentation of all underlying assumptions for baseline design and baseline economics, and
- a user manual and training material which will facilitate updating of the model for the future.

Execution Philosophy:

In order to carry out the study efficiently, the study has been divided into seven major tasks with each task having several identifiable subtasks. In Task I the study is defined. The baseline design is developed in Task II. The capital, operating and maintenance costs are developed in Task III. Mathematical models for computer simulation is developed in Task IV. Development and Validation of the model is conducted in Task V. Documentation of the process simulation and training program are conducted in Task VI. Whereas, the above mentioned six tasks are functional tasks, the remaining task, Task VII, is a level of effort task for project management, technical coordination and other miscellaneous support functions. Functional tasks (Tasks II through VI) are accomplished by a part time functional group while the project management and technical coordination are accomplished by a full time core management group (Task VII).

Topical/Task II Report Contents:

This is the Topical/Task Report for Task II. The report is divided into three volumes as follows:

Vol. No.	Table of Contents Sections Covered	Planned Date of Issue
I	1 - 23	November, 1991
II	24 - 42	November, 1991
III	43 - 50	December, 1991

Volumes I and II pertain to the baseline design while Volume III covers the options (alternates) which will be issued as indicated above.

The 42 Sections of the baseline design (Volume I & II) present information on 11 process facilities plants and 20 support facilities Plants. They have been subdivided into two volumes of approximately equivalent size. Volume I presents design information on all process plants plus the overall design considered, overall plant configuration and overall material and utility balances for the whole complex (process facilities and support facilities). Volume II, however, presents design information on all support facilities plus overall site plan, overall staffing plan, and overall Hydrogen flow, steam flow, and water flow distributions. It also contains the environmental strategy/plan for achieving environmental compliance.

For each process and support facilities plant, the following information is provided:

- o Design basis, criteria and consideration
- o Process description and block/process flow diagram
- o Material balance
- o Major Equipment List
- o Utility summary
- o Water summary (if applicable)

3. OVERALL DESIGN BASES, CRITERIA AND CONSIDERATIONS

3.1 Overall Plant Capacity And Equipment Design Criteria

This section includes the following five subsections:

3.1.1 Overall Plant Capacity
3.1.2 General Design Basis Considerations
3.1.3 Equipment Design Guidelines
3.1.4 Site Criteria
3.1.5 Utility Systems

3.1.1 Overall Plant Capacity

The strategy utilized to arrive at the Plant capacity is described below:

- The Bechtel/Amoco team in consultation with Hydrocarbon Research Institute (HRI) established the maximum size coal liquefaction train based on reactor design/fabrication/shipping limitations.

- The Maximum Plant capacity was then fixed at 61,945 BPSD of coal liquids (C_5-850°F) based on the expected production capacity of five maximum size trains, each producing 12,389 BPSD.

- The Coal throughput was then calculated to match the above output, depending on the design basis yield furnished by Bechtel/Amoco team.

Based on the above strategy, the total (5 trains) coal throughput capacity is 17,102 TPSD of moisture free (MF) coal or 15,140 TPSD of moisture and ash free (MAF) coal.

3.1.2 General Design Basis Considerations

The complex shall be designed with the following design considerations:

The only raw materials delivered to the plant shall be coal, natural gas, and untreated water. Distillate quality No. 2 Oil will be stored for start-up and as back-up.

Steam and Electric power shall be generated on-site as required by a cogeneration system powered by gas fired turbines. Fuel to the gas turbines will be medium BTU gas produced from the gasifiers. Feed gas pressure to the gas turbines will be 350 to 450 psig. Oxygen of 95 mol percent purity will be produced on-site as required. The co-product nitrogen of 99.9 mol percent purity will be used for purging and inert gas blanketing.

Coal shall be received by conveyor 2 shifts a day, 5 days a week. This complex shall be located next to coal mine(s) with no rail hopper cars or trucks are part of a mine mouth operation.

Raw water storage shall provide for seven days of plant supply.

The storage and handling facilities shall provide the following storage:

Raw coal storage	28 days
Clean coal storage	2 days
Dry coal storage before mills	8 hours
Dry coal storage after mills	8 hours
Intermediate feed storage	6 hours
Propane and mixed butane storage	15 days
Liquid products storage	30 days
Sulfur storage	15 days
Ammonia storage	15 days
Phenol storage	15 days
Distillate Fuel for Steam or Power Generation (Plant 31)	48 hours

Air cooling of process units shall be maximized, in lieu of using cooling water.

The complex shall conform to all applicable environmental,safety and health regulations. These will include the following:

EPA - Water pollution, air pollution, solid waste disposal.

FAA - Aircraft warning.

OSHA - Safety, noise, sanitary.

The complex shall further conform with the applicable standards within the latest editions of the following:

ACI - Concrete application

AISC - Design, Fabrication & Erection of Structural Steel for Buildings

ANSI - B31.1 Piping and Valve Design and Selection

API -	Plant Safety and Equipment Design
AREA -	Manual for Railway Engineering
ASME -	Pressure Vessel and Boiler
ASTM -	Materials of Construction
AWS -	D 1.1 Structural Welding
FM -	Factory Mutual Approved Guide
IEEE -	National Electric Safety Code
ISA -	Instrument Design Code
NEC -	Electrical Safety Code
NFPA -	National Fire Codes
TEMA -	Heat Exchanger Design
UL -	Electrical Safety Test

3.1.3 Equipment Design Guidelines

The plant equipment designs shall be based on the following guidelines:

Pumps

Pumps conforming to API Standard 610 and AVS Standards are acceptable as determined by process requirements.

Pump sparing is to be minimized. Power failure is a valid reason for sparing critical pumps.

Primary drive shall be electric. Turbine drives to be determined by steam balance and critical drive requirements.

Compressors

Process performance requirements shall determine the choice between centrifugal and reciprocating compressors.

Pressure Vessels and Boilers

Pressure vessels shall be designed per ASME Section 1.

Field erected vessels to be minimized.

Boilers to be fired by plant fuel gas.

Fired Heaters

Vertical firing with horizontal furnace tube arrangement is preferred for larger requirements.

Plant fuel gas shall be the primary fuel.

Shell and Tube Heat Exchangers

The preferred straight tube length is 20 feet. Tube diameter shall be 3/4" with 14 BWG min. wall thickness for low alloy materials (5 CR - 1/2 Mo and below) and 3/4" with 16 BWG min. wall thickness for high alloy materials (9 CR - 1 Mo and above.)

Cooling water thermal relief valves shall be set at 75 psig.

Air Coolers

A tube length of 40 feet is preferred.

Air coolers shall be designed for local site conditions.

Cooling Towers

Cooling towers shall be designed for local site conditions. Cold and hot water temperatures are defined in the Cooling Water System Section.

Piping

Relief valves handling non-flammable materials such as CO_2, air, etc shall be vented to the atmosphere a minimum of 10 feet above any platforms or other structures.

Relief valves handling light hydrocarbons and any other flammable materials shall be vented to a closed relief system discharging to a flare system.

Waste streams such as storm run off water, oily water, sanitary water,and chemical drains shall be segregated to allow individual treatment for recycle or disposal purposes.

Minimum overhead pipe clearance beneath main pipe racks is 14'. Minimum overhead pipe clearance over roadways is 22'.

All piping to be hydrostatically tested ANSI B 31.3.

Civil and Structural

Erection, wet and dry operations, and hydrostatic testing shall be considered in determining load conditions.

Wind velocity used for structural design purposes shall be as specified by the site conditions. Tornados or violent storms shall also be considered.

Wind pressure and its net coefficient shall be as specified in ANSI A 58.1.

A factor of safety against overturning of 1.5 shall be used for both erection and operation.

Snow loads shall be as specified by the site conditions.

Compressive strength of concrete for foundations and walls shall be 3000 psi.

Instrumentation

Basic instrument system is to be electronic except for local mounted controllers which may be pneumatic.

Field installation and instrument equipment must comply with the areas electrical classification.

Instrument signal wires to be both overhead and underground as conditions permit.

Instrument identification and symbols per ISA - S5.1

Sufficient instrumentation shall be specified to provide reliable semi-automatic operation and performance evaluation of the plant with a nominal sized operating staff.

Utility flow rates shall be metered and recorded as process unit totals.

Control room instrumentation shall be a distributed control system (DCS). Operator interface shall be located in one central control room and shall consist of consoles with each console containing operator stations and printers. Every operator station shall consist of a keyboard and a color CRT with graphics display and control capability. DCS system shall have data history, alarming, recording, integration and computation capability on all selected points.

Large logic systems shall be handled by programmable logic controllers, (PLCs), with operator interface through graphic displays on CRTs.

There shall be a supervisory computer for data management and advanced control which is connected into the DCS communications network. Operator interface shall be through the DCS operator stations and through computer system printers.

There shall be an engineering and maintenance console consisting of DCS and computer operator stations and printers.

Location of I/O modules and digital controllers shall be in the main control room or in satellite buildings depending upon the distances involved.

Analyzers shall be located in analyzer houses strategically located in the process areas.

Instrument hardware redundancy shall be provided to preclude loss of multi-loops due to one failure.

3.1.4 Site Criteria

The site criteria given below shall be used for the facility.

The plant shall be a "grass roots" facility.

The site is an "inland" location in Southern Illinois with rail and road accessibility.

Sufficient water at 50 PSIG and natural gas at 350 PSIG supplies are available at the plant fence.

The site is relatively level and dry. A soil load bearing capacity of 2500 pounds per square foot minimum shall be assumed for design calculations.

The plant shall be designed for seismic risk zone 2.

Other site information is given in subsection 3.3.

3.1.5 Utility Systems

The plant utilities shall be designed as shown in Table 3.1 through 3.3

Table 3.1

Steam Systems

		Pressure,psig	Temp.,°F
High Pressure (Super-heated)	maximum	650	750
	normal	600	720
	minimum	550	
High Pressure (Saturated)	maximum	650	497
	normal	600	489
	minimum	550	
Medium Pressure (Saturated)	maximum	180	380
	normal	150	366
	minimum	130	
Low Pressure (Saturated)	maximum	60	307
	normal	50	298
	minimum	45	

Table 3.2

Boiler Feedwater Systems

High Pressure	1100	240
Medium/Low Pressure	250	240

Table 3.3

Cooling Water System

Supply	50	87
Return	35	115

Start-up and Emergency Fuel Gas System - 350 psig at ambient temperature.

Air Systems

Plant Air	100 psig at ambient temperature.
Instrument Air	100 psig at ambient temperature.

Instrument Air shall be dried to -40°F dew point.

Electrical System

Steam and Power Generation (plant 31) will consist of a combustion turbine generator (CTG) based cogeneration plant providing both steam and electric power for the Liquefaction Facilities. In some cases Plant 31 will also contain fluid bed combustion boilers (FBC) for consumption of 1000°F+ material and any middlings from the coal cleaning operation. The design criteria for the plant is shown in Table 3.4 below:

Table 3.4

Operating Mode	Steam	Electricity	Fuel	Notes
Normal	Normal	Self-Sufficient	Synthesis gas to CTG,(s); solid fuel to FBC(s)	
Start-up	Start-up	Start-up	Distillate Oil to CTG(s) and FBC(s)	(1),(2)
Single largest electric generator or steam generator out-of-service	Above Normal	Self-Sufficient	Synthesis gas to CTG,(s); solid fuel to FBC(s)	(3)

Notes
1. Start-up utility loads will be minimized by sequential start-up of Plant 31 CTG(s).
2. Start-up distillate oil consumption will be minimized by shifting CTG(s) to synthesis gas as fuel becomes available.
3. Consumption of electricity to be minimized during outage period by shedding non essential loads and shifting rotating equipment to steam turbine drives. Above normal steam generation to be accomplished by use of synthesis gas grid burners in Plant 31 heat recovery equipment.

Generators will produce electric power at 13.8 Kv, 3 phase, 60 Hz. Distribution of electric power from the generators to the main switch-gear centers will be 69 Kv 3 phase, 60 Hz.

Electric motors will be connected as follows:

- 1/4 - 1/2 HP, 110 volts, 1 phase, 60 Hz
- 3/4 - 150 HP, 460 volts, 3 phase, 60 Hz
- 200 - 5,000 HP, 4160 volts, 3 phase, 60 Hz
- 5,000 HP plus, 13.8 Kv, 3 phase, 60 Hz

Lighting Distribution will be as follows:

- Incandescent and emergency 110/208 volts
- Mercury Vapor and Fluorescent 277/480 volts

Stand-by power will be required for lighting and instruments.

Nitrogen System

Nitrogen will be used for purging and inert gas blanketing. The liquid nitrogen from the air separation facility will be stored and vaporized as needed. Nitrogen will have 99.9% purity and be supplied at 150 psig. When steam-methane reforming is utilized, a nitrogen generating system will be installed. Nitrogen supplied from the generator will be 99.9% purity and supplied at 150 psig.

Fire Systems

Fire protection and control systems for all plants, structures and equipment within the complex are to be designed in compliance with all federal, state and local codes and standards and with the recommendations of the American Petroleum Institute, National Fire Protection Association, and Oil Insurance Association (OIA).

Sewage and Effluent Water Treatment Systems

Sewage and Effluent Water Treatment (Plant 34) and Raw Water Systems (Plant 32) are closely related and constitute the Project's Water Management System, with the purpose of minimizing raw water consumption and effluent discharge to public waters during normal plant operation. Wastewater streams are segregated on the basis of their compatibility and are treated as necessary to make them suitable for reuse, if practical, in lieu of importing fresh water.

3.2. Plant Numbering System

3.2.1 Philosophy

In this report the philosophy of numbering is addressed separately for Plants, Parallel Trains, Alternates and Individual Equipment related to a particular Plant. These are separately highlighted below.

Numbering For Plants:

The total coal liquefaction complex is comprised of a group of Plants. These Plants are divided into two categories:

1) Primary Process facilities, and 2) Support facilities.

Each plant in either of these two facilities is individually numbered. For example: coal cleaning and preparation is Plant number 1, coal liquefaction is Plant 2, etc.

Numbering For Parallel Trains:

Parallel trains are numbered by adding suffix A, B, C, etc. to the plant number. (For example if there are three parallel trains for coal liquefaction plant (Plant 2) it will be numbered as 2A, 2B and 2C).

Numbering For Sections:

Numbers for each primary process plant and support facilities are listed in Tables 3.5 and 3.6, respectively. Each plant is subdivided into various identifiable sections.

Numbering For Alternates:

The alternates are numbered by adding hyphenated suffix to the plant number, or plant section number (For example; 9 is the plant number of hydrogen production). The numbering for the alternate option for hydrogen production by natural gas reforming is 9-01.

Numbering of Individual Equipment at a particular location:

The basic numbering scheme for equipment is highlighted following an example as shown below:

2.3 - 01 - E101, where

"2" is the plant number
".3" is the section number of the plant
"-01" is the addition number if the plant or section is an option off the base-line design
"-E" is Bechtel's equipment letter code (listed in Table 3-xx) for the type of equipment
"101" is the sequential number for that type of equipment. If there are multiple equipment items in the same service, they will have unique numeric designations; the sequential equipment numbering will be for the entire plant, and not a separate sequence for each section. Each option will have its own unique numbering sequence, starting over with "101".

3.2.2 Primary Process Plants Numbering System

The main process plants numbers and description are shown in Table 3.5. However, the item numbers and section numbers of these plants along with any alternate section numbers are shown in the major equipment list section of each baseline and optional plant.

TABLE 3.5

Process Facilities Identification
Code for Baseline
and Alternates

Process Plant No.	Process Plant Title
1	Coal Cleaning and Preparation (Jigging)
1-01	Coal Cleaning Option (Heavy Medium Separation)
1-02	Coal Cleaning Option (Spherical Agglomeration)
2	Two Stages Coal Liquefaction
2.3 - 01	Reactor Configuration Option (Thermal-Catalytic Reactor)
2.3 - 02	Reactor Configuration Option (Catalytic-Catalytic with Vent Gas Separation)
3	Gas Plant
4	Naphtha Hydrotreater
5	Gas Oil Hydrotreater
6	Hydrogen Purification
7	OPEN
8	Critical Solvent Deashing Unit
8-01	Vacuum Bottoms Processing Option (Delayed Coker - deleted)
8-02	Vacuum Bottoms Processing Option (Fluid Bed Coker)
9	Hydrogen Production By Coal Gasification
9-01	Hydrogen Production Option (Hydrogen Production By Natural Gas Reforming)
10	Air Separation
11	By-Product Sulfur Recovery Unit

3.2.3 Support Facilities Numbering System

The main support facilities plant numbers and description are shown on Table 3.6. However, the sections numbers of these plants along with any alternate section numbers are shown in the major equipment list section of each baseline plant and optional plant.

TABLE 3.6

Support Facilities Identification Code for Baseline and Alternates

Support Facilities No.	Support Facilities
19	Relief and Blowdown
20	Tankage
21	Interconnecting Piping Systems
22	Product Shipping
23	Tank Car/Tank Truck Loading
24	Coal Refuse and Ash Disposal
25	Catalyst and Chemical Handling
26 - 29	OPEN
30	Electrical Distribution System
31	Steam and Power Generation (Co-generation)
31-01	Fluidized Bed Combustor/Steam Turbine Generator
32	Raw, Cooling and Potable Water Systems
33	Fire Protection Systems
34	Waste Water Treatment Systems
35	Instrument & Plant Air Systems
36	Purge and Flush Oil System
37	Solid Waste Management
38	Ammonia Recovery
39	Phenol Recovery
40	General Site Systems
41	Buildings
42	Telecommunications Systems

3.3 Plant Site Information

General Setting

The generic site for the Coal Liquefaction complex is Southern Illinois near the mouth of a surface mine. Perry Country is identified, only to serve as a location to obtain site information. The county is not part of a metropolitan area and had a population of 21,000 in 1980. The assumed site is in the southern part of the county as it contains the surface mines. The closest cities are Pinckneyville (1980 population: 3319) and DuQuoin (1980 population: 6554). St. Louis MO. is about 70 miles away. Pinckneyville is the county seat.

Community Support

Perry County is located in southwestern Illinois. In road miles the proposed site is approximately 240 miles from Memphis, 310 miles from Chicago, and 320 miles from Kansas City. A Junior College is located in Ina and a University in Carbondale, both about 30 miles away. A wide variety of activities is possible in the area including golf, tennis, fishing, boating, and hunting. In addition nearby St. Louis, one of the cultural meccas of the mid-west provides cultural and recreational activities year-round.

Population Trends

Population statistics are shown below for Perry County.

1970.....19,957 1980.....21,714 1985.....22,058

It can be seen that Perry County has had a slow steady growth. The county is still lightly inhabited, with an average density of 49 persons per square mile.

Industrial Employment

The coal industry has been the main support of the area's economy for the last 30 years and will continue to play an important role for years to come. About 10 percent of the total labor force is employed in coal production, preparation, processing, development and services.

Perry County is the largest coal producer in Illinois. There are one underground and six surface mines in the county.

Perry County has a work force of approximately 15,400 people. A few of the major employers in the area are listed in Table 3.7.

Table 3.7

MAJOR EMPLOYERS IN PERRY COUNTY

Employer	Product/Service	Employment
MCA Distribution	Phonographic Records	400
Consolidated Coal Co.	Coal Mining	375
Blazon-Flexible Flyer	Outdoor Recreational Equipment	350
Consolidation Coal	Coal Mining	275
Cablec Corporation	Wire Cable	275
Freeman United Coal	Coal Mining	265

Labor Availability

For the past year, unemployment in Perry County has been approximately 16 percent of the total civilian work force, indicating an excessive supply of labor.

Labor-Management Relations

All the large facilities, except Blazon-Flexible Flyer, are unionized.

Educational Facilities

The Pinckneyville school system operates three elementary schools, one junior high school and one high school. DuQuoin operates four elementary schools, one junior high school and one high school. There are also two private schools and a high school in the area, operated by religious denominations.

Southern Illinois University at Carbondale and Rend Lake Junior College are both about 30 miles away. Pinckneyville also has a technical career college.

Culture and Recreation

Local churches and schools are used for a variety of events such as community concerts, fashion shows, little theater presentations, and gospel,rock and country music concerts. Being a rural county, the residents also look to nearby St. Louis for many of their cultural attractions.

Recreational activities typically emphasize outdoor activities, such as public parks with tennis courts and golf courses; swimming pools and lakes; as well as private country clubs.

Transportation Facilities

The Pinckneyville-DuQuoin airport facilities include charter service, storage and maintenance for private planes, and a 4,000 foot lighted, hard surface runway. Lambert International airport (St. Louis) provides all cargo and passenger service.

Rail service is provided by Amtrak, Central Illinois and Missouri Pacific Railroads.

The area is served by US highway 51 and state highways 13, 127, 152, and 154. Additionally Interstate 57 has an interchange about 20 miles away. There are 10 interstate motor carriers serving the area.

Climate

Perry county has a continental climate. Summers are hot and winters are cold. In January the average temperature is 31.6°F and the average daily minimum temperature is 22.0°F. In Summer the average temperature is 76.3°F and the average monthly maximum temperature is 88.3°F. The elevation is 650 feet above sea level. Design frost line for this area is 18 inches.

The average total annual precipitation is 42.5 inches and the average annual snowfall is 14 inches. The average relative humidity at noon is 57 percent and 85 percent at night.

Prevailing winds are from the SW at 5 mph during the summer and from the W to WNW at 8 mph during the winter.

Tax Rates

A summary of the tax structure for the state of Illinois and Perry County is as follows:

Illinois Corporate Income Tax	4.8%
Illinois Corporate Personal Property Tax	2.5%
Individual Income Tax	3.0%
Illinois Sales Tax	6.25%
Average Incorporated city Property Tax	.065/$
Assessed value	33.3% of market value

Location

The location of this generic site is in Perry County, Illinois.

Meteorological Information

Meteorological site conditions, for design purposes shall be as shown in Table 3.8.

Table 3.8

Temperature	**Dry Bulb, °F**	**Relative Humidity, %**
Winter - Minimum	-18	
Winter - Design	-06	75
Summer - Maximum	104	
Summer - Design	95	45
Wind, Prevailing Direction		
Winter	NW	
Summer	SW	
Wind, Design Velocity	15	
Wind, Peak Gusts	90	
Rainfall, Inches		
Yearly	40	
Maximum in One Hour	2.6	
Maximum in 24 Hours	5.6	
Snowfall, Inches		
Yearly	14	
Maximum in 24 Hours	10	
Maximum in One Month	20	
Design Frost Line, Feet below surface	1'- 6"	
Plant elevation, feet above sea level	650	
Normal atmospheric pressure,psi	14.3	

Coal Characteristics

The facility shall be designed based on the following properties of Illinois No. 6 coal, Burning Star Coal.

ROM coal analysis[1] is shown in Table 3.9

Table 3.9

Proximate Analysis, wt.%	**As Received**	**Moisture Free**
Moisture	8.7	--
Ash	20.0	21.91
Volatile Matter	33.0	36.14
Fixed Carbon	38.3	41.95
	100.0	100.00

Ultimate Analysis, wt.	**Moisture Free**	**Moist & Ash Free**
Carbon	~~61.5~~ 61.1	~~78.75~~ 78.03
Hydrogen	4.2	~~5.38~~ 5.36
Nitrogen	1.2	~~1.54~~ 1.53
Sulfur	5.1	~~6.53~~ 6.51
Chlorine	0.1	~~0.12~~ 0.13
Ash	~~21.9~~ 21.7	---
Oxygen (by difference)	~~6.0~~ 6.6	~~7.68~~ 8.43
	100.0	100.00

Gross Heating Value, dry,BTU/lb. 10600

RAW WATER CHARACTERISTICS

Water Properties	Mid Continent Location Avg.	Max.
pH	6	8.5
Total Hardness as $CaCO_3$ ppm	140	180
Calcium as Ca, ppm	38	49
Magnesium as $MgCO_3$, ppm	96	130
Bicarbonate as HCO_3 ppm	137	173
Sulfate as SO_4 ppm	350	570
Silica as SiO_2 ppm	6.5	7.0
Total Dissolved Solids, ppm	662	1449
Temperature, °F (Range)	40--89	---
Conductivity, moh	1000	1200
Turbidity, Jackson turbidity units	---	20

(1) DOE's RFP on Direct Coal Liquefaction Project

3.4 Product and By-Product Specifications

None of the main liquid products from the coal liquefaction facility are considered to be finished or fungible. They will be stabilized for further treatment by hydrotreating. Products will be shipped by pipeline as a refinery feedstock for further refining to premium value stocks.

Naphtha

A hydrotreated coal naphtha will be produced which is suitable for additional refining (reforming) to gasoline. Primary specifications and estimated ranges for this product are shown in the table below. Other inspection data will depend on the specific processing scheme and operating conditions selected.

Distillates

Two coal distillate products, a light and a heavy distillate, will be produced. The products will be suitable for additional processing to fuel oil and gasoline (catalytic cracking, hydrocracking). Primary specifications and estimated ranges for these products are shown in the table below. Other inspection data will depend on the specific processing scheme and operating conditions selected.

Gas Oil

A coal-derived light vacuum gas oil will be produced. The product will be suitable for additional processing to fuel oil and gasoline (catalytic cracking, hydrocracking). Primary specifications and estimated ranges for this product is shown in the table below. Other inspection data will depend on the specific processing scheme and operating conditions selected. The liquefaction facility will not produce a saleable solid product. Gaseous product will be used as fuel.

Primary Product Specifications

The primary specifications of the four products are given in the Table below.

	Naphtha	Light Dist	Heavy Dist	Gas Oil
Boiling Range °F	C_5 - 350	350 - 450	450 - 650	650 - 850
Sulfur, ppm max	1.0	*	*	*
Nitrogen, ppm max	0.2	*	*	*
Oxygen ppm max	30	--	--	--

* **Note:** Sulfur and Nitrogen for the 350 to 850°F stream are 20 and 500 ppm respectively:

By-Products Specifications

Sulfur will be produced as a by-product of the gas clean-up operations. It will be produced as a marketable (99.5 wt. % S (dry) bright yellow.

Ammonia will be produced as a by-product of waste water treatment. It will be produced as a marketable (99.0 % min. purity) anhydrous ammonia stream.

The Phenolics stream will be produced as a by-product of waste water treatment. It will be produced as a marketable (87.0 % min. purity) Phenolic product.

3.5 Baseline

The Baseline design development for this complex is the primary activity of Task II. Therefore the design bases, criteria and considerations section (Section 3 of this report is dedicated to this purpose. Thus sections 1-43 of this report (Vols. I and II) pertain to the Baseline design while sections 44-50 are concerned with options or alternates to the baseline (Vol. III of this report).

3.6 Options (Alternates) to Baseline

As stated in Section 3.5, the options to the Baseline design are covered in sections 44-50 of this report. The design basis, criteria and considerations for the design of the options is given in the first subsection of each of these sections. For example, the design basis for plant 8-02 (fluid coking option) is subsection 49.1 of Section 49 of Vol. III of this report.

3.7 Overall Execution Methodology

The overall execution methodology is based on the following understandings between Bechtel/Amoco and D.O.E.:

1) The computer model <u>is</u> a planning/research guidance tool to guide DOE and its subcontractors' further research and to study the cost/economic impacts of existing and/or new designs of various plants on the overall complex economics. This will be done on "case study basis" to reduce convergence time and permit the model to be run on a 386 PC computer.

   Therefore, the model is not designed to be a plant design and sizing program for every plant in the complex. Due to the high level of interest in coal liquefaction technology Plant 2 (coal liquefaction plant) is the only plant where major equipment sizing information is printed. Plant 2 is also the only plant for which a kinetic model is provided as an additional item (not as an integral part of the model).

2) The cost estimates for the base case were rigorously prepared for the accuracy required by this study (± 30%). Other estimates for other capacities were not prepared in the manner as previously envisioned. Originally it had been assumed that other capacities would be computed by the model first (using equipment scaling factors) and then costed by the model using cost vs. capacity information. This is a time consuming and computer space consuming procedure which is not necessary for a ± 30% estimate.

   Bechtels' normal estimating techniques for this type of estimate is to prepare a reliable base case cost estimate and then directly scale that estimate (based on cost vs. capacity factors) to other capacities. The same technique will be used for this estimate.

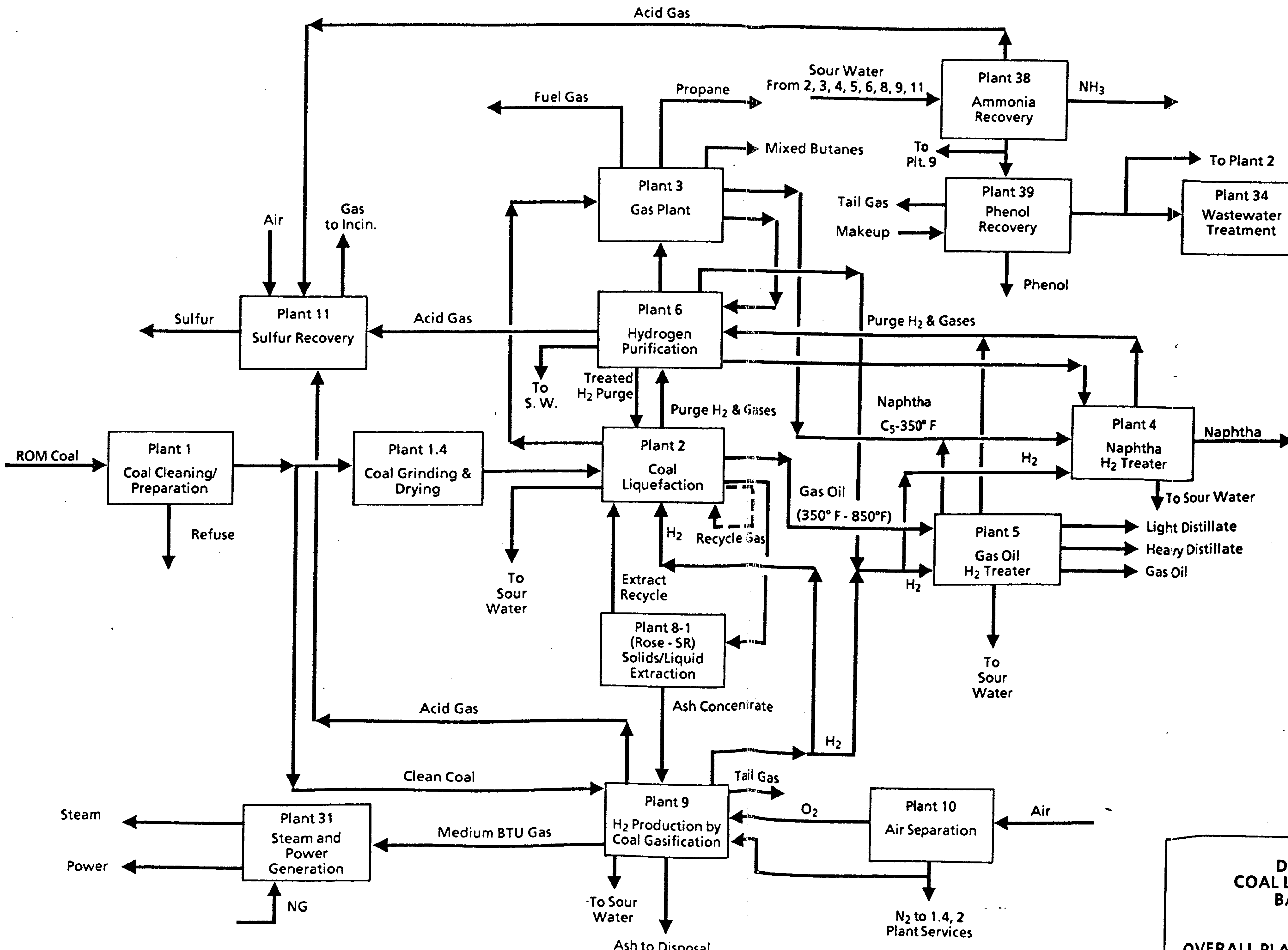
3) Heat and material balances suitable for major equipment sizing calculations were prepared by Bechtel on SimSCI Process Simulation Software, (PROCESS). Due to the inherent differences of this program from ASPEN, it was deemed appropriate to maintain Process as an internal document and work exclusively with ASPEN as the software on which all published material balances, heat balances, stream properties, elemental balances, etc. will be based. Bechtel inputed to Amoco all the necessary information to make the results of both softwares consistent.

   But, ASPEN will be the program of record for all information produced by the model.

   For this reason, the Task II report which is largely based on PROCESS will not contain any detailed material balances. These, and all process simulation results, will be published as part of the Task 5 (Aspen modeling) report.

4. Overall Plant Configuration

The plant configuration for the entire coal liquefaction complex is shown in Figure 4.1.



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DOE / PETC
COAL LIQUEFACTION
BASE CASE

OVERALL PLANT CONFIGURATION
Figure 4.1

5. Overall Material and Utility Balances

This section is divided into the following two subsections:

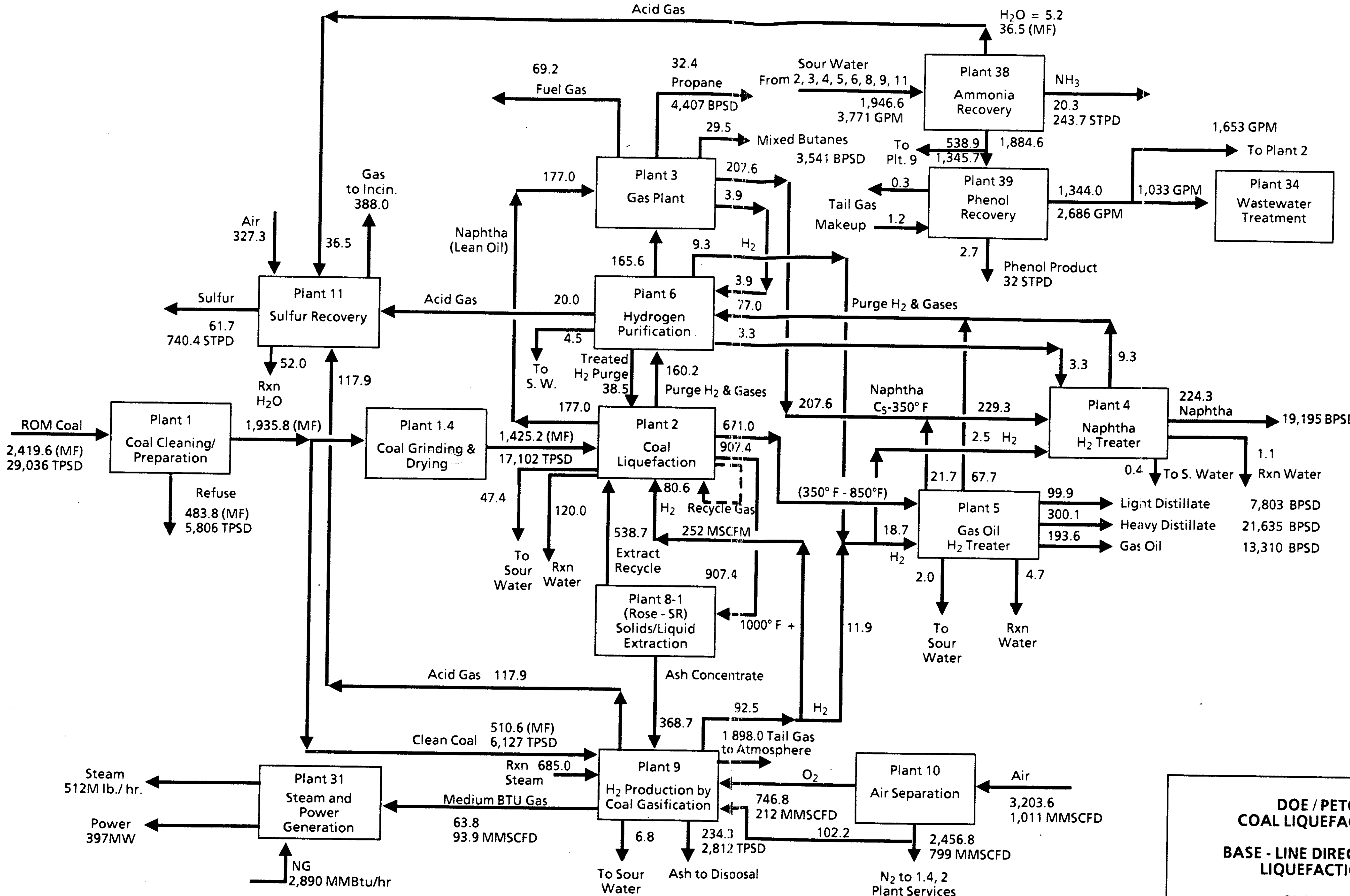
5.1 Material Balance
5.2 Utility Balance

5.1 Overall Material Balance

The overall material balance for the baseline coal liquefaction complex is shown in Figure 5.1.

5.2 Overall Utility Balance

The overall utility balance for the baseline coal liquefaction complex is shown in Table 5.1.



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Notes:

1. Flow rates are in MLB/HR unless noted and on dry basis
2. Simplied water flow distribution diagram is shown on Figure 41.1
3. Minor streams including steam, water, sour water, and make-up amine are not shown on this diagram
4. Flow rates around plants #38, 39, 34 are shown on wet basis

DOE / PETC
COAL LIQUEFACTION

BASE - LINE DIRECT COAL LIQUEFACTION

OVERALL MATERIAL BALANCE

Figure 5.1

27-Nov-91

TABLE 5.1
OVERALL UTILITY BALANCE
COAL LIQUEFACTION – BASE-LINE CASE

Plant No.	PLANT NAME	Fuel Gas MMbtu/hr	Electric Power KW	Steam Psig, Lb/hr 600 Sup @720 F	600 @489 F	150 @366 F	50 @298 F	Cond/BFW Net Cons lb/hr	Cooling Water gpm	Sour Water gpm	Waste Water gpm	Makeup Water gpm	Air scfm	Nitrogen scfm
	PROCESSING UNITS													
1	Coal Cleaning and Handling		8,289								785			
1.4	Coal Griding and Drying	486	12,114						240				5,956	75,063
2	Coal Liquefaction	1,090	58,988	5,565	(149,090)	(90,095)	56,115	424,715	8,095	(2,428)				715
3	Gas Plant	(1,473)	913		396,610		17,186	(413,796)	21,955	(5)				
4	Naphtha Hydrotreater	74	1,073		26,622			(26,622)	2,564	(40)		(38)		
5	Gas Oil Hydrotreater	162	2,179		65,048	(33,864)		(26,184)	8,228	(88)		(73)		
6	Hydrogen Purification		43,944				59,473	(59,473)	6,662	(70)		(60)		
8	Critical Solvent Deashing	212	4,363			25,419				(51)				
9	H2 Production by Coal Gasification	(981)	63,210		(291,800)	(762,900)	(781,300)	1,836,000	52,600	(966)	95	(1,196)		23,611
10	Air Seperation		159,994											(555,197)
11	By-product Sulfur Recovery	69	3,062		59,956	(186,541)	54,468	72,117	8,126	(123)				
ALL	Common Users			73000		40000	90000	(193,000)				(532)	9044	
	Sub Total (Process)	(361)	358,129	78,565	107,346	(1,007,981)	(504,058)	1,613,757	108,470	(3,771)	880	(1,899)	15,000	(455,808)
	OFFSITES													
19	Relief and Blowdown		19											
20	Tankage		6,758			22,000	3,300	(25,300)	100					
30	Electrical Distribution													
31	Steam and Power Generation	3,871	(396,295)	(511,963)		381,996	436,191	(593,853)	65,667		(322)	(898)		
32	Raw, Cooling and Potable Water Systems		13,652						(255,529)		(1,011)	12,042		
33	Fire Protection		44											
34	Sewage & Effluent Treatment		7,313				3,000	(3,000)				(1,011)		
35	Instrument & Plant Air Systems		2,894										(15,000)	
36	Purge and Flish Oil System		258											
37	Solid Waste Disposal		48								90			
38	Ammonia Recovery		1,531		326,052	581,269		(907,321)	76,821	3,771				
39	Phenol Recovery		814			22,716	61,567	(84,283)	4,471		(1,033)			
Other	Bldg 41,42, Light, Etc.		4835											
	Natural Gas Imported	(3,510)												
	Off Site BL and Evapration										1,396	(8,234)		
	Sub Total (Offsite)	361	(358,129)	(511,963)	326,052	1,007,981	504,058	(1,613,757)	(108,470)	3,771	(880)	1,899	(15,000)	0
	GRAND TOTAL	0	0	(433,398)	433,398	0	0	0	0	0	0	0	0	(455,808)

6. Plant 1 (Coal Cleaning and Handling)

6.0 Design Basis, Criteria and Considerations

ROM coal enters the fence line of the complex. The analysis of ROM coal is shown in Table 6.1. ROM coal is cleaned by Jig Cleaning operation.

Storage capacities for ROM Coal and clean coal are as follows:

> For ROM coal, "Active Storage" Pile is of 4 days capacity and remaining 24 days of "Inactive Storage".
>
> For clean coal, storage is for 2 days of "Active Storage" pile.

Operation Basis for the base case cleaning plant (Jig) is:

> 2 shifts a day 5 days a week and 50 weeks per year
> Third shift for maintenance
> 14.5 hrs of operation with 1.5 hrs for start-up, planned and unplanned outages.

Ash content of base case clean coal are as shown below:

Base case, Jig Cleaning	11.47%
Alternate 1, Heavy Medium	8.6%
Alternate 2, Spherical Agglomeration	< 4%

Size analysis for ROM Coal is shown in Table 6.2.

The basis for analysis of clean coal is shown in Table 6.3.

The production rate of this plant is based on the requirement of coal liquefaction plant (Plant 2) and coal gasification plant (Plant 9, the source for hydrogen supply)..

Coal from this coal cleaning and handling plant (Plant 1) is further processed through coal grinding and drying plant (Plant 1.4) where the coal moisture is reduced to 2 wt% and the particle size is reduced to 50% through 200 mesh screen.

Plant 1 design is on the basis of six train each train processing nominal 5000 TPSD on MF basis. The product size, ash and moisture content of clean coal from Plant 1 is shown in Table 6.4.

Production of Middlings

The possibility of utilizing a potential middlings product with an ash content of around 20 percent for boilers or coal gasification was examined. The recovery of a middlings product is justified when the process sink material (refuse) contains a significant amount of material of economically usable quality. A review of the float/sink data for the design coal revealed that only a clean coal with 11.47 percent ash (Base Case) will be produced from the ROM coal, the refuse will contain no coal of the required middlings quality. The refuse will be exclusively material with an ash content above 68 percent.

Under these circumstances, additional cleaning, dewatering, handling and storage circuits will not be justified to produce a middlings product. Since the boilers or gasifier can use a fuel of 20 percent ash this requirement can be more economically met by ROM Coal.

Table 6.1

Illinois No. 6 Coal Burning Star Mine ROM Coal Analysis [1]

Proximate Analysis	Wt. %
Volatile Matter	33.0
Fixed Carbon	38.3
Ash	20.0
Moisture	8.7
Ultimate Analysis	
Carbon	61.1
Hydrogen	4.2
Nitrogen	1.2
Sulfur	5.1
Chlorine	0.1
Ash	21.7
Oxygen (by difference)	6.6
Sulfur Forms	
Pyrite	3.0
Sulfitic	0.3
Organic	1.8
Ash Composition	
Phosphorus pentoxide, P_2O_5	0.1
Silica, SiO_2	43.8
Ferric Oxide, Fe_2O_3	24.1
Alumina, Al_2O_3	17.1
Titania, TiO_2	0.8
Lime, CaO	5.6
Magnesia, MgO	1.0
Sulfur Trioxide, SO_3	4.1
Potassium Oxide, K_2O	2.1
Sodium Oxide, Na_2O	0.6
Undetermined	0.7

(1) DOE's RFP on Direct Coal Liquefaction Project with Adjustment

Table 6.2

Size Analysis for ROM Coal

Size (inch or mesh)	Percent
3 x 3/8	51
3/8 x 28M	37
28M x 0	12
Total	100

Table 6.3

Analysis of Feed Coal to Liquefaction
Illinois No. 6 (Burning Star Mine)

Proximate Analysis (wt. %, Dry Basis)	
Volatile Matter	42.2
Fixed Carbon	46.3
Ash	11.5
Total	100.0
Ultimate Analysis (wt. %, Dry Basis)	
Carbon	71.0
Hydrogen	4.8
Nitrogen	1.4
Sulfur	3.2
Chlorine	0.1
Ash	11.5
Oxygen (by difference)	8.0
Total	100.0
Moisture %	8.6

Table 6.4

Product Size from Coal Cleaning Plant

Base Case		Wt %
Particle Size		
	3 x 1-1/2	9.4
	1-1/2 x 100M	90.6
Moisture		8.6
Ash		11.47

6.1 Process Description, Block Flow Diagram and Process Flow Diagram

Jigs effectively remove mine dilutions and other high ash components from ROM coal. They are extensively used by the coal industry for the preparation of power plant fuel. Depending upon the characteristics of the ROM coal, Jigs can dramatically improve its quality at a very nominal cost.

Jigs are operated at a relatively high specific gravity of separation of 1.8. The washability characteristics of the raw coal and the desired ash/sulfur contents in the clean coal generally determine the specific gravity of separation. A lower ash content in the clean coal usually requires a lower separation gravity. The separation efficiency of Jigs is low below the specific gravity of 1.6. This reduced efficiency is reflected in lower than theoretically possible yield of clean coal and/or increased clean coal ash content.

Figure 6.1 presents a Simplified Block Flow Diagram for the Base Case. ROM coal (3 inch x 0) from the storage pile is cleaned in a Baum Jig at a separation gravity of 1.8 providing a float (clean coal) and a coarse refuse. A clean coal screen classifies the coal into three size fractions. The coarse size fraction (3 inch x 1-1/2 inch x 1/8 inch) is dewatered using a centrifuge. The fine clean coal slurry (1/8 inch x 0) is classified at 100 mesh using classifying cyclones. The cyclone underflow joins the 1-1/2 inch x 1/8 inch coal ahead of the centrifuge. The cyclone underflow slurry containing high ash clay solids is diverted to a static thickener where it is thickened to a consistency of approximately 25 percent (solids by weight). The thickened fine refuse is pumped to a settling pond. Clarified water from the thickener overflow is reused in the plant. The crushed and dewatered clean coal is sent to the clean coal storage facility. The dewatered coarse refuse from the Jig is transported to the Refuse Handling Section.

The more detailed description of the plant is shown in Figure 6.2 (process flow diagram for Jig cleaning) which also includes product crushing, dewatering and water recovery.

FIGURE 6.1

JIG CLEANING - BASE CASE

SIMPLIFIED BLOCK FLOW DIAGRAM

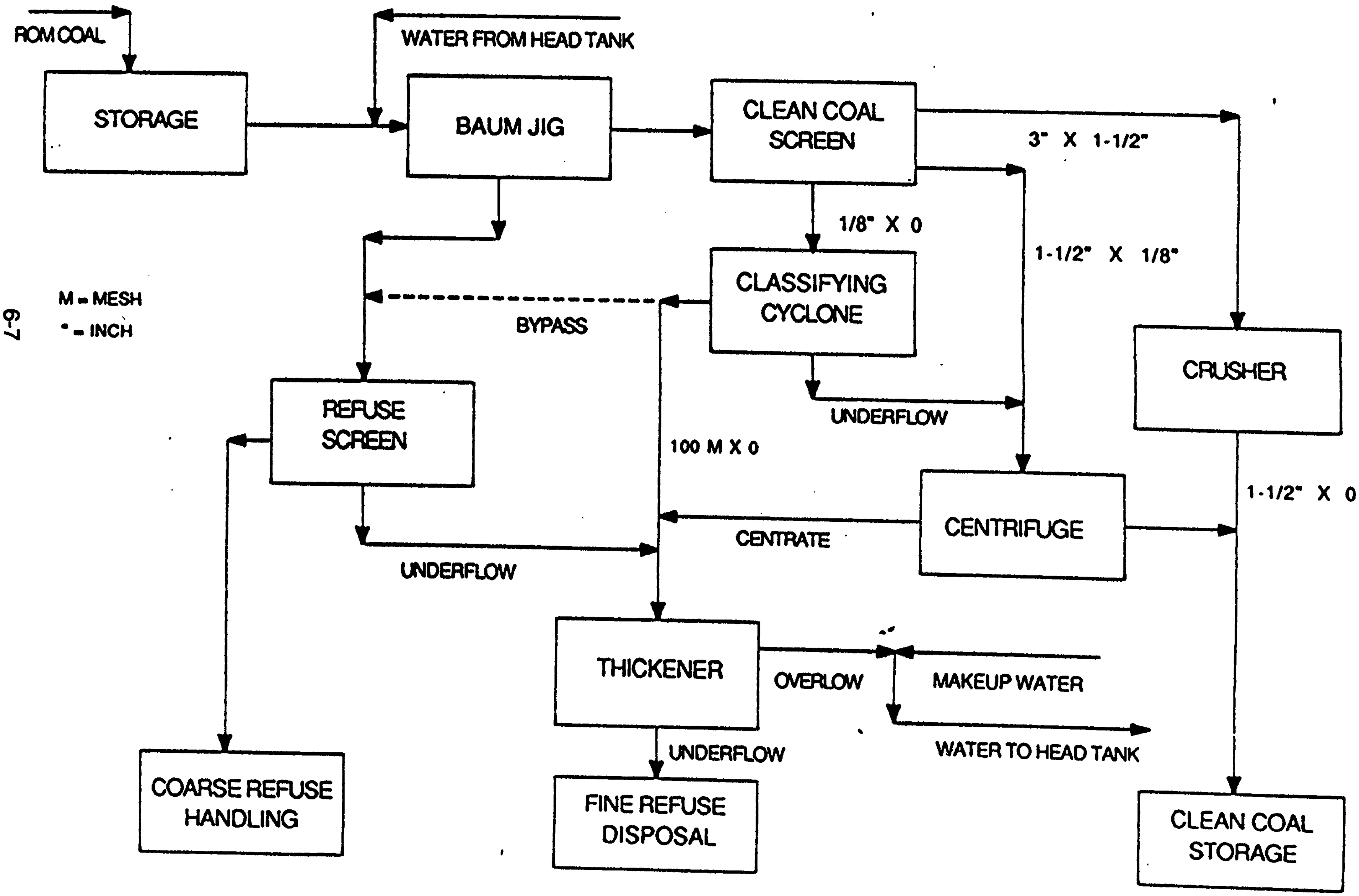
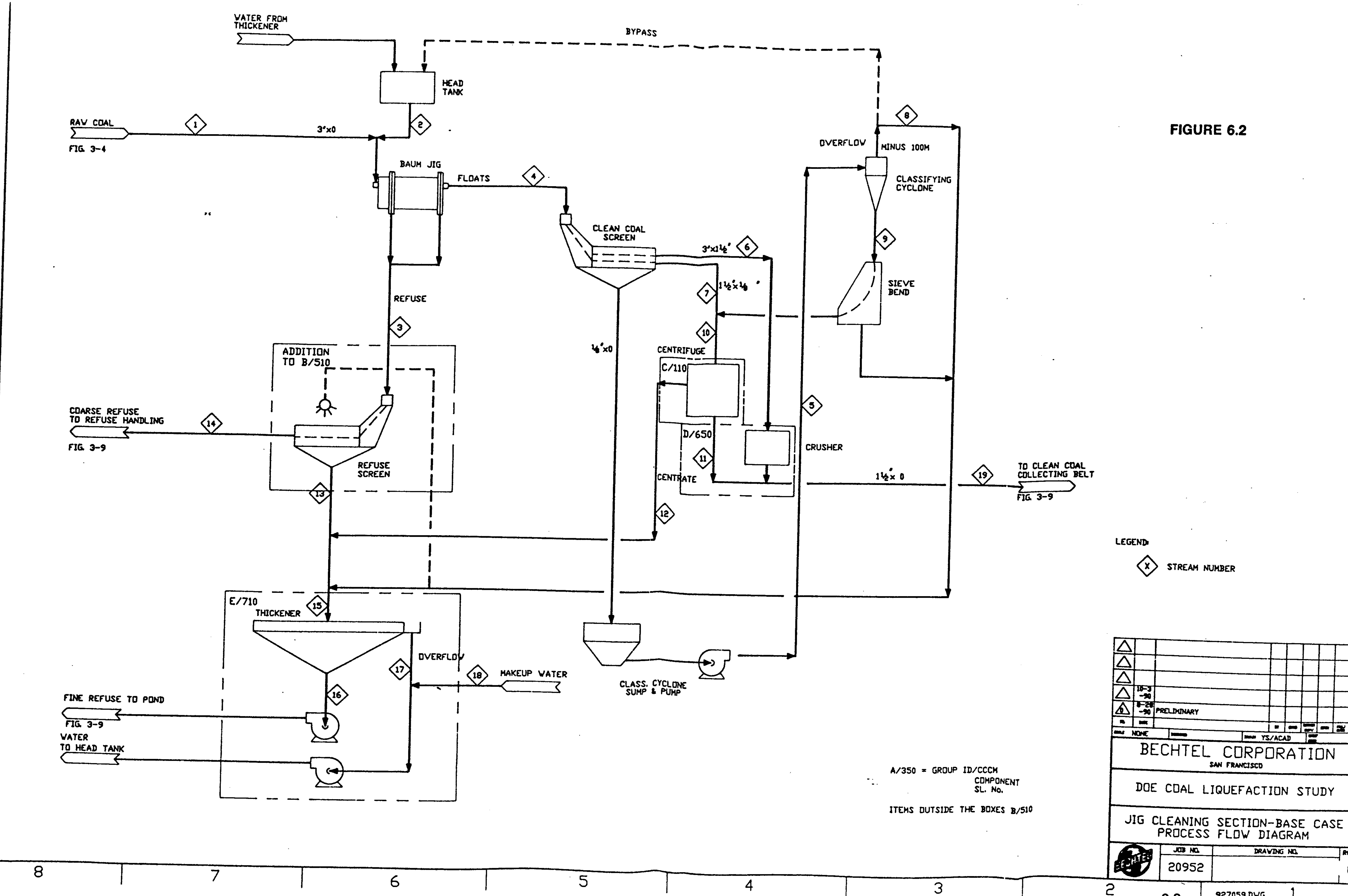


FIGURE 6.2


WATER FROM THICKENER
BYPASS
HEAD TANK
RAW COAL
FIG. 3-4
3"x0
BAUM JIG
FLOATS
REFUSE
CLEAN COAL SCREEN
3"x1½"
1½"x⅛"
⅛"x0
OVERFLOW
MINUS 100M
CLASSIFYING CYCLONE
SIEVE BEND
ADDITION TO B/510
COARSE REFUSE TO REFUSE HANDLING
FIG. 3-9
REFUSE SCREEN
CENTRIFUGE
C/110
D/650
CENTRATE
CRUSHER
1½"x 0
TO CLEAN COAL COLLECTING BELT
FIG. 3-9
LEGEND:
STREAM NUMBER
E/710
THICKENER
OVERFLOW
MAKEUP WATER
FINE REFUSE TO POND
FIG. 3-9
WATER TO HEAD TANK
CLASS. CYCLONE SUMP & PUMP
A/350 = GROUP ID/CCCM COMPONENT SL. No.
ITEMS OUTSIDE THE BOXES B/510
10-3-90
8-28-90 PRELIMINARY
NONE
YS/ACAD
BECHTEL CORPORATION
SAN FRANCISCO
DOE COAL LIQUEFACTION STUDY
JIG CLEANING SECTION-BASE CASE PROCESS FLOW DIAGRAM
JOB NO.
20952
DRAWING NO.
REV
0
927059.DWG
8
7
6
5
4
3
2
1

The coal receiving, storage and mixing section of the plant incorporates a storage and homogenization facility. Additional items are shown in a separate box. For this plant a circular storage system with a bridge type scraper reclaimer will be used.

Figure 6.3 represents the process flow diagram for this section of the plant.

ROM coal will be delivered from the mine to the preparation plant by a belt conveyor. The coal will be transferred to a storage feed conveyor which will deliver the coal to the circular or longitudinal pile(s).

The circular storage system will use a stacker with slewing and luffing capabilities mounted on a column at the center of the storage pile. The column supports the stacker which slowly rotates and deposits the coal in the storage area in thin layers. A bridge scraper reclaimer will reclaim the coal to the center of the storage pile area where it will be transferred through a central hopper to an underground reclaim conveyor. Additional reclaim hoppers and feeders will be provided below the pile.

The longitudinal systems will use a traveling stacker mounted on rails. The stacker, with luffing and slewing capability, can build coal piles on either side of the track. The coal will be laid in thin layers as the stacker moves slowly up and down the rail track. Reclaiming will be done by a bridge mounted bucket wheel reclaimer. As the reclaimer advances into the pile it will take a cut across the entire cross section of the pile thus ensuring a well blended feed to the cleaning plant.

Coal from the reclaim conveyor will be conveyed to a surge bin which will ensure a steady feed for the downstream units. A tramp iron magnet will be provided to collect tramp magnetic metal found in the plant feed.

A weigh feeder located below the surge bin will deliver measured quantities of raw coal to the plant feed conveyor for transport to the cleaning plant building.

The following description addresses a single train.

ROM coal, 3 inch x 0 in size, will be fed into a Baum Jig for a separation at a specific gravity of 1.8. The jig will produce a float and a sink product. The float product (clean coal) will be dewatered on a double deck vibrating screen with decks of 1-1/2 inch and 1/8 inch openings. The screen underflow will be pumped to a bank of classifying cyclones. The cyclone overflow slurry with minus 100 mesh solids which are essentially high ash-clay solids will be sent to the Refuse Thickening and Water Recovery section. The sink product from the Baum Jig will be the coarse refuse. This product will be dewatered on a refuse screen and sent to the Coarse Refuse Handling section shown in Figure 6.4.

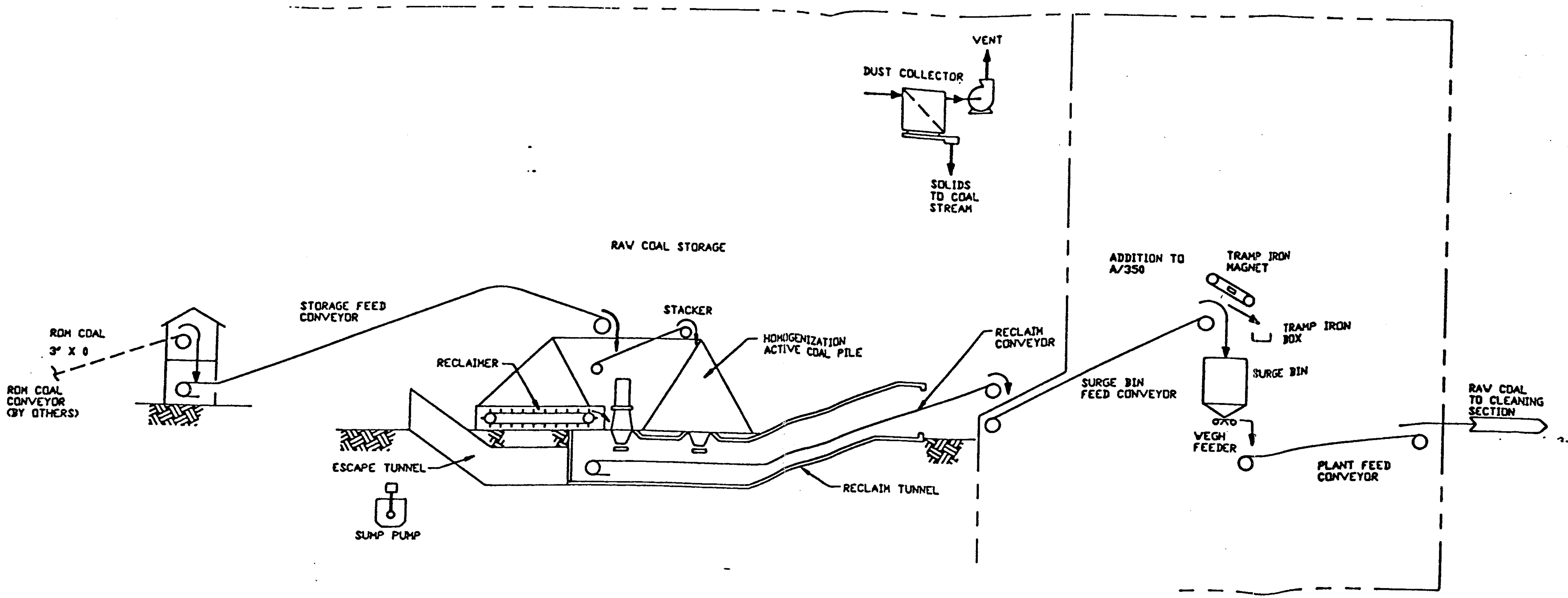


Figure 6.3

BECHTEL CORPORATION SAN FRANCISCO		
DOE COAL LIQUEFACTION STUDY		
ROM COAL RECEIVING AND STORAGE SECTION PROCESS FLOW DIAGRAM		
JOB NO.	DRAWING NO.	REV.
20952		0

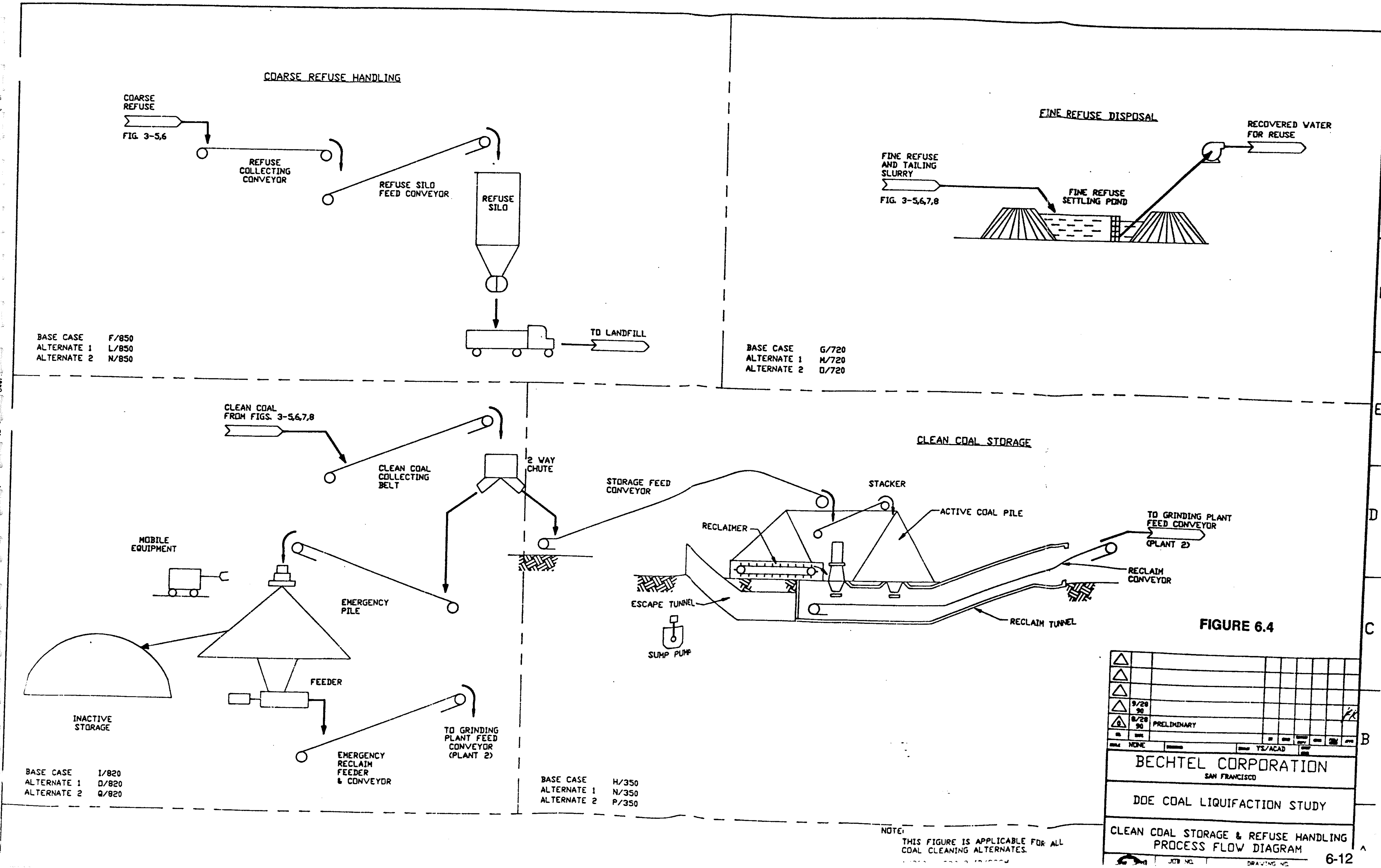

COARSE REFUSE HANDLING
COARSE REFUSE
FIG. 3-5,6
REFUSE COLLECTING CONVEYOR
REFUSE SILO FEED CONVEYOR
REFUSE SILO
TO LANDFILL
BASE CASE F/850
ALTERNATE 1 L/850
ALTERNATE 2 N/850
FINE REFUSE DISPOSAL
RECOVERED WATER FOR REUSE
FINE REFUSE AND TAILING SLURRY
FIG. 3-5,6,7,8
FINE REFUSE SETTLING POND
BASE CASE G/720
ALTERNATE 1 M/720
ALTERNATE 2 O/720
CLEAN COAL FROM FIGS. 3-5,6,7,8
CLEAN COAL COLLECTING BELT
2 WAY CHUTE
MOBILE EQUIPMENT
EMERGENCY PILE
INACTIVE STORAGE
FEEDER
TO GRINDING PLANT FEED CONVEYOR (PLANT 2)
EMERGENCY RECLAIM FEEDER & CONVEYOR
BASE CASE I/820
ALTERNATE 1 O/820
ALTERNATE 2 Q/820
CLEAN COAL STORAGE
STORAGE FEED CONVEYOR
STACKER
ACTIVE COAL PILE
RECLAIMER
TO GRINDING PLANT FEED CONVEYOR (PLANT 2)
RECLAIM CONVEYOR
ESCAPE TUNNEL
RECLAIM TUNNEL
SUMP PUMP
BASE CASE H/350
ALTERNATE 1 N/350
ALTERNATE 2 P/350
NOTE:
THIS FIGURE IS APPLICABLE FOR ALL COAL CLEANING ALTERNATES.
PRELIMINARY
NONE
TS/ACAD
BECHTEL CORPORATION
SAN FRANCISCO
DOE COAL LIQUIFACTION STUDY
CLEAN COAL STORAGE & REFUSE HANDLING PROCESS FLOW DIAGRAM
JOB NO.
DRAWING NO.

FIGURE 6.4

6.2 The overall material balance for Plant I is shown in Table 6.5. This material balance is per train.

6 TRAINS

Table 6.5

COAL CLEANING: BASE CASE (JIG)											
OVERALL MATERIAL BALANCE PER TRAIN FOR PLANT1											
		Material In			Material Out						
		Raw Coal	Make-up Water	Total In	Clean Coal	Clean Coal	Total Clean Coal	Refuse	Refuse	Total Refuse	Total Out
STREAM #		1	18		6	11	19	14	16	18	
Coal	TPH	627.6	0.0	627.6	47.1	455.0	502.1	90.8	34.7	125.5	627.6
Water	TPH	59.9	101.7	161.5	2.1	45.0	47.1	9.9	104.6	114.5	161.6
	GPM	239.6	406.9	646.5	8.4	180.0	188.4	39.8	418.5	458.2	646.6
Total Stream	TPH	687.5	101.7	789.1	49.2	500.0	549.2	100.7	139.3	240.0	789.2
Ash	TPH	137.5	0.0	137.5	5.4	52.2	57.6	59.9	20.0	79.9	137.5

NOTE: (1) Stream #1 Is Raw (MF) Coal Entering The Coal Cleaning Plant
(2) Stream #19 Is Clean Coal (MF) Leaving The Coal Cleaning Plant
(3) This Plant Operates Two Shifts A Day With 14.5 Hours On Stream Per Day, 5 Days A Week And 50 Weeks A Year
(4) Each Train Will Deliver Approximately 1.82 MM TPY Of Clean Coal (MF) To Storage

8.576% H_2O

6.3 MAJOR EQUIPMENT LIST

The major equipment list for the plant is included in Table 6.6.

TABLE 6.6

JIG CLEANING - BASE CASE

ROM Coal Receiving and Storage

Surge Bin Feed Conveyor

Tramp Iron Magnet

Surge Bin

Weigh Feeder

Plant Feed Conveyor

Baum Jig

Addition

Refuse Screen

Clean Coal Centrifugal Drying

Clean Coal Crushing

Fine Refuse Thickening and Water Recovery

Coarse Refuse Handling

Fine Refuse Disposal

Clean Coal Storage

Clean Coal Storage - Emergency File

6.4 UTILITY SUMMARY

TOTAL   AVG PER SD ? SEE TAB 5.1

Utility requirement for the plant ~~(per train)~~ is listed in Table 6.7.

TABLE 6.7

ELECTRIC POWER 8,289 KW

WATER 785 gpm ✓

7. Plant 1.4 - (Coal Grinding and Drying)

7.0 Design Basis Criteria and Considerations and Block Flow Diagram

This plant crushes and drys the coal from the coal cleaning and handling (Plant 1) before it enters the coal liquefaction plant (Plant 2).

Clean coal enters the plant sized at 1-1/2"x0 with up to 15% moisture content and exits (with 50% of the coal minus 200 mesh) and with 2% moisture content. The block flow diagram, Figure 7.1, illustrates the coal grinding and drying process. Moist clean coal from a feeder bin enters the coal mill where it is ground. Dry, heated nitrogen blown into the mill dries and carries the fine coal dust out of the mill to a dust collector unit. The dust collector separates the coal dust from the nitrogen and directs the coal to a crushed coal bin. The nitrogen is recirculated through a heater back into the mill except for a small amount which is purged. The crushed coal product is delivered to Plant 2 via a pneumatic type pump.

This plant is designed to grind and dry 17,102 TPD (dry) of 1-1/2" and less, 55 Hargrove coal with 15% maximum feed moisture content. The final product has a size of 50% of the coal less than 200 mesh and a 2% moisture content.

The plant is intended to be operated 24 hours per day.

Inert nitrogen gas is utilized for drying and transferring the coal dust to minimize auto-ignition and explosive hazards.

The coal feed and crushed coal product bins must each provide 8 hrs of storage capacity (600 tons at 75 TPH).

7.1 Plant Description

The design for this plant is based on coal roller-mills. Twelve identical grinding circuits or trains are required, 2 trains for each Coal Liquefaction (Plant 2) train. The trains are rated at 75 TPH each. Ten trains are normally operating with two spares corresponding to the spare Plant 2 train. The following paragraphs describe one of the trains (see Figure 7.1).

A weigh belt feeder (1.4-T101A-L), located below the 600 ton coal feed bin (1.4-D101A-L), delivers clean coal to the mill (1.4-Y101A-L). The coal is fed to the center of the mill's grinding table where it is ground between the rotating table and three large cylindrical rollers. The ground coal spills over the table's edge and is lifted by a stream of rising hot (660 °F) nitrogen gases to the top of the mill. The hot gases cause an immediate evaporation of the moisture in the coal. A separator at the top of the mill allows the finished coal dust to exit the mill while the coarse fraction falls back to the center of the grinding table for further crushing.

The coal dust leaving the mill dust enters the dust collector baghouse (1.4-Y102A-L). The baghouse directs the stream up through fabric bag filters which separate the coal particles from the nitrogen gas. Particles collect on the outside of the bags and are removed periodically with short bursts of pressurized plant air. The recovered particles are transferred by screw conveyor (1.4-T102A-L) to the 600 ton crushed coal product bin (1.4-D102A-L). The crushed coal is conveyed to Coal Liquefaction (Plant 2) via a pneumatic type feed pump (1.4-G101A-L).

The nitrogen gas enters the nitrogen recycle fan (1.4-K101A-L) where about 90% of the gas is recycled to the mill through a fuel gas fired heater (1.4-F101A-L). The hot nitrogen gas is tempered with make-up nitrogen from the nitrogen bleed damper (1.4-L101A-L) to about 660 °F before entering the mill. The portion of nitrogen gas which is not recycled is vented to the atmosphere through a final bag filter.

A feeder hoist (1.4-T103A-L) and overhead travelling crane (1.4-T104A-L) are included with each train for maintenance purposes.

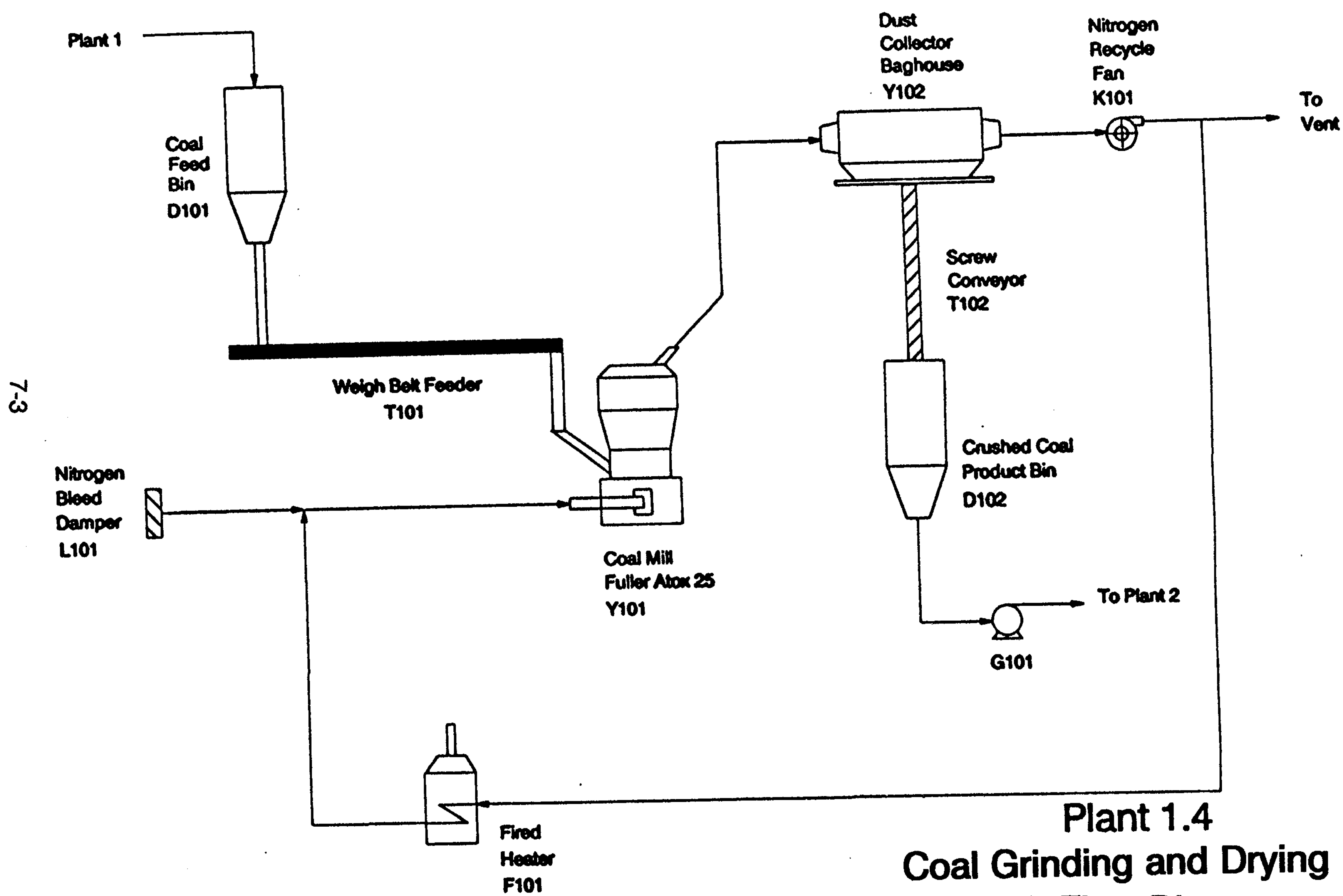


Figure 7.1

Plant 1.4
Coal Grinding and Drying
Block Flow Diagram

7.2 Major Equipment List

The major equipment for this plant is listed in Table 7.4.

Table 7.1
Major Equipment List

Equipment No.	Type
1.4-D101A-L	Coal Feed Bin
1.4-D102A-L	Crushed Coal Product Bin
1.4-F101A-L	Fired Heater
1.4-G101A-L	Finished Coal Feed Pump
1.4-K101A-L	Nitrogen Recycle Fan
1.4-L101A-L	Nitrogen Bleed Damper
1.4-T101A-L	Weigh Belt Feeder
1.4-T102A-L	Crushed Coal Screwed Conveyor
1.4-T103A-L	Hoist for Feeders
1.4-T104A-L	Overhead Travelling Crane
1.4-Y101A-L	Fuller Atox 25 Coal Mill
1.4-Y102A-L	Dust Collector Baghouse

7.3 Utility Summary

ENTIRE

The utility summary for this Plant is included in Table 7.2.

Table 7.2

TOTAL **Utility Summary**

		PER OPERATING TRAIN
Electricity	12114 KW	1211.4
Water Circulation	240 gpm	24.0
Fired Duty	486 MMBTU/hr	48.6
Nitrogen Consumed	108 MMSCFD	10.8
Air Consumed	8.6 MMSCFD	.86

8. Plant 2 (Coal Liquefaction)

8.0 Design Basis Criteria and Considerations

Coal Feed

The coal to be fed to the coal liquefaction plant is washed and dried Illinois No. 6 coal (Burning Star Mine) from Plant 1, the Coal Preparation Plant. Analysis of the basis coal is presented in Table 8.1.

5 TRAINS

Coal feed to Plant 2 is approximately:

Basis	Per Train	Total
MF basis	3,420 TPSD	17,102 TPSD
MAF basis	3,028 TPSD	15,140 TPSD

Moisture in the coal feed is 2.0 wt. % and ash (MF) content is 11.47 wt. %.

Process

The process used will be close-coupled, catalytic-catalytic, two-stage coal liquefaction (H-COAL by HRI) with ashy recycle, recycle of extract from the critical solvent deashing plant, and recycle of 850°F+ to extinction.

Maximum Reactor Size

The maximum size reactor will process a feed rate of 3,028 TPSD (MAF basis), producing 12,389 BPSD of liquid products (C_5-850°F). Five trains of the maximum size operating at capacity will be required to reach the proposed production of approximately 62,000 BPSD.

Nominal Reaction Conditions

Nominal coal liquefaction conditions will be:

Reactor Inlet Pressure, psig	3200
Reactor Outlet Pressure, psig	3000
Hydrogen Partial Pressure, psia (outlet of Stage 2 reactor)	1950
Coal Feed Rate (space velocity), lb MAF coal/lb catalyst/hr	1.12
Solvent/MAF Coal Ratio	2.454
Solvent, wt.%	
Distillate (1000°F-)	38.0
Residuum (1000°F+)	49.9
Ash + Unconverted Coal	12.1
Slurry Tank Temperature, °F	400
Reaction Temperature, °F	
Stage 1	790
Stage 2	760
Coal Feed Analysis	(Table 8.1)

Table 8.1

Analysis of Feed Coal to Liquefaction
Illinois No. 6 (Burning Star Mine)

Proximate Analysis (wt. %, Dry Basis)	
Volatile Matter	42.2
Fixed Carbon	46.3
Ash	11.5
Ultimate Analysis (wt. %, Dry Basis)	
Carbon	71.0
Hydrogen	4.8
Sulfur	3.2
Nitrogen	1.4
Ash	11.5
Chlorine	0.1
Oxygen (by difference)	8.0
Sulfur Forms (wt. %, Dry Basis)	
Pyrite	1.0
Sulfitic	0.1
Organic	1.9
Ash Composition (wt. % oxidized)	
Phosphorus pentoxide, P_2O_5	0.2
Silica, SiO_2	49.8
Ferric Oxide, Fe_2O_3	17.6
Alumina, Al_2O_3	19.2
Titania, TiO_2	1.0
Lime, CaO	6.3
Magnesia, MgO	1.0
Sulfur Trioxide, SO_3	2.9
Potassium Oxide, K_2O	2.0
Sodium Oxide, Na_2O	0.5
Undetermined	-0.5

Catalyst and Catalyst Addition Rates

Catalyst for the coal liquefaction process will be Amocat 1C in both stages. The catalyst has a bulk density of 35 lb/cf. Catalyst addition rates are specified to be 3.0 lb of catalyst per ton of MAF coal feed for the first stage reactor and 1.5 lb of catalyst per ton of MAF coal feed for the second stage reactor.

Hydrogen Makeup

Makeup hydrogen gas is available at the pressure required for the coal liquefaction plant at a temperature of 110°F. The composition of the makeup hydrogen is approximately 99.9% (vol.) hydrogen and 0.1% (vol.) nitrogen.

Yields

Yields from the coal liquefaction plant are presented in Table 8.2 below:

Table 8.2

Coal Liquefaction Yields

Yields *Wt.% MAF Coal Feed	First Stage Outlet	*Overall	
H_2S	2.00	2.86	
H_2O	7.10	9.51	
NH_3	0.95	1.39	
CO	0.04	0.06	
CO_2	0.09	0.14	
C_1	0.91	1.84	
C_2	0.74	1.43	
C_3	0.85	1.52	
C_4	0.40	0.79	
C_5-350°F	8.00	16.12	
350-450°F	4.90	7.51	
450-650°F	6.76	25.11	
650-850°F	15.20	21.36	
850-1000°F	18.52	0.66	
1000°F+	26.80	8.39	
Unconverted Coal	10.20	7.21	
Phenols	0.24	0.30	106.2
Ash	12.96	12.96	
Hydrogen Consumption	-3.70	-6.20	

* At the outlet of 2nd stage

Product Quality

The API gravity of the product cuts from the first stage and second stage reactors is given in Table 8.3 below:

Table 8.3

Product Quality

Product Cuts	Gravity (°API) First Stage	Gravity (°API) Overall
C_5-350°F	44.5	45.5
350-450°F	24.5	25.0
450-650°F	13.5	14.0
650-850°F	5.0	6.0
850-1000°F	-1.6	-1.5
1000°F+	-11.5	-10.5

Design Considerations

The design of the slurry preparation system of the Coal Liquefaction Plant was based on the concept used in the Breckinridge Project.

The capacity of an individual train was determined by HRI and it was based on reactor diameter and weight considerations. The coal throughput using the maximum-sized reactors was based on the space velocity used in Wilsonville run 257E. The liquefaction of each section including primary separation was designed by HRI. The HRI design report is included in Section 8.6.

The process, except the reaction system and slurry and hydrogen preheat system, was modeled using Simulation Science's PROCESS simulation software. The model was used to develop equipment sizing, product separations, and utility requirements for those portions of the plant not provided by HRI. Sizing information supplied by HRI was used directly in the study. This included the slurry and hydrogen preheat system, the reactor system, the high pressure separations system, and the recycle hydrogen compressor.

The material balance was developed from the information provided by HRI and extended to the rest of the plant. The overall material balance for the plant is shown in subsection 8.2, Table 8.4.

The process was developed making maximum use of air-fin coolers and condensers to a process outlet temperature of 130°F. The cooling of hot process fluids by steam generation was utilized wherever possible. Most of the steam generated was at the 150 psig level.

The separation system was developed with a two pressure level configuration. Each pressure had three separators: hot at the reactor effluent temperature (760°F), warm at approximately 550 °F, and cold at 130°F. The three separators configuration improved the purity of the recycle hydrogen stream, reduced the amount of heavy material in the gas stream sent to Plant 3, and reduced the amount of gases sent to atmospheric fractionation.

The coil outlet temperatures of the atmospheric and vacuum heaters was maximized to tube coking limitations. The pressures in the two fractionation towers was set at the lowest level practical by overhead condensing and vacuum system limitations in order to achieve as high a bottoms cut point as possible.

Two separate flush oil systems have been incorporated in the design. The light flush oil is taken from the atmospheric tower sidestream and is used for flushing instruments in slurry service. The heavy flush oil is taken from the vacuum tower upper sidestream and is used for flushing the seals of pumps in slurry service.

8.1 Process Description and Process Flow Diagrams

The plant consists of three sections. These are: Slurry Preparation and Liquefaction Reaction, Primary Separation and Product Fractionation. These three sections are schematically shown in process flow diagrams, Figures 8.1 through 8.3, respectively.

<u>Slurry Preparation and Liquefaction Reaction</u>

Coal, which has been pulverized and dried in Plant 1.4, is mixed with recycled oil from the downstream plants to form a slurry for feed to the coal liquefaction reactors. The recycle oil used as solvent for the process comes from four sources: slurry oil from the Slurry Hold Tank, slurry oil from atmospheric bottoms, the lower sidestream product (850-1000°F) from the Vacuum Distillation Tower, and the extract product from the Critical Solvent Deashing Plant (ROSE-SM), Plant 8.

Slurry preparation consists of prewetting and mixing. The prewetting occurs in a twin screw mixer in which the 850-1000°F product, which have been cooled in air-fin exchanger (E102) to 180°F, are sprayed on the pulverized coal as it is being turned over in the mixer. This mixture is then fed into the Mix Tank (C101) where it is mixed with the solvent from the other three sources. Those oils are cooled in an exchanger producing 600 psig steam (E101) before entering the Mix Tank. The Mix Tank is equipped with a high-speed agitator. Coal slurry flows to the Slurry Surge Tank

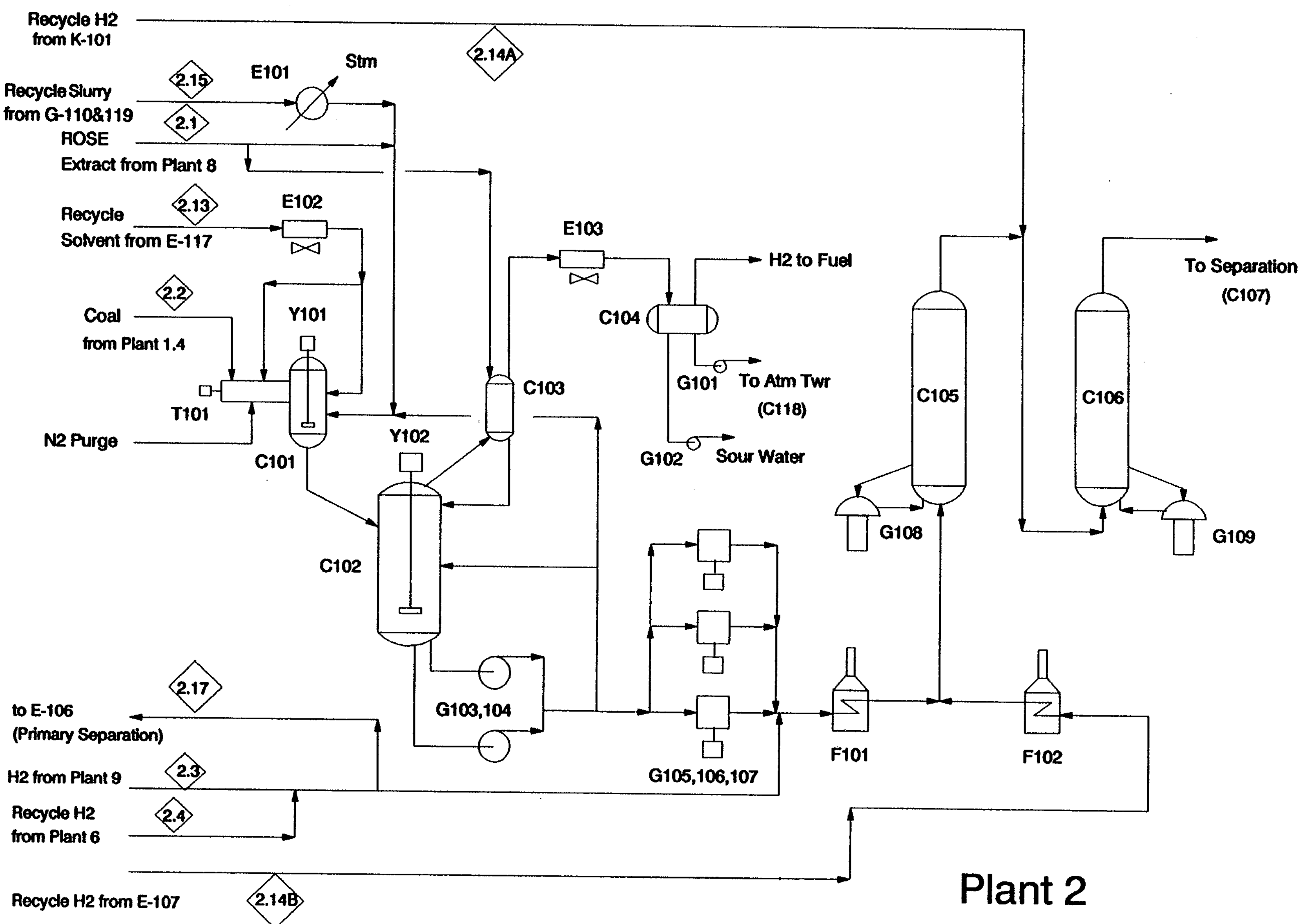
(C102) which also contains a mixer to keep the coal in the slurry. The Surge Tank is vented through a scrubber (C103), where the vapors are contacted with a portion of the solvent. The vapors from the scrubber are cooled in an air-fin exchanger (E103) to 130°F and separated into three phases in the Slurry Overhead Receiver (C104). The vapor from that drum is sent as fuel to the Hydrogen Heater (F102). The hydrocarbon liquid is sent to the Atmospheric Tower. The sour water stream is withdrawn to the Sour Water Flash Drum (C116) and eventually to Plant 34.

Material from the Slurry Surge Tank is pumped by two Slurry Booster Pumps (G103/G104) operating in parallel to the suction of three high pressure reciprocating Reactor Feed Pumps (G105/G106/G107) in parallel. About one-third of the flow from the Slurry Booster Pumps is recycled back to the Slurry Surge Tank to help eliminate dead spots in the tank where coal can settle out of the slurry. The Reactor Feed Pumps move the coal slurry through the Slurry Feed Heater (F101) to the reactors.

As shown on the attached Figure 8.1, the make-up hydrogen stream from coal gasification plant (Plant 9) is combined with the recycle hydrogen stream from Plant 6. A portion of the combined stream is then mixed with the recycle hydrogen stream from the Recycle Hydrogen Compressor (K101) and is first preheated by exchange against the reactor effluent in exchangers E106 and E107. The other portion of the combined hydrogen stream is injected into the coal slurry feed at the inlet to the Slurry Feed Heater (F101) in order to reduce the possibility of coke formation in the heater. Injection of hydrogen into the coal feed slurry also tends to reduce the viscosity of the slurry mixture and results in a lower pressure drop in the slurry feed heater coil. The hydrogen and coal slurry mixture is heated and then fed to the First Stage Reactor (C105). The remaining reactor hydrogen stream is fed to the First Stage Reactor after being heated in the Hydrogen Heater (F102).

The coal slurry-hydrogen mixture from the Slurry Feed Heater and the hydrogen stream from the Hydrogen Heater are introduced into the bottom of the First Stage Reactor. The reactor operates at approximately 790°F average reactor temperature and 3050 psig. The reactor operating temperature is controlled by adjusting the Slurry Feed heater outlet temperature to achieve the desired conversion level in the reactor.

Effluent from the First Stage Reactor is quenched with cold hydrogen from the recycle hydrogen compressor and introduced into the bottom of the Second Stage Reactor (C106). The interstage quench hydrogen flow rate is controlled to maintain the Second Stage Reactor in heat balance at 760°F average reactor temperature.



Plant 2
Slurry Preparation / Reaction

Figure 8.1 Process Flow Diagram

The reactors incorporate the principle of the ebullated-bed operation. The entire mass in the reactor is held in a fluidized ebullated state by recirculating coal-oil slurry from the top of the reactor through the Ebullating Pumps (G108 and G109) and back into the bottom of the reactor. The ebullated bed ensures the reactor's near isothermal operation under the extreme exothermic reactions involved. Since the coal particles are much finer than the extrudate catalyst, a separation can be made between these solids such that the coal and ash particles are entrained with the liquid-gaseous reactor effluent products, while the catalyst remains behind in suspension in the reactor.

Primary Separation

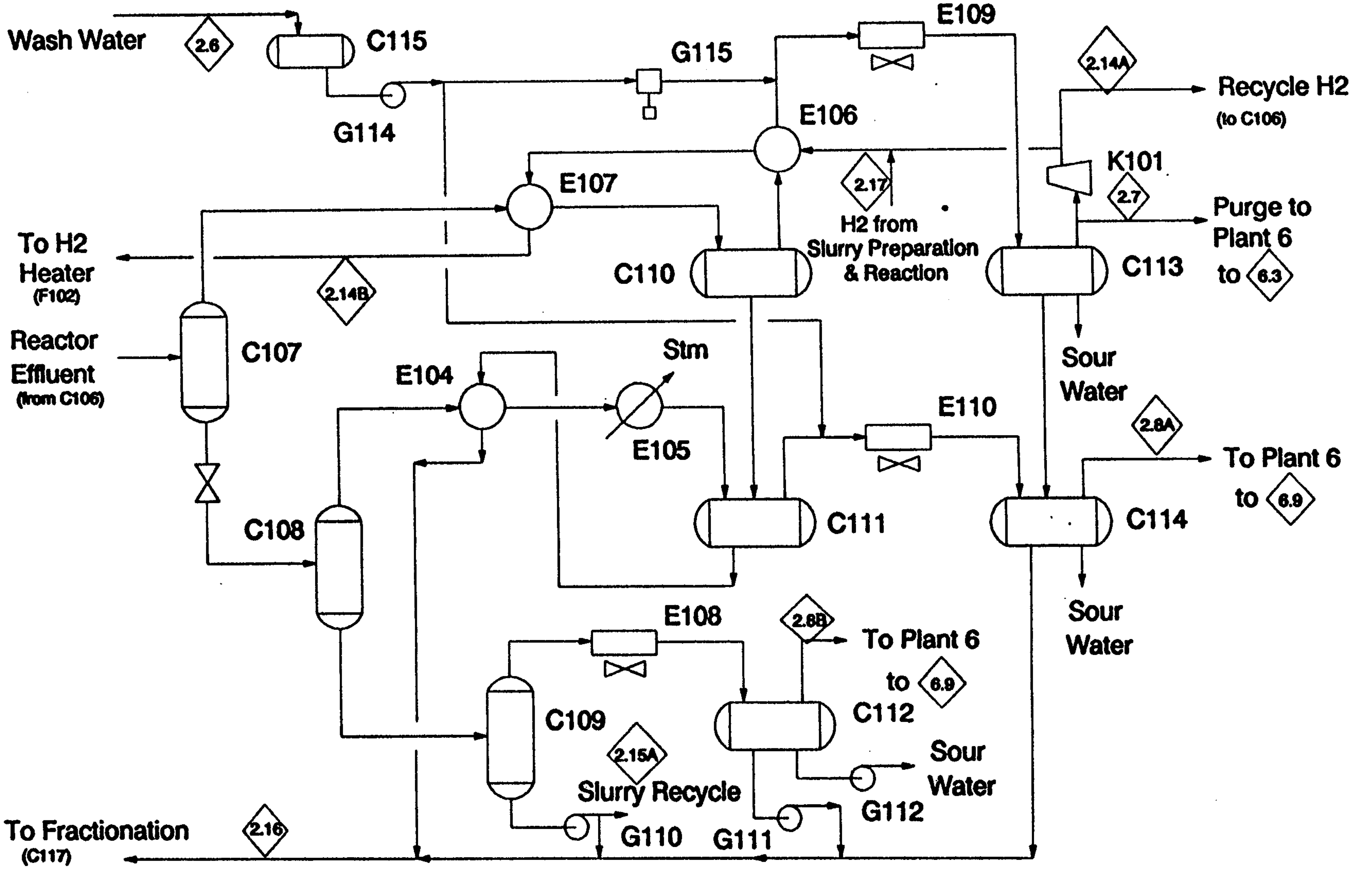
As shown on Figure 8.2, the effluent from the top of the Second Stage Reactor is sent directly to the Hot High Pressure Separator (C107) at 760°F and 3000 psig. The overhead vapor, after being separated from the slurry, is cooled to 550°F in exchange with recycle and makeup hydrogen in exchanger E107 and enters the Warm High Pressure Separator (C110). The temperature of the separator is set high enough to prevent precipitation of ammonium salts. The overhead of the warm separator is cooled to 130°F in exchange with recycle and makeup hydrogen in exchanger E106 and in air-fin exchanger E109 before entering the Cold High Pressure Separator (C113). Wash water is injected ahead of the air-fin cooler for control of ammonium salt deposition as the vapor is cooled. The vapor from that separator is compressed in the Recycle Hydrogen Compressor (K101) and returned to the reactors as recycle hydrogen. A portion of the stream is purged from the system to the Hydrogen Purification Unit (Plant 6) to prevent the build-up of methane and other non-condensables in the system. The water phase is withdrawn to the Sour Water Flash Drum (C116) and eventually to Plant 38 for recovery of anhydrous ammonia.

The liquid from the Hot High Pressure Separator is flashed to 100 psig in the Hot Low Pressure Separator (C108). The vapor from the Hot Low Pressure Separator is cooled from 750°F to 450°F in exchange with the Warm Low Pressure Separator liquid (E104) and in an exchanger producing 150 psig steam (E105). The mixed vapor-liquid stream then enters the Warm Low Pressure Separator (C111) along with the liquid from the Warm High Pressure Separator. The temperature of the separator is set high enough to prevent precipitation of ammonium salts. The vapor from the Warm Low Pressure Separator is cooled to 130°F in an air-fin cooler (E110) and enters the Cold Low Pressure Separator (C114) along with the liquid from the Cold High Pressure Separator. Wash water is injected ahead of the air-fin cooler for control of ammonium salt deposition. The vapor from this separator is sent to the Hydrogen Purification Plant, the liquid is sent to the Product Distillation Feed Flash Drum (C117), and the water phase is sent to the Sour Water Flash Drum (C116). The liquid from the Warm Low Pressure Separator is heated by the feed to that vessel in exchanger E104 and sent to the Product Distillation Feed Flash Drum. The liquid from the Hot Low Pressure Separator is further flashed to 20 psig in the Recycle Slurry Hold Drum

(C109). The vapor is cooled in air-fin exchanger E108 and sent to the Recycle Slurry Hold Drum Overhead Accumulator. The vapor from this drum is sent to Plant 6, the liquid is sent to the Product Distillation Surge Drum, and the water phase is sent to the Sour Water Flash Drum. Some of the liquid from the Recycle Slurry Hold Drum is pumped as recycle solvent back, using pump G110 to the coal slurry tank with the remainder being sent to the Product Distillation Surge Drum.

Product Fractionation

As shown in Figure 8.3, the liquid products from the three low pressure separators in the coal liquefaction plant are sent to the Product Distillation Feed Flash Drum (C117) from where the combined stream is pumped through the Atmospheric Feed Heater (F103) to the Atmospheric Distillation Tower (C118). Vapors from the surge drum are vented directly to the atmospheric tower. Two products are taken from the Atmospheric Tower Overhead Accumulator (C120): a naphtha product (IBP - 350°F) which is sent as feed to the Naphtha Hydrotreater via Plant 3, where Naphtha is used as lean oil; and a sour water stream which is sent to the Wash Water Surge Drum (C115) on the Coal Liquefaction Plant. An overhead vapor stream from the overhead accumulator is compressed by Compressor (K102) and sent to Plant 6. A sidestream product (350°F+) is withdrawn from a sidestream stripper (C119) and sent as feed to the Gas Oil Hydrotreater via cold (110°F) intermediate tankage. The bottoms stream off the atmospheric tower is pumped to two dispositions: as recycle solvent back to the Coal Slurry Tank with the remainder being sent to the Vacuum Feed Heater (F104) ahead of the Vacuum Distillation Tower(C122). Three products are taken overhead from the vacuum tower: a light gas oil product (450°F+) which is sent to the Gas Oil Hydrotreater via the same intermediate tankage as the atmospheric sidestream product; an overhead vapor stream which is compressed and sent to Plant 6; and a sour water stream which is sent to the wash water surge drum on the Coal Liquefaction Plant. An upper sidestream product (≈550°F - 850°F) is withdrawn and sent to the Gas Oil Hydrotreater via hot (400°F) intermediate tankage. A lower sidestream product (850°F - 1000°F) is withdrawn from a sidestream stripper (C123) and sent as recycle solvent back to the Coal Slurry Tank. The vacuum tower bottoms stream is sent to the solids-liquids separation unit, Plant 8.



Plant 2
Primary Separation

Figure 8.2 Process Flow Diagram

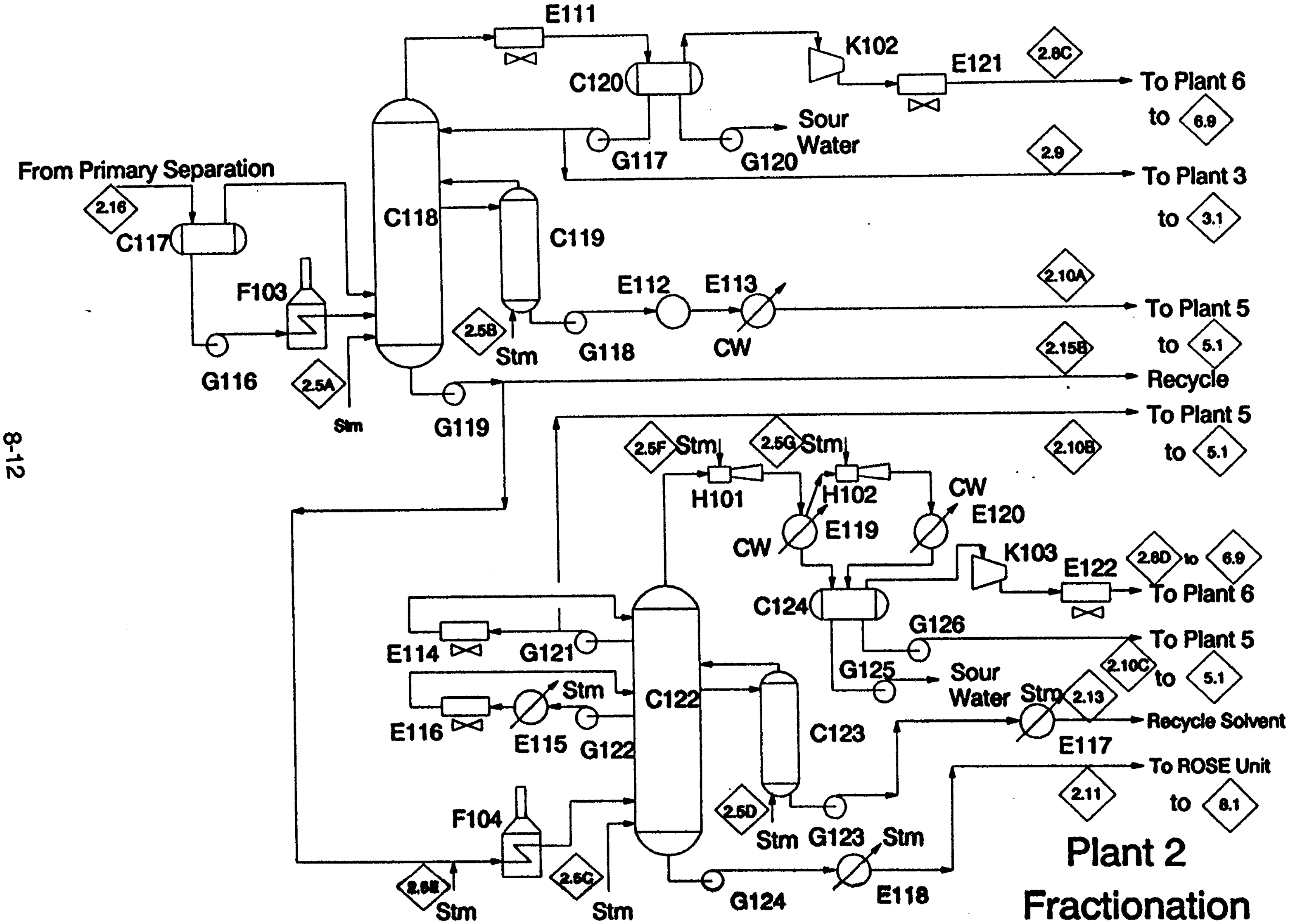


Figure 8.3 Process Flow Diagram

8.2 Material Balance

Plant 2 consists of 5 operating trains. The material balance per train is shown in Table 8.4. The overall material balance per train for the Plant 2 is schematically shown in Figure 8.4.

TABLE 8.4
PLANT 2
MATERIAL BALANCE

+5=99930

PLANT INPUT

STREAM NO.	2.2	2.3	2.4	2.6	2.5	2.1		
COMPONENTS	COAL #/hr	MAKEUP H2 #/hr	RECYCLE H2 FR #6 #/hr	WASH WATER #/hr	STEAM #/hr	ROSE EXTRACT #/hr	RXN DELTA #/hr	TOTAL INPUT #/hr
N2		223	70					293
H2		15900	4086				-15645	4341
H2O	5817		102	165232	49442		23997	244589
H2S			2		-		7217	7218
NH3			0				3507	3507
CO			42				151	193
CO2			0				353	353
C1			1501				4643	6144
C2			862				3608	4470
C3			668				3835	4503
C4			291				1993	2284
C5-350			171				40676	40847
350-450			1				18950	18952
450-650			0				63361	63361
650-850			0			76	53898	53974
850-1000			0			353	1665	2018
1000+			0			107317	21171	128488
PHENOLS			0				757	757
MEA			1					1
UNCONVERTED COAL	252333		0				-234140	18193
ASH	32702		0					32702
TOTAL	290852	16123	7794	165232	49442	107746	0	637189
TEMPERATURE (F)		100	100	126	*	300		
PRESSURE (psig)		3425	3425	3200	*	100		

PLANT OUTPUT

STREAM NO.	2.7	2.8	2.9	2.10	2.11	2.12	
COMPONENTS	PURGE H2 #/hr	PURGE GASES #/hr	NAPHTHA #/hr	GAS OIL #/hr	ROSE UNIT FEED #/hr	SOUR WATER #/hr	TOTAL OUTPUT #/hr
N2	205	88					293
H2	3442	899					4341
H2O	307	967	486			242829	244589
H2S	2022	2042	23			3131	7218
NH3	483	40				2984	3507
CO	135	58					193
CO2	173	80				100	353
C1	3623	2518	2			1	6144
C2	2007	2436	27				4470
C3	1315	3032	156				4503
C4	161	1781	342				2284
C5-350	85	5349	34227	1098		88	40847
350-450	0	54	631	15858		2409	18952
450-650		5		63354		2	63361
650-850				53898	76		53974
850-1000					2019		2019
1000+					128488		128488
PHENOLS						757	757
MEA		1					1
UNCONVERTED COAL					18193		18193
ASH					32702		32702
TOTAL	13958	19350	35894	134208	181478	252301	637189
TEMPERATURE (F)	130	130	140	110	688	**	
PRESSURE (psig)	2970	50	1	**	13	**	

Table 8.4 Material Balance - continued

Notes:

* *This stream represents the combination of steam levels utilized by the plant.*

** *This stream represents a combination of output streams; a single temperature/pressure cannot be reported.*

-- *The combined sour water flow rate for the plant is represented by 2.12*

2.5 = 2.5A+2.5B+2.5C+2.5D+2.5E

2.8 = 2.8A+2.8B+2.8C+2.8D

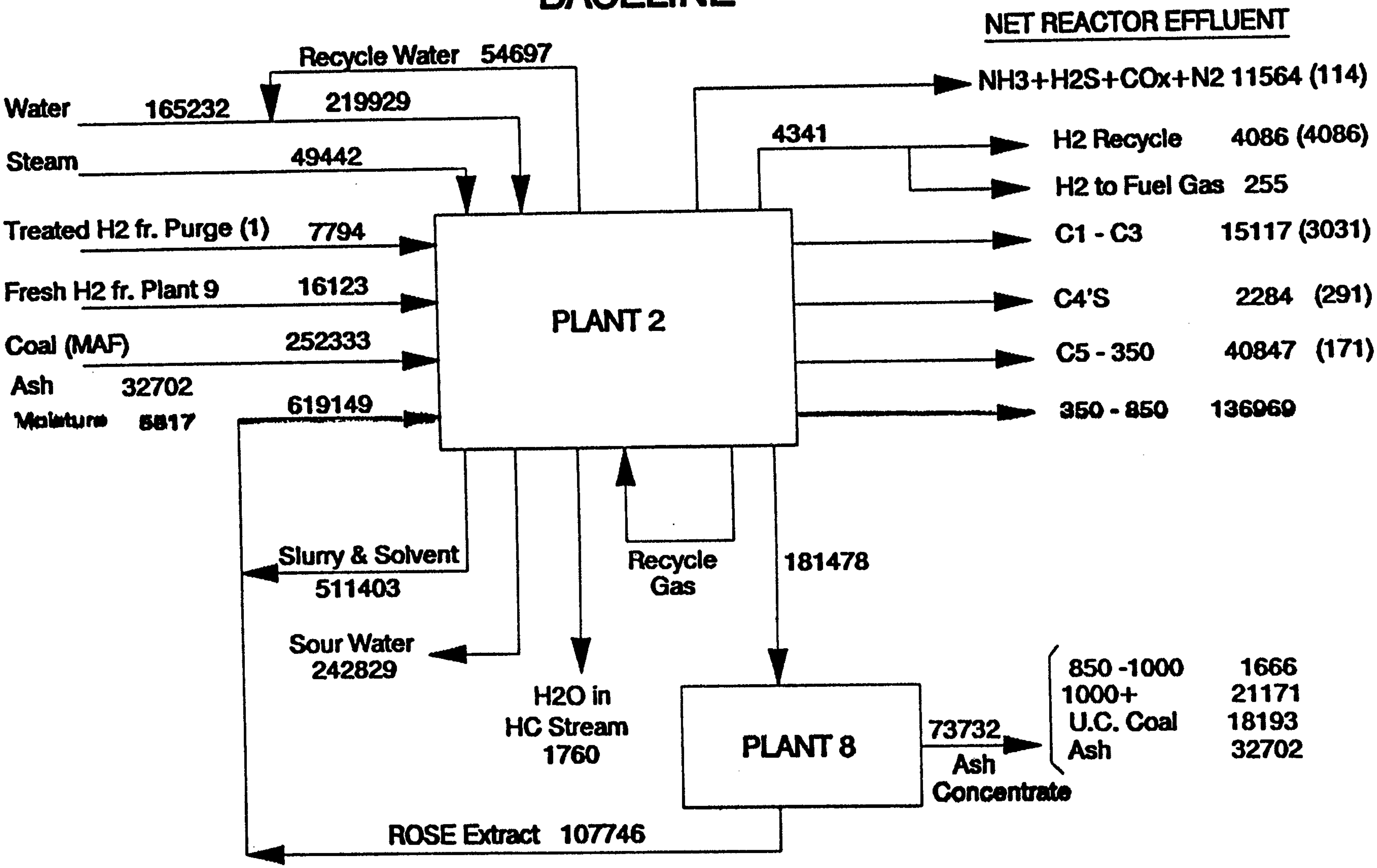
2.10 = 2.10A+2.10B+2.10C

2.14 = 2.14A+2.14B

2.15 = 2.15A+2.15B

These stream numbers are with reference to PFD of Plant 2.

BASELINE



NOTES: . Flow rate in lbs/hr
. Number in parenthesis is due to input stream (1). Therefore, to calculate the % yield subtract out the corresponding number in parenthesis.
. The HC's in sour water are shown in net reactor effluent.

Figure 8.4 Overall Material Balance

8.3 Major Equipment List

The major equipment list for Plant 2 is shown in Table 8.5.

Table 8.5 Major Equipment List

PLANT 2 — COAL LIQUEFACTION

Reactors and Vessels

Equipment No.	Equipment Description
2.1- C101	Slurry Vortex Mx Tnk Innr Shll
	Slurry Vortex Mx Tnk Outr Shll
2.1- C102	Slurry Surge Tank
2.1- C103	Slurry Srge Tank Vent Scrubber
2.1- C104	Scrubber Overhead Receiver
2.3- C105	Coal Liquefaction – Stage 1
2.4- C106	Coal Liquefaction – Stage 2
2.5- C107	Hot High Pressure Separator
2.5- C108	Hot Low Pressure Separator
2.5- C109	Recycle Slurry Hold Drum
2.5- C110	Warm High Pressure Separator
2.5- C111	Warm Low Pressure Separator
2.5- C112	Recy Slry Hold Drm Cond Accum
2.5- C113	Cold High Pressure Separator
2.5- C114	Cold Low Pressure Separator
2.5- C115	Wash Water Drum
2.5- C116	Sour Water Drum
2.5- C117	Compressor Knockout
2.6- C117	Feed Flash Drum
2.6- C120	Overhead Accumulator
2.6- C121	Vent Gas Compress Knockout
2.6- C124	Vacuum Jet Condenser Accum
2.6- C125	Vent Gas Compress Knockout
2.6- C118	Atmospheric Fractionator
2.6- C119	AGO Stripper
2.6- C122	Vacuum Fractionator
2.6- C123	Vac Recy Solvent Stripper

Fired heaters

Equipment No.	Equipment Description
2.2- F101	Coal Slurry Heater
2.3- F102	Gas Heater
2.6- F103	Atmospheric Fractionator Htr
2.6- F104	Vacuum Fractionator Preheater

Table 8.5 Major Equipment List - continued

Pumps

Equipment No.	Equipment Description
2.1- G101	Scrubber Condenser Product Pump
2.1- G102	Scrubber Sour Water Pump
2.1- G103	Coal Slurry Booster Pump
2.3- G105	Slurry Charge Pump
2.3- G108	Stage 1 Ebullating Pump
2.4- G109	Stage 2 Ebullating Pump
2.5- G110	Slurry Recycle Pump
2.5- G111	Recy Slr Hld Drm Cnd Acm Liq Pump
2.5- G112	Recy Slr Hld Dr Cnd Ac Sr Wtr Pump
2.5- G113	Sour Water Drum Pump
2.5- G114	Wash Water LP Pump
2.5- G115	Wash Water HP Pump
2.6- G116	Atmos Heater Charge Pump
2.6- G117	Reflux/Naphtha Product Pump
2.6- G118	AGO Product Pump
2.6- G119	Bottoms Product Pump
2.6- G120	Atmos Ovhd Accum Sour Water Pump
2.6- G121	LVGO Pumparound Pump
2.6- G122	HVGO Pumparound Pump
2.6- G123	HVGO Product Pump
2.6- G124	Bottoms Product Pump
2.6- G125	Vacuum Jet Accum Sour Water Pump
2.6- G126	Vacuum Jet Accum Product Pump

Miscellaneous

Equipment No.	Equipment Description
2.1- T101	Coal Prewetting Feeder
2.1- Y101	Slurry Vortex Mix Tank Mixer
2.1- Y102	Slurry Surge Tank Mixer
2.9/2.10	Catalyst Handling System
2.6- H101	1st Vacuum Steam Jet
2.6- H102	2nd Vacuum Steam Jet

Table 8.5 Major Equipment List - continued

Heat Exchangers

Equipment No.	Equipment Description
2.1- E101	Recycle Slurry Steam Generator
2.1- E102	Recycle Solvent Cooler
2.1- E103	Slurry Srge Tank Overhead Cond
2.5- E104	Hot LP Sep Vap Cooler
2.5- E105	Hot LP Sep Vap Steam Gen
2.5- E106	1st Hydrogen Preheater
2.5- E107	2nd Hydrogen Preheater
2.5- E108	Slurry Hold Drum Vap Cond
2.5- E109	Warm HP Sep Vap Cooler
2.5- E110	Warm LP Sep Vap Cooler
2.6- E111	Atmos Frac Ovhd Condenser
2.6- E112	AGO Product Cross Exchanger
2.6- E113	AGO Product Cooler
2.6- E121	Atm VG Comp Aftercooler
2.6- E114	LVGO Pumparound Cooler
2.6- E115	HVGO Pumparound Steam Gen
2.6- E116	HVGO Pumparound Cooler
2.6- E117	HVGO Product Steam Gen
2.6- E118	Bottoms Product Steam Gen
2.6- E119	1st Steam Jet Condenser
2.6- E120	2nd Steam Jet Condenser
2.6- E122	Vac VG Aftercooler

Compressors

Equipment No.	Equipment Description
2.7- K101	Recycle Hydrogen Compressor
2.6- K102	Atmospheric Vent Gas Compressor
2.6- K103	Vacuum Vent Gas Compressor

8.4 Utility Summary

Table 8.6 below presents a summary of the utilities required per train for the Coal Liquefaction Plant (Plant 2).

Table 8.6

PLANT 2 UTILITY SUMMARY

Steam, lb/hr		
600 psig	-28,705	
150 psig	-18,019	
50 psig	11,223	
Demineralized Water gpm	173	
Cooling Water, gpm	1,619	
Water, gpm	341	
Fuel Gas, MMSCFD	5	208.3 M BTU/hr
Electricity, Kw	11,798	
Nitrogen, MMSCFD	0.2	138 SCF/M

Note: Negative values represent utility production.

8.5 Water Summary

Table 5 below presents the water balances per train for the Coal Liquefaction Plant (Plant 2).

Table 8.7

WATER SUMMARY

Water Inlet	lb/hr
Coal feed	5,817
Produced in Reactors	23,997
Wash Water	219,929
Steam	49,442
Plant to Recycle	102
TOTAL	299,287
Water Outlet	
Sour Water to NH_3 Recovery	242,829
Recycled to Wash Water	54,697
Hydrogen Purge	307
Low Pressure Gases to Plant 3	967
Products to Plants 4 and 5	486
TOTAL	299,287
Wash Water Summary	
Requirements	219,929
Separator Water Recycled	54,697
Makeup Wash Water	165,232

8.6 HRI's Reactor/Primary Separation Design

Attached on the next pages is the complete HRI report on their portion of the Plant 2 design.

DOE DIRECT COAL LIQUEFACTION
BASELINE DESIGN AND SYSTEMS ANALYSIS

HRI REPORT

Prepared by

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Prepared For

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DISCLAIMER

This HRI Design Report for the DOE Baseline Design for Direct Coal Liquefaction presents information necessary for Bechtel to complete a preliminary conceptual commercial design of a two-stage catalytic coal liquefaction facility. The design information contained herein is not sufficient to proceed with the detailed design, procurement or construction for a commercial plant. No licensees, guarantees, or warrantees are contained in this report either written or implied

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DOE DIRECT COAL LIQUEFACTION

BASELINE DESIGN AND SYSTEM ANALYSIS

HRI SCOPE OF WORK

As part of Bechtel's project with the U.S. Department of Energy (DOE) for development of a direct coal liquefaction baseline design and systems analysis, HRI has performed the preliminary process design of a portion of the liquefaction section, under a subcontract with Bechtel. HRI's scope of work included the following portions of the liquefaction section:

- preheating
- first-stage reactor
- second-stage reactor
- high pressure separation
- recycle hydrogen compression

In addition to the baseline design described above, HRI prepared one alternative case based on an interstage separator, between the two reactor stages.

The baseline design is for a full-scale commercial facility, containing maximum-sized reaction trains, based on design basis information provided by Bechtel. The engineering work performed by HRI was preliminary in nature, but of sufficient detail for Bechtel to prepare a meaningful budget-type cost estimate. The engineering design is not adequate to proceed into detailed engineering nor to purchase equipment. A summary of work performed is as follows:

- Process Flowsheets were prepared showing major equipment and flows.
- Process Description is provided explaining the process flow through the steps identified within the battery limits.
- Overall Material Balance showing composition, conditions and rates of all battery limits process streams.
- Utility, Catalyst and Chemicals requirements were estimated.
- Equipment List and Process Data for the major equipment were prepared.

HRI's work on this project was based on our experience in the design of direct coal liquefaction plants and design basis information provided by Bechtel. It is important to note that HRI's scope of work did not include development of any design basis information. HRI did, however, review the design basis prepared by Bechtel and comment on the reasonableness of this basis.

BACKGROUND

Hydrocarbon Research, Inc. is the inventor, developer, and licensor of the ebullated-bed reactor, the H-Oil® Process, the H-Coal® Process and Catalytic Two-Stage Coal Liquefaction (CTSL™) Process. As part of the H-Coal Process development program, HRI scaled-up the H-Coal® Process from bench, to PDU (3 ton/day), to the Catlettsburg H-Coal Pilot Plant. The H-Coal Pilot Plant was designed to convert up to 600 tons of coal a day into as many as 1800 barrels of heavy fuel oil and distillate products or 675 barrels of synthetic crude from 200 tons a day. The plant was operated from 1979 to 1982. The demonstration program was very successful confirming scaleup of the H-Coal process chemistry, establishing process mechanical operability and reliability in commercial scale equipment, and collecting engineering design data on critical equipment such as slurry preheaters, let-down valves, and slurry pumps.

Based on the successful demonstration program at Catlettsburg, HRI participated with Ashland Oil and Bechtel in the complete design and engineering of a commercial-scale H-Coal plant producing 50,000 Barrels per day of synthetic crude product. HRI provided the process design "A" package for the H-Coal section including:

- coal slurry mixing and pumping
- preheat and reaction section
- product separation
- recycle slurry preparation
- product fractionation
- hydrogen compression and recycle

More than 35,000 manhours were expended by HRI to complete this design. As the price of oil fell from over 40$ per barrel to 25$ per barrel, the decision was made not to proceed with this commercial H-Coal project.

From 1982 to the current time, HRI has continued to conduct research and development on direct coal liquefaction. In 1983, HRI invented the direct-coupled two-stage catalytic coal liquefaction process (CTSL). This work was sponsored by the U.S. Department of Energy and cost shared by HRI.

The CTSL Process concept utilized a low temperature first stage (<800°F) for hydrogenation of coal feed and recycle slurry followed by a higher temperature (>800°F) hydroconversion second stage. Improvements were further made in optimizing recycle to extinct VGO and/or residuum produced from coal liquefaction. Slurry recycle rates were reduced from greater than 2.5 parts recycle per part coal to about 1.2/1 recycle to coal ratio. Deep coal cleaning was further demonstrated to increase liquid yields from coal.

Liquid yields were increased from 3 barrels per ton of moisture ash free coal to over 5 barrels per ton for the Illinois No. 6 bituminous coal. Product qualities were improved in that the distillate products were more aliphatic (more petroleum like) and had significantly reduced sulfur and nitrogen levels and improved compatibility and stability. HRI will be scaling up the CTSL™ Process concept during 1991 in HRI's 3 ton coal per day of Process Development Unit (PDU) in Lawrenceville, New Jersey.

HRI prepared conceptual commercial plant designs for the U.S. Department of Energy for CTSL processing of Illinois No. 6 bituminous coal and Wyoming sub-bituminous coal. Economics were prepared showing substantial reduction in the cost of producing liquid fuels from coal. Separate studies by Mitre confirmed that the CTSL Process as developed by HRI could produce liquid fuels from coal at lower cost than second generation technologies (EDS, H-Coal Process) or evolving third generation technologies (NTSL, ITSL, RITSL, CTSL). As stated by Dr. G.V. McGurl, DOE Program Manager, in the October 10, 1988 edition of Chemical & Engineering News- "of the several processes proposed thus far, the best is the catalytic-two-stage process developed by Hydrocarbon Research, Inc. He says it produces the highest yield of liquid product having the highest quality and does it at a lower cost than previous processes".

HRI also provided the process design for a catalytic ebullated-bed PDU for the Wilsonville Advanced Liquefaction facility. That unit now contains two-ebullated bed reactors in series and has been run in several modes (NTSL, ITSL, RITSL, CTSL) of direct coal liquefaction under sponsorship of the U.S. Department of Energy, Southern Company Services, Electric Power Research Institute, and Amoco. DOE has selected the team of Bechtel and Amoco to prepare a Baseline Design for Direct Coal Liquefaction based on a selected period of operation at Wilsonville in a direct-coupled catalytic-catalytic liquefaction mode. HRI has not participated in setting the operating conditions for the selected Wilsonville operations, has not evaluated the detailed technical data from this operation, has not done economic assessments as to the attractiveness of the selected operation, and has not developed the design basis for the Baseline Design. At the request of DOE and Bechtel, HRI has developed a preliminary conceptual process design for the high pressure section (preheaters to high pressure product separators) of a two-stage ebullated-bed catalytic reactor system for this Baseline Design.

The 1-1/2 inch x 1/8 inch clean coal separated on the clean coal screen will be fed to a centrifugal dryer for dewatering. The unit will also receive the underflow from the classifying cyclones. The underflow will be partially dewatered on a sieve bend ahead of the centrifuge.

The dewatered clean coal will be deposited on the Clean Coal Collecting Belt which is common to all process trains (Figure 6.4).

The 3 inch x 1/1/2 size fraction of the clean coal separated on the top deck of the clean dewatering screen will be crushed to a size below 1-1/2 inch and delivered to the Clean Coal Collecting Belt.

Fine Refuse Thickening and Water Recovery. The fine coal slurry overflow of the classifying cyclones will be collected in a static thickener. The clarified thickener overflow will be sent to a head tank for reuse. The underflow, the thickened fine refuse from the plant, will be pumped to the refuse settling pond (Figure 6.4).

DESIGN BASIS

The design basis for the baseline design was prepared by Bechtel and provided to HRI. It was based on operations at Wilsonville on Illinois No. 6 coal (Run 257E). These operations used a two-stage catalytic reactor configuration, and the catalyst in this run was 1/12-inch, Amocat-1C extrudates. The first-stage reactor was operated at 790°F, while the second-stage was operated at a lower temperature, 760°F. The total MAF coal feedrate was set at 15,140 tons per stream day to achieve a total liquid product rate of approximately 60,000 barrels per stream day.

The design basis information provided by Bechtel is summarized in Table 1.

TABLE 8.1

DESIGN BASIS INFORMATION

(Provided by Bechtel)

1) REACTOR OPERATING CONDITIONS

Coal Feed

Rate, lb MAF/hr - 1,261,667
% Ash (MF) - 11.47
% Moisture (as is) - 2.0

Temperature, °F

First-Stage - 790
Second-Stage - 760

Catalyst

Space Velocity, lb MAF coal/hr/lb catalyst - 1.12
Catalyst Replacement Rate, lb catalyst/ton MAF coal
First-Stage - 3.0
Second-Stage - 1.5
Catalyst Type - 1/12-inch, Amocat-1C extrudates
Catalyst Bulk Density, lb/ft^3 - 35

Hydrogen Partial Pressure

PSIA (min.) at second-stage outlet - 1950

2) YIELDS AND PRODUCT QUALITIES

Product Yields, W% Maf	First-Stage	Overall
H2S	2.00	2.86
H2O	7.10	9.51
NH3	0.95	1.39
CO	0.04	0.06
CO2	0.09	0.14
C1	0.90	1.84
C2	0.74	1.43
C3	0.85	1.52
C4	0.40	0.79
C5-350 °F	8.00	16.12
350-450 °F	4.90	7.51
450-650 °F	7.00	25.41
650-850 °F	15.20	21.36
850-1000 °F	18.52	0.66
1000 °F+	26.80	8.39

Unconverted Coal	10.20	7.21
Ash	12.96	12.96
Hydrogen Consumption	-3.70	-6.20
Product Quality (API Gravity)		
IBP-350 °F	44.5	44.5
350-450 °F	24.5	25.0
450-650 °F	13.5	14.0
650-850 °F	5.0	6.0
850-1000 °F	-1.6	-1.5
1000 °F+	-11.5	-10.5

3) **BATTERY LIMIT STREAM CONDITIONS**

Coal Slurry

Temperature, °F - 400
Solvent/MAF Coal Ratio - 2.457
Composition of Solvent, W%
    Distillate - 38.0
    1000 °F+ Resid - 49.9
    Unconverted Coal + Ash - 12.1
GC Simulated Distillation of Distillate, W%
    350-450 °F - 0.1
    450-650 °F - 7.5 → 18 %
    $\left.\begin{array}{l}\text{650-750 °F - 14.5}\\ \text{750-850 °F - 24.9}\end{array}\right\}$ 39.4   48
    $\left.\begin{array}{l}\text{850-950 °F - 44.8}\\ \text{950-1000°F - 8.2}\end{array}\right\}$ 53.0   vs 32 (nom)

Make-Up Hydrogen

Temperature, °F - 110
Composition, V%
    H2 - 99.0
    C1 - 1.0

Recycle Hydrogen

Temperature, °F - 130
Composition, V%
    H2 - 93.4
    C1 - 3.9
    C2 - 1.3
    C3 - 0.8
    C4 - 0.3
    C5+ - 0.3
    H2O - 0.1

SCHEME 1

BASE CASE DESCRIPTION

Based on the design basis information provided by Bechtel, HRI has completed the design of the base case, identified in this description as Scheme 1. This schemes contains two direct-coupled catalytic ebullated-bed reactors in series with the first stage operating at higher (790°F) temperature and the second stage at lower temperature (760°F). As a part of this design HRI has performed a heat and material balance, major equipment sizing, and estimated utility, catalyst and chemicals requirements. The following is provided to describe the base case:

- Process description
- Process flow diagram (Figure 1)
- Overall material balance (Table 2)
- Battery limits stream compositions and conditions (Table 3)
- Major equipment list (Table 4)
- Utility, catalyst and chemicals requirements (Table 5)

Five reactor trains are required to process the amount of coal specified in the design basis. The major equipment in each of the reactor trains is numbered with the series 100 to 500 for each reactor train. Each train is identical and processes 3,028 tons per stream day of MAF coal per train, or 15,140 tons per stream day total. The description which follows is for a single train (100).

As shown in Figure 1, the make-up hydrogen stream, and external recycle hydrogen stream, are combined with the internal recycle hydrogen stream from the recycle Hydrogen Compressor (R-101). The combined hydrogen stream is first preheated by heat exchange against the reactor effluent vapor in M-101. A portion of this hydrogen stream is injected into the coal slurry feed at the inlet to the Slurry Feed Heater (L-101) in order to reduce the possibility of coke formation in the slurry heater. Injection of hydrogen into the coal feed slurry also tends to reduce the viscosity of the slurry mixture and results in a lower pressure drop in the slurry feed heater coil. The hydrogen and coal slurry mixture is heated and then fed to the First-Stage Reactor (K-101). The outlet temperature of the slurry feed heater is controlled by adjusting the rate of fuel fired to the heater.

The remaining reactor hydrogen feed stream is fed to the first-stage reactor after being heated in the Hydrogen Heater (L-102). The heater outlet temperature is controlled by adjusting the rate of the fuel fired.

Extra capacity is designed into the hydrogen and slurry feed heaters to ensure flexibility in controlling the reactor operating temperature and for startup operations.

Coal slurry-hydrogen mixture from the slurry feed heater and the hydrogen stream from the hydrogen heater are introduced into the bottom of the First-Stage Reactor (K-101). The reactor operates at approximately 790°F average reactor temperature and 3050 psig. The reactor operating temperature is controlled by adjusting the slurry feed heater outlet temperature to achieve the desired conversion level in the reactor.

Effluent from the first stage reactor is quenched with cold hydrogen from the recycle hydrogen compressor, and introduced into the bottom of the second-stage reactor. The interstage quench hydrogen flow rate is controlled to maintain the second stage reactor in heat balance at 760°F average reactor temperature. In the event of loss of the recycle gas quench or excessive reaction in the reactor, provisions need to be made to provide a solid free liquid stream for use as an emergency interstage quench. (Not shown in Figure 1)

The Reactors (K-101 and K-102) incorporate the principle of the ebullated-bed operation as successfully demonstrated in the Catlettsburg H-Coal® Pilot Plant and in the several operating commercial H-Oil® units around the world. The entire mass in the reactor is held in a fluidized ebullated state by recirculating coal-oil slurry from the top of the reactor through the ebullating pump (J-101 and J-102) and back into the bottom of the reactor. The ebullated bed ensures the reactor's near isothermal operation under the extreme exothermic reactions involved. Increasing the upward flow of the liquid mainly increases the catalyst bed expansion with essentially constant pressure drop. Since the coal particles are much finer than the extrudate catalyst, a separation can be made between these solids such that the coal and ash particles are entrained with the liquid-gaseous reactor effluent products, while the catalyst remains behind in suspension in the reactor. The level of the expanded catalyst bed is observed by radiation detector points, and is maintained by adjusting the ebullating pump flow.

The catalyst in the reactor is maintained at the desired activity level by daily addition of fresh catalyst and corresponding withdrawal of equilibrium catalyst. A catalyst handling system (S-101) is used to add and withdraw catalyst to and from the reactor during normal operation. This system was not designed under the scope of work but the system cost has been estimated.

Hot reactor effluent from the top of the Second-Stage Reactor (K-102) is separated into vapor and slurry phases outside the reactor in the Hot High Pressure Separator (Q-101). The resulting slurry is a net product from the battery limits defined.

Vapor from Q-101 is cooled by heat exchange against hydrogen feed gas in Exchanger (M-101). The vapor and liquid phases resulting from cooling reactor effluent vapor in M-101 are separated in the Warm High Pressure Separator (Q-102). Operating temperature of Q-102 is set high enough to prevent precipitation of ammonium chloride salts. The liquid leaving Q-102 is a net product from the battery limits defined.

Vapor from the Warm High Pressure Separator (Q-102) is water-washed to dissolve any ammonium sulfide, ammonium chloride and bicarbonate salts which might form as the reactor vapor is further cooled. The vapor-water mixture is then air-cooled in the Warm High Pressure Vapor Cooler (M-102). The mixture leaving M-102 is separated into three phases, vapor, hydrocarbon liquid and water in the Cold High Pressure Separator (Q-103). Vapor from Q-103 is fed to the Recycle Hydrogen Compressor (R-101). A purge stream is taken upstream of the Recycle Hydrogen Compressor to prevent build-up of methane, and other non-condensables in the recycle hydrogen stream. The quantity of the purge gas controls the hydrogen purity of the recycle gas. A knockout drum (Q-104) is provided upstream of R-101

The remaining recycle gas stream from R-101, is combined with the make-up hydrogen stream, and sent back to the reactor after being preheated as previously described.

Separation of hydrocarbon liquid from sour water takes place in the Cold High Pressure Separator Q-103. The three phase mixture entering Q-103 is separated in the settling compartment of the drum into a water phase, a light oil phase which floats on top of the water, and a vapor phase which exists above the liquid phase. The oil phase overflows a vertical baffle and is accumulated in the oil compartment of the drum. Water is removed by means of a boot in the settling compartment. The sour water is withdrawn under interface level control.

FIGURE 1
BASELINE DESIGN - SCHEME 1
BATTERY LIMITS STREAMS

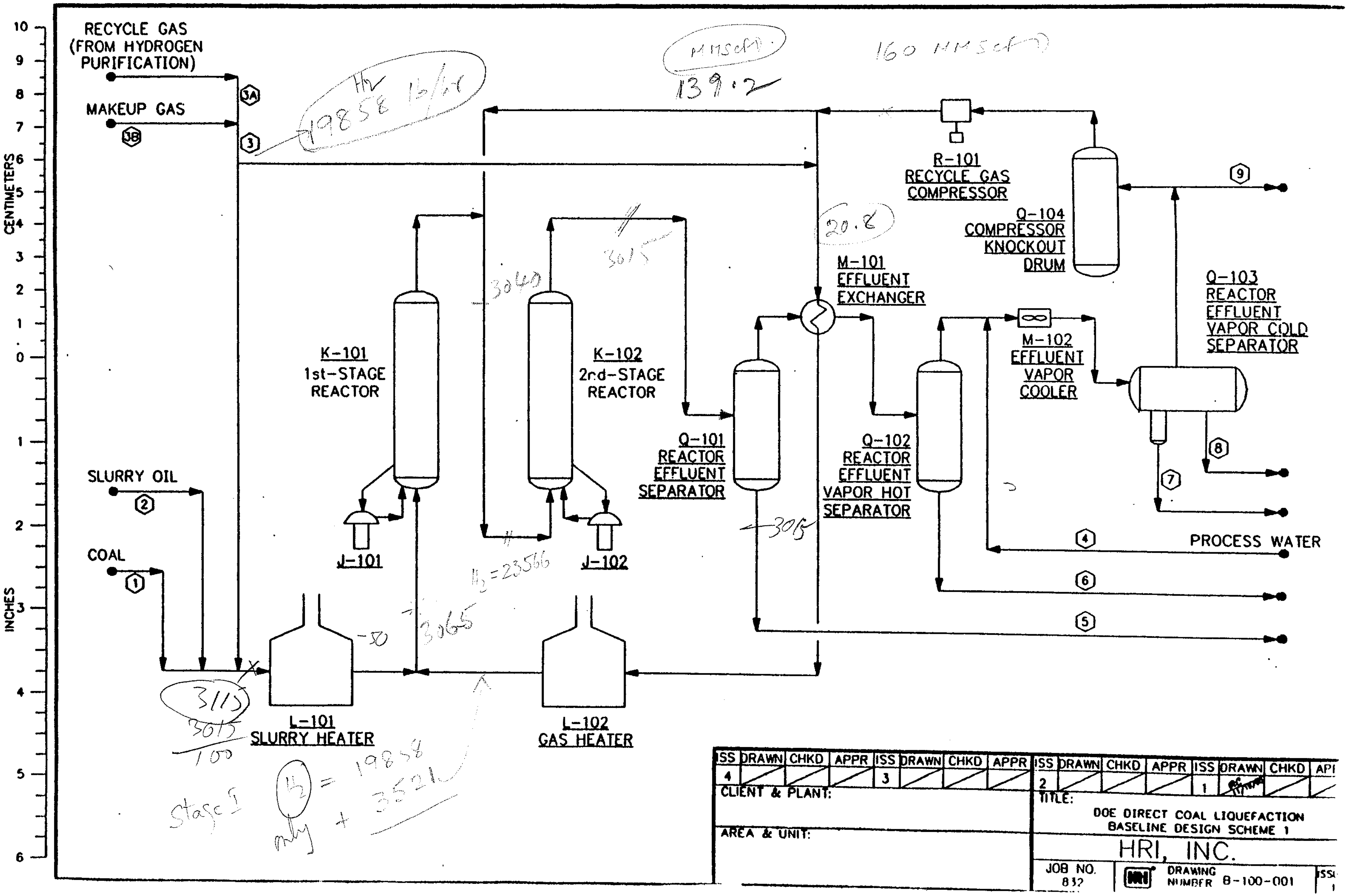

TABLE 8.2

OVERALL MATERIAL BALANCE FOR SCHEME 1

	OVERALL	PER TRAIN
COAL *	1,454,264	290,852
SLURRY OIL	3,095,752	619,150
MAKEUP GAS	85,196	17,039
RECYCLE GAS	38,972	7,794
QUENCH WATER	1,099,643	219,929
TOTAL IN	5,773,827	1,154,764
Q-101 LIQUID	3,851,854	770,370
Q-102 LIQUID	227,161	45,431
Q-103 LIQUID	376,817	75,362
SOUR WATER	1,246,377	249,277
PURGE	71,618	14,324
TOTAL OUT	5,773,827	1,154,764

* Includes 2.0% Moisture on as is coal and 11.47 w% ash on moisture free coal basis

TABLE 8.3

BATTERY LIMITS STREAMS - SCHEME 1

(refer to Figure 1)
Flows given are per train

Stream	1	2	3	3A	3B	4	5	6	7	8	9
Descrip	Coal	Slur Oil	Total Gas	Recycle	Make-Up	Water	Q-101 Liq	Q-102 Liq	Q-103 Liq	Sour Wat	Purge
Flow lb/hr											
H2	0	0	19858	4087	15771	0	923	102	176	2	3010
CO	0	0	0	0	0	0	29	3	9	0	110
H2S	0	0	0	0	0	0	1564	230	2218	1294	1911
NH3	0	0	0	0	0	0	473	74	812	1742	406
H2O	5817	0	39	39.	0	219929	2992	415	97	246213	66
CO2	0	0	0	0	0	0	81	11	70	13	178
C1	0	0	2624	1356	1268	0	1773	226	915	9	4344
C2	0	0	847	847	0	0	1123	164	1370	4	1794
C3	0	0	765	765	0	0	1059	171	2167	0	1203
C4	0	0	252	252	0	0	447	80	1390	0	328
C5-350	0	0	448	448	0	0	8686	2048	29425	0	965
350-450	0	235	0	0	0	0	6721	1986	10470	0	8
450-650	0	17646	0	0	0	0	47172	13737	20854	0	1
650-850	0	92699	0	0	0	0	122004	19466	5128	0	0
850-1000	0	124697	0	0	0	0	119487	6615	261	0	0
1000 +	0	308956	0	0	0	0	330024	103	0	0	0
Subtotal	5817	544233	24033	7794	17039	219929	644558	45431	75362	249277	14324
Coal	252333	26780	0	0	0	0	44973	0	0	0	0
Ash	32702	48137	0	0	0	0	88839	0	0	0	0
TOTAL	290852	619150	24033	7794	17039	219929	778370	45431	75362	249277	14324
Temp. ,F	400	400	100			130	760	550	130	130	130
P., psia	3115	3115	3250			3050	3015	2980	2970	2970	2970
Mol Wt		489.9	2.466			18.0	296.2	153.1	83.8	18.1	7.344
Lb/Ft3 @ cond		47.3	1.103			61.6	41.2	39.8	43.8	61.0	3.210

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TABLE 8.4

EQUIPMENT LIST - SCHEME 1

MAJOR EQUIPMENT CLASSIFICATIONS

J	-	PUMPS
K	-	REACTORS
L	-	FIRED HEATERS
M	-	HEAT EXCHANGERS
Q	-	DRUMS
R	-	COMPRESSORS
S	-	SPECIAL

SCHEME 1 EQUIPMENT LIST PUMPS			
ITEM NO.	SERVICE	NO. REQ'D	COMMENTS
J-101 to J-501	Ebullating Pump, First Stage Reactor	5	Estimated Cost/Pump = $950,000
J-102 to J-502	Ebullating Pump, Second Stage Reactor	5	Estimated Cost/Pump = $950,000

SCHEME 1
EQUIPMENT LIST
REACTORS

ITEM NO.	SERVICE	DIAM.,[4] FT.	LENGTH, FT.	THICK, INCHES	WEIGHT, ST	NO.	COMMENTS
K-101 to K-501	First-Stage Reactor	15'0"	85'6"	11.6	1282	5	2-1/4 Cr 1 Mo[1],[2],[3]
K-102 to K-502	Second-Stage Reactor	15'0"	85'6"	11.6	1282	5	2-1/4 Cr-1 Mo[1],[2],[3]

(1) 347SS Overlay

(2) Refractory Lined (6")

(3) Design Shell Temperature, °F=600
Design Pressure, psig = 3,300

(4) ~~Material~~ METAL Inside Diameter

SCHEME 1
EQUIPMENT LIST

HEATERS

		HEAT ABSORBED, MMBTU/HR PER HEATER				DESIGN PRESSURE
ITEM NO.	SERVICE	NORMAL	DESIGN	NO.	MATERIAL OF CONSTRUCTION	psig
L-101 to L501	Slurry Heater	75.9	87.5	5	347 SS Tubes	3,300
L-102 to L502	Gas Heater	39.8	43.8	5	347 SS Tubes	3,300

SCHEME 1
EQUIPMENT LIST
EXCHANGERS

			PER EXCHANGER		DESIGN CONDITIONS		MAT'LS OF CONSTRUCTION		
ITEM NO.	SERVICE	NO. REQ'D	DUTY MMBTU/HR	SURFACE FT^2	TEMP., °F	PRESS., psig	TUBES	SHELL	TOTAL EST. BHP
M-101 to M-501	Effluent Exchanger	5	60.8	4,980	850	(1)	321 SS	2 1/4 Cr - 1 Mo	-
M-102 to M-502	Effluent Vapor Cooler	5	129.2	18,900(2)	550	3,300	316 LSS	316 LSS(3)	2,600

(1) 3300 psig for tubes, 3250 psig for shell

(2) Bare tube surface

(3) Header box

SCHEME 1
EQUIPMENT LIST
DRUMS

ITEM NO.	NO. REQ'D	SERVICE	POSITION (H=Horizontal V=Vertical)	DIAM, FT	LENGTH FT	DESIGN CONDITIONS TEMP., °F	DESIGN CONDITIONS PRESSURE, PSIG	MATERIALS OF CONSTRUCTION
Q-101 to Q-501	5	Reactor Effluent Drum	V	10'0"	22'6"	600(1)	3300	SA387-GR22 CL2(2)
Q-102 to Q-502	5	Reactor Effluent Vapor Hot Separator	V	9'6"	19'0"	750	3300	SA387-GR22 CL2(2)
Q-103 to Q-503	5	Reactor Effluent Vapor Cold Separator	H	9'6"	27'0"	450	3300	SA387-GR11 CL2(3)
Q-104 to Q-504	5	Compressor Knockout Drum	V	6'6"	13'0"	450	3300	Carbon Steel

(1) Refractory lined. Design temperature for shell and heads.
(2) 347SS Clad
(3) 321SS Clad

SCHEME 1 EQUIPMENT LIST COMPRESSORS	
ITEM	R-101 to R-501
NUMBER REQUIRED	5
FLOW, MMSCFD PER COMPRESSOR Normal Design	 160 176
PRESSURES, psig Inlet Exit	 2935 3235
TEMPERATURE, °F Inlet Exit	 130 150
GAS PROPERTIES Mol. Wt. Cp/Cv	 7.344 1.3224
ESTIMATED BHP, per compressor Normal Design	 1086 1200

SCHEME 1
EQUIPMENT LIST
SPECIAL

ITEM NO.	SERVICE	NO. REQ'D	COMMENTS
S-101 to S-501	Catalyst Handling System	5	Estimated Cost/System = $1,850,000

TABLE 8.5

UTILITY, CATALYST AND CHEMICAL REQUIREMENTS - SCHEME 1

	SCHEME 1
Utilities (Normal)	
Fuel Fired, MMBTU/HR	723
Power, Kwh/Hr	8,597
CATALYST, LBS*	
Initial	2,253,000
Daily Addition (Total)**	68,130

* Amocat 1C 1/12" extrudates

SCHEME 2 - ALTERNATE CASE DESCRIPTION

In Scheme 2, (Figure 2) the alternative of adding vapor-liquid separation between the two catalytic ebullated-bed reactor stages was evaluated. The objective of adding interstage separation between the reactor stages is to remove the light ends formed in the hydrocracking reactions in the first-stage reactor, which in turn reduces the the vapor loading in the second-stage reactor. For the situation where the capacity per reactor train is limited by the vapor loading in the second-stage, the option of adding interstage separation could potentially reduce the total number of reactor trains. For the design of the base case (Scheme 1), the capacity of the reactor trains is limited by the total reactor weight, such that the addition of interstage separation has no impact on the number of reactor trains required.

The description of the process flows for Scheme 2 is essentially the same as for Scheme 1. As described in the preceding discussion, Scheme 2 differs from Scheme 1 by the addition of a vapor-slurry separator (Q-105) between the two reactor stages. The effluent from the first-stage reactor enters Q-105 where it is separated into vapor and slurry streams. The vapor stream joins the vapor exiting from the second-stage reactor separator (Q-101) and enters the M-101 reactor effluent exchanger. The slurry from Q-105 flows directly to the second-stage reactor with recycle hydrogen from the recycle gas compressor (R-101). The description of all other process flows in Scheme 2 is identical to that provided for Scheme 1.

In the flowsheet depicted for Scheme 2, Q-105, the interstage separator, will require a pressure control valve on the vapor exiting the separator (since Q-105 has a higher operating pressure than Q-101). Q-105 will also require a level controller on the slurry exiting the separator. As a result, the situation exists where the flow of the first-stage reactor effluent can be blocked which is potentially very dangerous. To provide a safe and operable design of the interstage separator option, a relief header system designed for reactor blowdown is required. Design of this relief header system was beyond the scope of HRI's work on this alternate scheme.

The design basis for this Alternate Case (Scheme 2) is the same as Case 1 except for the interstage separation.

FIGURE 2

BASELINE DESIGN - SCHEME 2

BATTERY LIMITS STREAMS

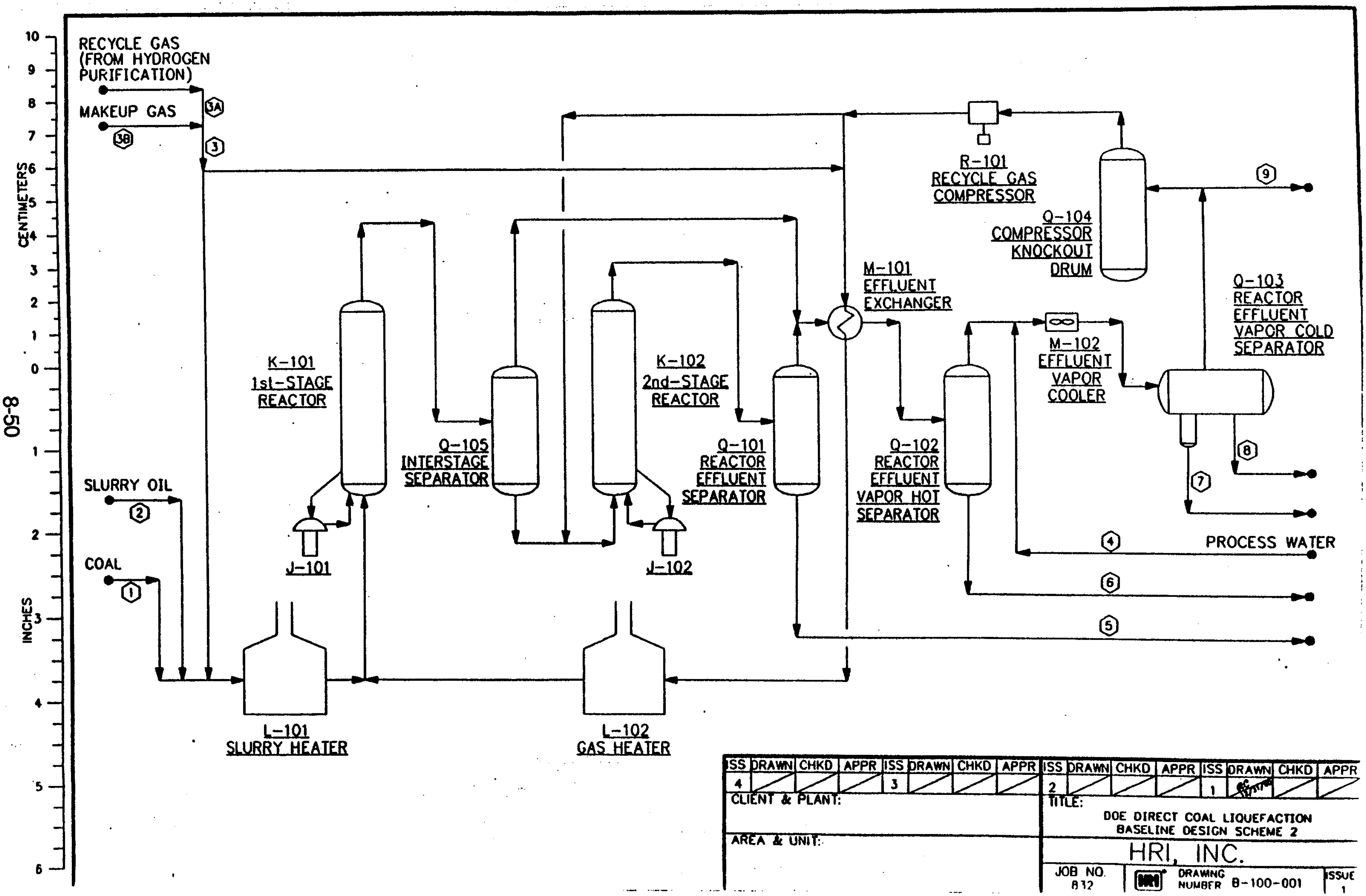


TABLE 8.6

OVERALL MATERIAL BALANCE FOR SCHEME 2

	Overall	Per Train
Coal*	1,454,264	290,852
Slurry Oil	3,095,752	619,150
Makeup Gas	85,641	17,128
Recycle Gas	42,275	8,455
Quench Water	1,099,643	219,929
Total In	5,777,575	1,155,514
Q-101 Liquid	3,833,583	766,717
Q-102 Liquid	245,553	49,110
Q-103 Liquid	371,018	74,203
Sour Water	1,252,539	250,509
Purge	74,882	14,975
Total Out	5,777,575	1,155,514

* Includes 2.0% Moisture on as is coal and 11.47 w% ash on moisture free coal basis

TABLE 8.7

BATTERY LIMITS STREAMS - SCHEME 2

(refer to Figure 2)
Flows given are per train

Stream	1	2	3	3A	3B	4	5	6	7	8	9
Descrip	Coal	Slur Oil	Total Gas	Recycle	Make-Up	Water	Q-101 Liq	Q-102 Liq	Q-103 Liq	Sour Wat	Purge
Flow lb/hr											
N2	0	0	20253	4470	15783	0	891	108	174	2	3433
CO	0	0	0	0	0	0	34	3	8	0	106
H2S	0	0	0	0	0	0	1736	222	2043	1236	1980
NH3	0	0	0	0	0	0	479	73	781	1736	438
H2O	5817	0	43	43	0	219929	1702	410	93	247512	70
CO2	0	0	0	0	0	0	94	10	61	12	176
C1	0	0	2754	1409	1345	0	2076	210	794	8	4309
C2	0	0	928	928	0	0	1320	156	1235	3	1822
C3	0	0	838	838	0	0	1218	167	2033	0	1255
C4	0	0	276	276	0	0	503	79	1340	0	347
C5-350	0	0	491	491	0	0	8900	1994	29243	0	1030
350-450	0	235	0	0	0	0	6161	2041	10975	0	8
450-650	0	17646	0	0	0	0	49685	12646	19432	0	1
650-850	0	92699	0	0	0	0	119075	21866	5656	0	0
850-1000	0	124697	0	0	0	0	117047	8982	335	0	0
1000 +	0	308956	0	0	0	0	329984	143	0	0	0
Subtotal	5817	544233	25583	8455	17128	219929	640905	49110	74203	250509	14975
Coal	252333	26780	0	0	0	0	44973	0	0	0	0
Ash	32702	48137	0	0	0	0	80839	0	0	0	0
TOTAL	290852	619150	25583	8455	17128	219929	766717	49110	74203	250509	14975
Temp. ,F	400	400	100			130	760	550	130	130	130
P., psia	3115	3115	3250			3050	3015	2980	2970	2970	2970
Mol Wt		489.9	2.488			18.0	301.4	159.9	85.3	18.1	6.912
Lb/Ft3 @ cond		47.3	1.194			61.6	41.3	40.0	44.0	61.0	3.008

TABLE 8.8

EQUIPMENT LIST - SCHEME 2

MAJOR EQUIPMENT CLASSIFICATIONS

J	-	PUMPS
K	-	REACTORS
L	-	FIRED HEATERS
M	-	HEAT EXCHANGERS
Q	-	DRUMS
R	-	COMPRESSORS
S	-	SPECIAL

SCHEME 2 EQUIPMENT LIST PUMPS			
ITEM NO.	SERVICE	NO. REQ'D	COMMENTS
J-101 to J-501	Ebullating Pump, First Stage Reactor	5	Estimated Cost/Pump = $950,000
J-102 to J-502	Ebullating Pump, Second Stage Reactor	5	Estimated Cost/Pump = $950,000

SCHEME 2
EQUIPMENT LIST
REACTORS

ITEM NO.	SERVICE	DIAM.,[4] FT.	LENGTH, FT.	THICK, INCHES	WEIGHT, ST	NO.	COMMENTS
K-101 to K-501	First-Stage Reactor	15'0"	85'6"	11.6	1282	5	2-1/4 Cr 1 Mo[1],[2],[3]
K-102 to K-502	Second-Stage Reactor	15'0"	85'6"	11.6	1282	5	2-1/4 Cr-1 Mo[1],[2],[3]

(1) 347SS Overlay

(2) Refractory Lined (6")

(3) Design Shell Temperature, °F=600
Design Pressure, psig = 3,300

(4) ~~Material~~ Inside Diameter
METAL

SCHEME 2
EQUIPMENT LIST
HEATERS

		HEAT ABSORBED, MMBTU/HR PER HEATER				DESIGN PRESSURE
ITEM NO.	SERVICE	NORMAL	DESIGN	NO.	MATERIAL OF CONSTRUCTION	psig
L-101 to L501	Slurry Heater	75.9	87.5	5	347 SS Tubes	3,300
L-102 to L502	Gas Heater	39.8	43.8	5	347 SS Tubes	3,300

SCHEME 2
EQUIPMENT LIST
EXCHANGERS

			PER EXCHANGER		DESIGN CONDITIONS		MAT'LS OF CONSTRUCTION		
ITEM NO.	SERVICE	NO. REQ'D	DUTY MMBTU/HR	SURFACE FT^2	TEMP., °F	PRESS., psig	TUBES	SHELL	TOTAL EST. BHP
M-101 to M-501	Effluent Exchanger	5	60.8	4,980	850	(1)	321 SS	2 1/4 Cr - 1 Mo	-
M-102 to M-502	Effluent Vapor Cooler	5	129.2	18,900(2)	550	3,300	316 LSS	316 LSS(3)	2,600

(1) 3300 psig for tubes, 3250 psig for shell

(2) Bare tube surface

(3) Header box

SCHEME 2
EQUIPMENT LIST
DRUMS

			POSITION			DESIGN CONDITIONS		
ITEM NO.	NO. REQ'D	SERVICE	(H=Horizontal V=Vertical)	DIAM, FT	LENGTH FT	TEMP., °F	PRESSURE, PSIG	MATERIALS OF CONSTRUCTION
Q-101 to Q-501	5	Reactor Effluent Drum	V	10'0"	22'6"	600(1)	3300	SA387-GR22 CL2(2)
Q-102 to Q-502	5	Reactor Effluent Vapor Hot Separator	V	9'6"	19'0"	750	3300	SA387-GR22 CL2(2)
Q-103 to Q-503	5	Reactor Effluent Vapor Cold Separator	H	9'6"	27'0"	450	3300	SA387-GR11 CL2(3)
Q-104 to Q-504	5	Compressor Knockout Drum	V	6'6"	13'0"	450	3300	Carbon Steel

Q-105 to Q 505 5

(1) Refractory lined. Design temperature for shell and heads.
(2) 347SS Clad
(3) 321SS Clad

MISSING

~~7-61~~ 8-58

SCHEME 2 EQUIPMENT LIST COMPRESSORS	
ITEM	R-101 to R-501
NUMBER REQUIRED	5
FLOW, MMSCFD PER COMPRESSOR Normal Design	 160 176
PRESSURES, psig Inlet Exit	 2935 3235
TEMPERATURE, °F Inlet Exit	 130 150
GAS PROPERTIES Mol. Wt. Cp/Cv	 7.344 1.3224
ESTIMATED BHP, per compressor Normal Design	 1086 1200

SCHEME 2 EQUIPMENT LIST SPECIAL			
ITEM NO.	SERVICE	NO. REQ'D	COMMENTS
S-101 to S-501	Catalyst Handling System	5	Estimated Cost/System = $1,850.00

TABLE 8.9

SUMMARY OF UTILITIES

UTILITY, CATALYST AND CHEMICAL REQUIREMENTS - SCHEME 2

	SCHEME 2	
UTILITIES (NORMAL)		
Fuel Fired, MMBTU/HR	699	
Power, Kwh/Hr	9.230	9230(?)
CATALYST,*		
Initial	2,253,000	
Daily Addition (Total)	68,130	

* Amocat 1C 1/12" extrudates

DISCUSSION AND SUMMARY

HRI's design of a portion of the liquefaction section for the DOE Direct Liquefaction Baseline Design and System Analysis for Bechtel has shown that five parallel reaction trains are required to achieve the desired processing capacity. What is limiting in the capacity of the reaction trains is the maximum weight constraints for the reactors.

In the design basis specified by Bechtel from Wilsonville operations, the reactor temperatures are both less than 800 F. In order to obtain high coal conversions and yields of distillate liquid products, long residence time (low space velocity) is required. As a result, the volume of the reactors required is large and the absolute size of the reactors becomes limiting. In addition to the long residence time required by the low reactor temperatures to obtain high conversions and liquid yields, the design basis also required a very high rate of recycle slurry, with an extremely high concentration of unconverted residual oil (1000 F+). It is well known that a high recycle ratio contributes to reduced thermal efficiency and increases the size and cost of equipment upstream and downstream of the reactors. The high concentration of unconverted residual oil (1000 F+) in the recycle slurry, while not necessarily contributing to higher costs, will make unit operations more difficult due to the high resulting viscosity in the entire system.

One aspects of this design which requires comment is the high/low temperature staging. This temperature sequencing is thermally inefficient due to the exothermic nature of the reactions involved. The effluent from the first-stage must be quenched to a temperature well below the operating temperature of the second-stage due to the exothermic heat of reaction. Because of this extreme quenching requirement the overall flow of hydrogen is split almost equally between the two reactor stages, * even though more than 50% of the hydrogen is consumed in the first-stage. From a thermal efficiency point of view it would be greatly preferred to operate the second-stage at a higher temperature than the first-stage. HRI's research also shows advantages in process performance with low/high temperature staging.

The results shown for the alternate case with interstage separation were initially surprising. It was hoped that the use of an interstage separator might show some benefit, even as much as a reduction in the number of reaction trains. In evaluating the observations in the preceding discussions, it is obvious that the intent of interstage separation did not address the limitations in the baseline design. The use of interstage separation may be beneficial, especially with a different design basis, that would include higher overall temperatures and low/high temperature staging.

With higher overall temperatures, the total reactor size/weight would no longer be limiting. This would then reduce to the situation where the vapor loading in the

* Please see attached note from HRI

Comments on hydrogen flow to reactors:

copied June 10, 1991 fax to Syamal Poddar - Bechtel Corporation, from John Duddy - HRI, Inc.

"Subject: DOE Direct Coal Liquefaction Baseline Design and Systems Analysis - HRI Report - Hydrogen Flows

The following is being provided to help clarify some of the comments on hydrogen flows in the subject report:

Hydrogen, lb/hr	*Per Train*	
Total Make-up (Stream 3)	19,858	
Total Recycle (Internal)	27,087	
Total Hydrogen	46,945	
First-Stage		
Make-up	19,858	
Recycle	3,521	
Total	23,379	
Second-Stage		
Recycle	23,566	
Split by Stage		%
Total First-Stage	23,379	50
Total Second-Stage	23,566	50
Total both stages	46,945	100
Recycle to First-Stage	3,521	13
Recycle to Second-Stage	23,566	87
	27,087	100

As shown by this analysis, the comments made in the report (Page 36) and in the response to Bechtel questions/comments (March 6, 1991, Ref: 91/JED/24) are correct, namely:

1) Page 36 of report - "...the overall flow of hydrogen is split almost equally between the two reactor stages..."

2) *Response to question/comment 2 - "87% of this stream (recycle hydrogen) goes to K-102 (second-stage)."*

The factor which is causing some confusion in our discussions regarding this analysis relates to the hydrogen which is not consumed in the first-stage and flows to the second-stage. The following shows the same analysis on this basis:

Hydrogen, lb/hr	*Per Train*	
First-Stage		
Make-up	*19,858*	
Recycle	*3,521*	
Total	*23,379*	
Consumed	*(9,336)*	
• Second-Stage	*14,043*	
Second-Stage		
Recycle	*23,566*	
From First-Stage	*14,043*	
Total	*37,609*	
Split by Stage		*%*
First-Stage	*23,379*	*38*
Second-Stage	*37,609*	*62*
Total	*60,988*	*100*

This analysis is not as precise as the previous analysis since the hydrogen which is not consumed in the first-stage is double counted in the calculation of the total flow. HRI believes that the information previously provided gives Bechtel the most accurate depiction of the split of hydrogen flows between the first- and second stage reactors."

second-stage reactor would limit the capacity per reactor train. Use of interstage separation could then be effective in increasing the capacity per reactor train, by decreasing the total vapor loading in the second-stage.

In low/high temperature staging, it may be possible to utilize the exothermic heat of reaction to maintain the heat balance in the second-stage reactor. In this case the hydrogen requirement to the second-stage would no longer be set by heat balance considerations. The hydrogen requirement would be set by the desired hydrogen partial pressure in the reactor. This would result in minimum hydrogen requirement in the second-stage of a two-stage system, and a potentially attractive option for interstage separation.

9. Plant 3 (Gas Plant)

9.0 Design Basis, Criteria and Considertions

The Gas Plant consists of the following sections:

- Absorber/deethanizer
- Lean oil stripper/debutanizer
- Depropanizer
- LPG Merox for propane and butane products
- Makeup lean oil stripper

The Gas Plant will be designed as a two-train unit to increase the overall plant reliability.

Feed to the Gas Plant will be tail gas from the Hydrogen Purification Plant (Plant 6). Makeup lean oil is the naphtha product from Plant 2. Lean oil purge from the lean oil stripper/debutanizer will be sent to the Naphtha Hydrotreater (Plant 4). Lean oil stripper offgas is returned to Plant 6 for treating.

The products from the Gas Plant include:

- C_3 LPG
- C_4 LPG
- Fuel Gas

The specification for the two LPG products are as follows:

Specification	C_3 LPG Max.	C_4 LPG Max.
Ethane (vol.%)	2.0	--
Propane (vol.%)	--	2.0
Butane (vol.%)	2.0	--
Pentane (vol.%)	--	2.5
Mercaptans (wt.ppm)	20	20

Design Considerations

Lean oil absorption was used as the method to recover light hydrocarbons from miscellaneous gas streams being sent to the fuel gas system because of the high recovery of propane and butanes, smaller equipment sizes, and recovery without the need for a refrigeration system. The lean oil absorber and rich oil deethanizer operations were combined into a single tower to reduce capital and operating costs.

Naphtha product from the Coal Liquefaction Plant is used as the lean oil makeup. Because the makeup rate is close to the Plant 2 production rate, the total naphtha stream was sent to the Gas Plant. The lean oil purge (Stripper/Debutanizer bottoms) is sent as feed to the Naphtha Hydrotreater. This scheme simplifies the operation of the system and reduces the amount of light components in the feed to Plant 4. The makeup lean oil is stripped prior to entering the absorption system to remove ammonia and acid gases which may contaminate LPG products or fuel gas and light hydrocarbons which may be lost to fuel gas in the absorber overhead stream.

Some pentanes are lost to fuel gas in the absorber overhead vapor stream due to vapor-liquid equilibrium. A refrigerated cooler would be required to recover this lost material; however, little additional propane recovery would be made since the recovery rate is already quite good.

Plant 3 was modeled using the PROCESS simulation software. Equipment was sized using these simulations.

9.1 Process Description and Process Flow Diagram

As pointed out earlier, the plant consists of five sections. These are: Absorber/Deethanizer, Lean Oil Stripper/Butanizer, Depropanizer, Makeup Lean Oil Stripper and LPG Merox for propane and butane products. The first four sections of this plant are shown in Process Flow Diagram, Figure 9.1 and the Merox section is shown in Process Flow Diagram, Figure 9.2. The description of each of these sections is included below.

Absorber/Deethanizer

Compressed tail gas from the pressure swing absorption (PSA) unit of the Hydrogen Purification Plant (Plant 6) is fed to a combined lean oil absorber/deethanizer tower to recover the propane and heavier liquids in the stream. The tower is operated at 160 psig. The tower has a reboiler to strip the ethane and lighter components from the rich oil exiting the bottom of the tower. The overhead gas is mixed with the recirculated lean oil and the makeup lean oil and sent to a water cooler before entering the precontactor drum. The vapor product is sent directly to the fuel gas system. The liquid from the precontactor is fed to the top tray of the absorber/deethanizer tower. The precontacting operation improves the recovery of propane from fuel gas and reduces the losses of pentanes in the lean oil to fuel gas by operating the vapor/liquid separation at a lower temperature than available at the top tray of the tower.

Lean Oil Stripper

The lean oil is the naphtha product from Plant 2. The Plant 2 naphtha contains hydrogen sulfide and ammonia which would contaminate the fuel gas, propane, and mixed butane products. Prior to being sent to the absorber/deethanizer as makeup lean oil, the naphtha is stripped of the contaminants in a stripper tower, which operates as a debutanizer. The stripped gases are sent to the low pressure gas compressor in Plant 6 for scrubbing, and recovery of hydrogen and light gases.

Stripper/Debutanizer

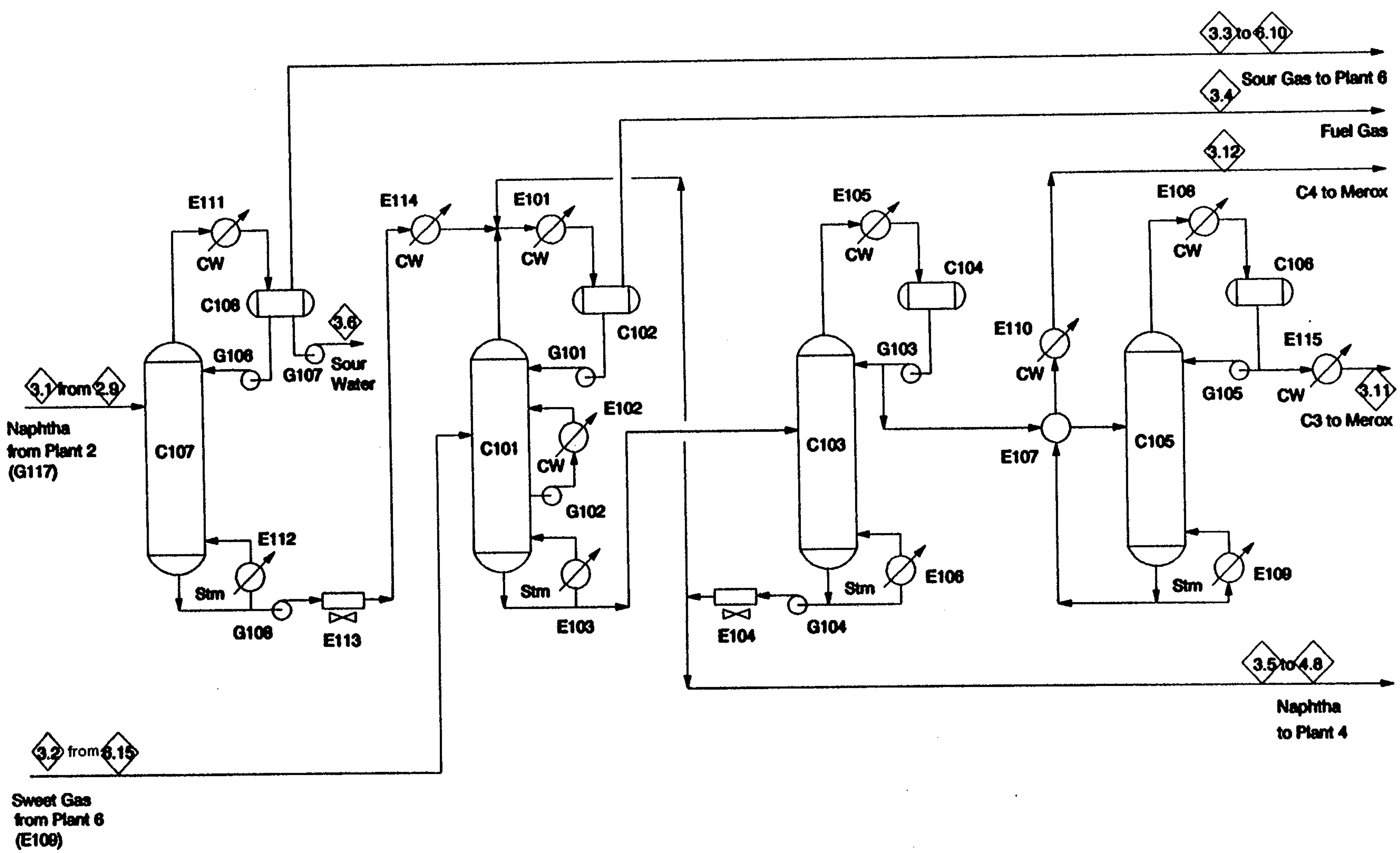
Rich oil from the absorber/deethanizer tower is sent to the rich oil stripper tower to remove propane and butanes from the lean oil. The lean oil from the stripper tower is returned to the absorber/deethanizer via heat exchange with the rich oil feed. Some of the recirculated lean oil is purged to the Naphtha Hydrotreater (Plant 4) to remove pentane and heavier hydrocarbons recovered from the feed gases and maintain a more consistent lean oil composition.

Depropanizer

The overhead product from the stripper is fed to a depropanizer tower to separate the light hydrocarbons into two LPG products (propane and mixed butanes). The two products are sent to individual Merox units for the removal of trace sulfur compounds prior to being sent to product storage.

Merox Treating

The Merox units remove mercaptan sulfur compounds from the LPG products (propane and mixed butanes) by dissolving the compounds in a 20° Baume caustic soda solution. This is accomplished in a multistage, vertical Merox extraction. A caustic prewash of 10° Baume solution is required to remove any remaining hydrogen sulfide from the feed stream before entering the extraction tower. The mercaptan-rich caustic solution from the bottom of the extraction towers of the two Merox units flow to the common regeneration section through a small steam heater. The caustic stream is regenerated with air in the presence of a catalyst, where the mercaptans are converted to disulfides. The caustic stream is returned to the Merox units, and the disulfide stream is sent to the Naphtha Hydrotreater feed for disposal.



Plant 3
Gas Plant

Figure 9.1 Process Flow Diagram

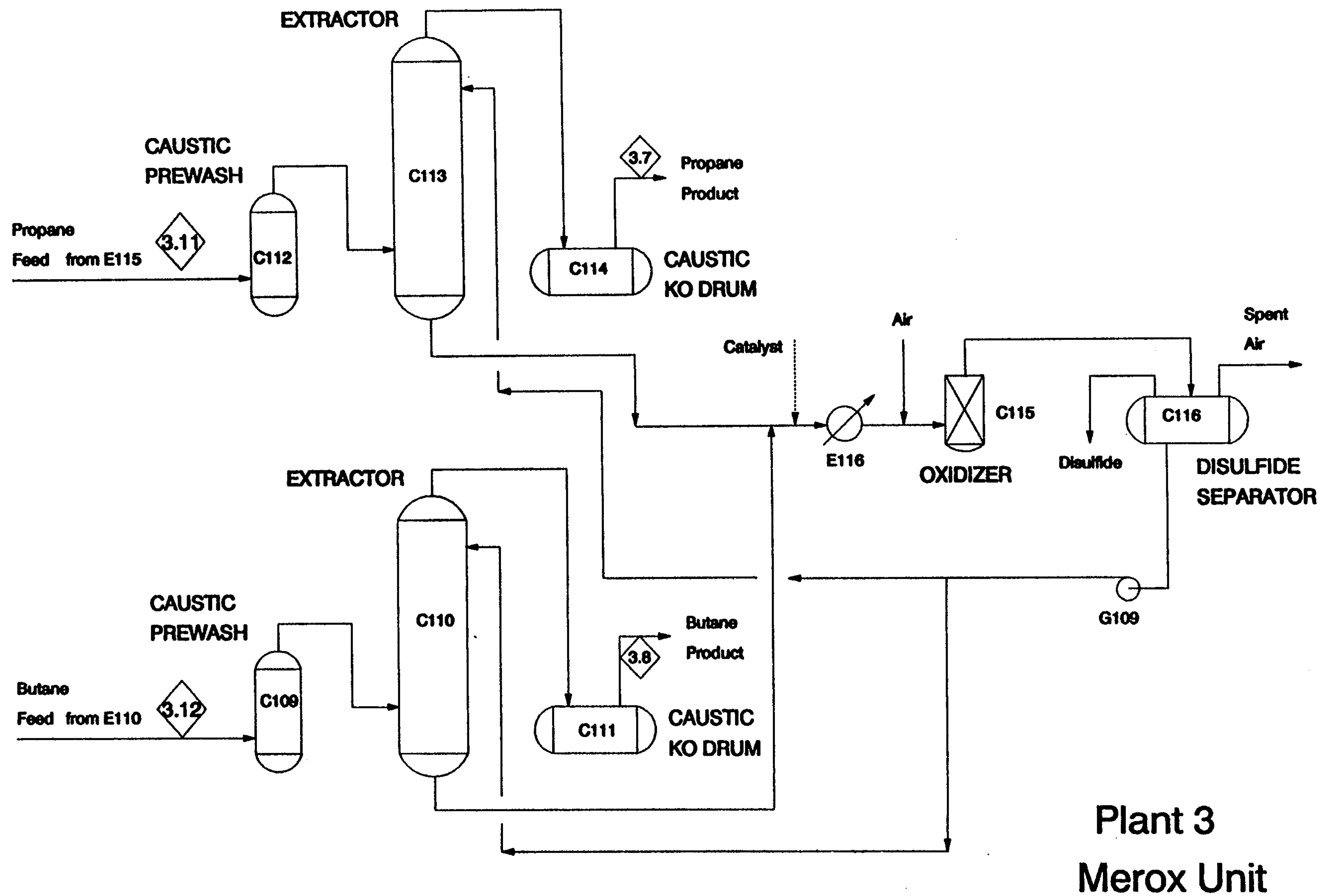


Figure 9.2 Process Flow Diagram

9.2 Material Balance

The material balance for a single train of this plant train is shown in table 9.1 on the next page. Two trains are required.

TABLE 9.1
PLANT 3
MATERIAL BALANCE

	PLANT INPUT			PLANT OUTPUT						
STREAM NO.	3.1	3.2		3.3	3.6	3.4	3.7	3.8	3.5	
COMPONENTS	PLT 2 NAPH #/hr	PSA TG FR PLT 6 #/hr	TOTAL INPUT #/hr	LO OH GAS TO #6 #/hr	SOUR WATER #/hr	FUEL GAS #/hr	C3 PROD #/hr	MIX. C4 PROD #/hr	NAPHTHA TO PLT 4 #/hr	TOTAL OUTPUT #/hr
N2	0	639	639			639				639
H2	0	1127	1127	0		1127				1127
H2O	1215	1519	2734	23	1192	1519				2734
H2S	58	22	79	58		22				79
NH3	0	0	0			0				0
CO	0	378	378			378				378
CO2	0	0	0			0				0
C1	5	16238	16243	5		16237	1			16243
C2	68	13192	13259	68		12972	216	4	0	13259
C3	390	15875	16265	390		24	15641	210	0	16265
C4	855	14688	15543	855		70	353	14164	101	15543
C5–350	85568	20549	106117	599		3150		347	102021	106117
350–450	1578	52	1630	0		1			1629	1630
450–650	0	13	13	0					13	13
650–850	0	0	0						0	0
850–1000	0	0	0						0	0
1000+	0	0	0						0	0
PHENOLS	0	0	0						0	0
MEA		18	18						18	18
UNCONVERTED COAL	0	0	0						0	0
ASH	0	0	0						0	0
TOTAL	89735	84309	174044	1997	1192	36139	16211	14725	103781	174044
TEMPERATURE (F)	140	95		137	137	100	100	100	130	
PRESSURE (psig)	50	210		40	40	200	265	265	248	

9.3 Major Equipment List

The major equipment list for this plant per train is shown in Table 9.2.

Table 9.2 Major Equipment List

PLANT 3 – GAS PLANT

Reactors and Vessels

Equipment No.	Equipment Description
3.1– C102	Abs/Deeth Ovhd Precontactor
3.2– C104	Lean Oil Stripper Ovhd Accum
3.3– C106	Depropanizer Ovhd Accum
3.1– C108	LO Makeup Strp Ovhd Accum

Towers

Equipment No.	Equipment Description
3.1– C101	Absorber/Deethanizer
3.2– C103	Lean Oil Stripper
3.3– C105	Depropanizer
3.1– C107	Lean Oil Makeup Stripper

Heat Exchangers

Equipment No.	Equipment Description
3.1– E101	Absorb/Deeth Ovhd Cooler
3.1– E102	Absorb/Deeth Intercooler
3.1– E103	Absorb/Deeth Reboiler
3.2– E104	Lean Oil Cooler
3.2– E105	Lean Oil Str Ovhd Condenser
3.2– E106	Lean Oil Str Reboiler
3.3– E107	Depropanizer Feed/Btms Exch
3.3– E108	Depropanizer Ovhd Condenser
3.3– E109	Depropanizer Reboiler
3.3– E110	Depropanizer Btms Cooler
3.1– E111	LO Makeup Strp Ovhd Cond
3.1– E112	LO Makeup Strp Reboiler
3.1– E113	LO Makeup Strp 1st Btms Clr
3.1– E114	LO Makeup Strp 2nd Btms Clr
3.3– E115	Depropanizer Ovhd Cooler

Table 9.2 Major Equipment List - continued

Pumps

Equipment No.		Equipment Description
3.1-	G101	Absorb/Deeth Precontact Liq
3.1-	G102	Intercooler Pumparound
3.2-	G103	Lean Oil Strp Reflux Pump
3.2-	G104	Lean Oil Circulation Pump
3.3-	G105	Depropanizer Reflux Pump
3.1-	G106	LO Makeup Strp Reflux Pump
3.1-	G107	LO Makeup Strp Sour Water Pump
3.1-	G108	LO Makeup Strp Btms Pump

9.4 Utility Summary

The utility requirement for this plant is summarized in Table 9.3

Table 9.3

Utility Summary

	Utility per Train
Steam, 600 psig	198,305 lbs/hr
Steam, 50 psig	8,593 lbs/hr
Electricity	457 kW
Cooling Water	10,977 gpm
Fuel Gas Production	737 MM Btu/hr

9.5 Water Summary

Sour Water = 2.5 gpm

10. Plant 4 (Naphtha Hydrotreater)

10.0 Design Basis, Criteria and Considerations

The Naphtha Hydrotreater will be designed as a two-train unit to increase the overall plant reliability.

Feed to the Naphtha Hydrotreater will be the naphtha (C_5-350°F) product from the atmospheric fractionator tower at the Coal Liquefaction Plant (Plant 2) via the Gas Plant (Plant 3). An additional stream from Gas Oil Hydrotreater (Plant 5) is sent to the downstream section (Stabilizer) of Naphtha Hydrotreater (Plant 4).

Design criteria for the Naphtha Hydrotreater is based on the following:

Reactor Inlet Pressure, psig	1,000
Reactor Outlet Pressure, psig	950
Hydrogen Partial Pressure, psia (outlet of reactor)	700
Reactor Inlet Temperature, °F	525
Reactor Outlet Temperature, °F	575
Chemical Hydrogen Consumption, SCF/B	125
LHSV, V/V/hr	2.0
Catalyst Type	NiMo

Characteristics of the product from the Naphtha Hydrotreater is given in Table 10.1 below:'

Table 10.1

Naphtha Hydrotreater Product Characteristics

	C_5-350°F
Gravity, °API	52.2
Distillation, TBP °F	
IBP	72
5% (wt.)	110
10%	138
30%	207
50%	246
70%	284
90%	328
95%	340
EP	344
Sulfur, wppm (maximum)	1.0*
Nitrogen, wppm (maximum)	0.2*

*Note: When Nitrogen Specification is achieved at 0.2 wppm, Sulfur level will go down further (0.1 wppm).

Design Considerations

Because the Naphtha Hydrotreater is a vapor phase reaction, only cold separators (both high pressure and low pressure) are included in the design of the plant.

The naphtha stream from the Gas Oil Hydrotreater is sent directly to the Plant 4 fractionator because it has already been treated at more severe conditions than those in the Naphtha Hydrotreater.

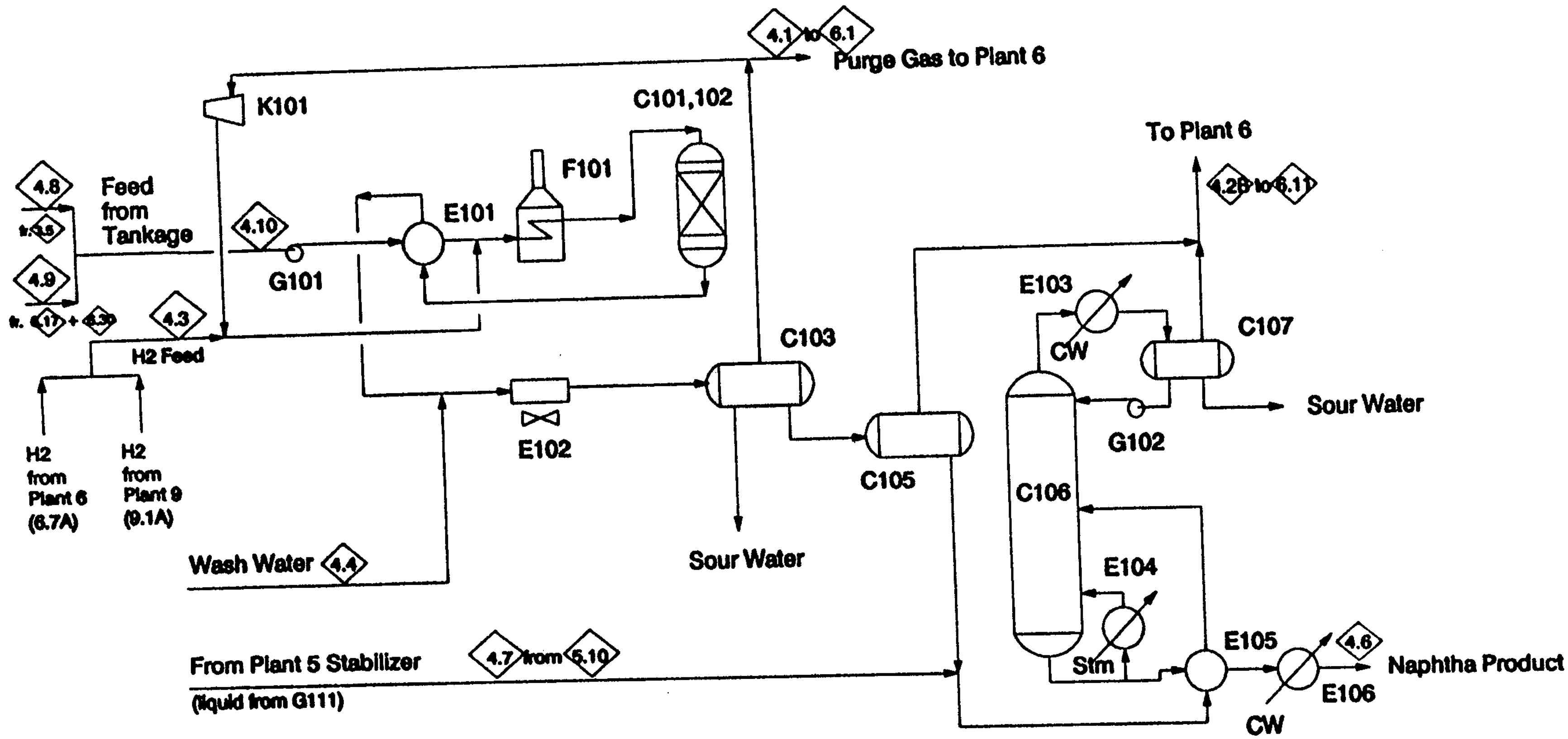
The fractionator was designed to use 600 psig steam for the reboiler because the bottoms temperature is 450°F and to reduce capital costs.

10.1 Process Description and Process Flow Diagram

The Plant includes two distinct sections. These are: 1) Reaction and Primary Separation section and 2) Product Distillation section. Figure 10.1 is the Process Flow Diagram of the Plant.

Reaction and Primary Separation Section. Feed is pumped from intermediate tankage directly to heat exchange with the reactor effluent (E101). Hydrogen is then mixed with the heated feed prior to entering the reactor feed preheater (F101), where the mixture is raised to the reactor inlet temperature. The feed/hydrogen mixture enters the fixed bed reactors (C101 and C102), which operate at a pressure of 1000 psig. The reactor effluent passes through heat exchange with the cold feed (E101), through an air-fin cooler (E102), and to the high pressure separator (C103). Wash water is injected into the reactor effluent line prior to the air-fin cooler for control of ammonium salts. Hydrogen rich vapor is compressed in K101 and returned to the front end of the unit to mix with makeup hydrogen and the unit feed. A purge gas stream is sent to Plant 6 to remove light hydrocarbons from the system. Sour water is removed from the high pressure separator to be eventually sent for ammonia recovery. The hydrocarbon liquid is flashed to 185 psig in the low pressure separator (C105) to remove light hydrocarbons, which are sent to Plant 6. The low pressure separator liquid is then sent to product distillation.

Product Distillation. The low pressure separator liquid is preheated with the product from the fractionator in exchanger E105 before entering the tower (C106). The overhead vapor from the tower is partially condensed and only a vapor product is withdrawn for further processing at Plant 6. The fractionator has a steam-heated reboiler (E104). The bottoms naphtha product is cooled in exchangers (E105 and E106) and sent to product storage.



Notes: (1) Combined sour water rate is represented by 4.5

Plant 4
Naphtha Hydrotreater

Figure 10.1 Process Flow Diagram

10.2 Material Balance

Material balance for Plant 4 is shown in Table 10.2. There are two trains in the plant. The material balance shown in Table 10.1 is for a single train only.

Table 10.2 Material Balance - Plant 4

PLANT 4.

PLANT INPUT

COMPONENTS	4.8 NAPHTHA FROM #3 #/hr	4.9 PLT 6 KO LIQ #/hr	4.3 H2 FROM PLT. 9 & 6 #/hr	4.4 WASH WATER #/hr	4.7 PLT 6 NAPH #/hr	TOTAL INPUT #/hr	RXN DELTA #/hr
N2	0	0	10		0	10	
H2	0	0	1250		0	1250	-246
H2O	0	128	0	9386	0	9514	567
H2S	0	0	0		0	0	11
NH3	0	0	0		0	0	207
CO	0	0	0		0	0	
CO2	0	0	0		0	0	
C1	0	2	3		1	6	543
C2	0	7	0		2	9	364
C3	0	25	0		49	74	350
C4	101	59	0		455	615	1391
C5–350	102021	1393	0		9573	112087	-8065
350–450	1629	175	0		763	2566	5795
450–650	13	0	0		0	13	-1
650–850	0	0	0		0	0	
850–1000	0	0	0		0	0	
1000+	0	0	0		0	0	
PHENOLS	0	0	0		0	0	
MEA	18	0	0		0	18	-16
UNCONVERTED COAL	0	0	0		0	0	
ASH	0	0	0		0	0	
TOTAL	103781	1787	1263	9386	10843	127059	0
TEMPERATURE (F)	130	130	100	112	136		
PRESSURE (psig)	148	150	1075	950	200		

PLANT OUTPUT

COMPONENTS	4.1 H2 PURGE #/hr	4.2A SW VAP TO PLT 6 #/hr	4.5 SOUR WATER #/hr	4.2B GAS TO PLT 6 #/hr	4.6 NAPH PROD #/hr	TOTAL OUTPUT #/hr
N2	7			3		10
H2	948	1		56		1004
H2O	24	1	10033	23		10081
H2S	0		11	0		11
NH3	8		162	37		207
CO				0		0
CO2				0		0
C1	380	1		168		549
C2	128			245		373
C3	61			363		424
C4	103			1724	179	2006
C5–350	288			143	103591	104022
350–450					8361	8361
450–650					12	12
650–850					0	0
850–1000					0	0
1000+					0	0
PHENOLS					0	0
MEA					2	2
UNCONVERTED COAL					0	0
ASH					0	0
TOTAL	1947	2	10206	2761	112144	127059
TEMPERATURE (F)	135		135	131	100	
PRESSURE (psig)	916		*	160	153	

* This stream represents a combination of streams; a single pressure cannot be reported.

10.3 Major Equipment List

The major equipment list for the Plant is shown in Table 10.3.

Table 10.3 Major Equipment List

PLANT 4 -- NAPHTHA HYDROTREATER

Reactors and Vessels

Equipment No.	Equipment Description
4.1- C101	Hydrotreater Reactor
4.1- C102	Hydrotreater Reactor
4.1- C103	High Pressure Separator
4.1- C104	Compressor Knockout Drum
4.1- C105	Low Pressure Separator
4.2- C107	Fractionator Ovhd Accum
4.1- C108	Wash Water Drum
4.1- C109	Sour Water Drum

Towers

Equipment No.	Equipment Description
4.1- C106	Fractionator

Fired Heaters

Equipment No.	Equipment Description
4.1- F101	Reactor Feed Preheater

Heat Exchangers

Equipment No.	Equipment Description
4.1- E101	Reactor Feed/Effluent Ex
4.1- E102	Reactor Effluent Cooler
4.2- E103	Fractionator Condenser
4.2- E104	Fractionator Reboiler
4.2- E105	Fractionator Feed/Bottoms Ex
4.2- E106	Fract Bottoms Prod Cooler

Table 10.3 Major Equipment List - continued

Compressors

Equipment No.	Equipment Description
4.1- K101	Recycle Hydrogen

Pumps

Equipment No.	Equipment Description
4.1- G101	Reactor Charge Pump
4.2- G102	Fractionator Reflux Pump
4.1- G103	Sour Water Drum Pump
4.1- G104	Wash Water Pump

10.4 Utility Summary

The utility summary for this Plant is included in Table 10.4.

Table 10.4

Utility Summary

	(per Train)
Steam, lbs/hr 160 psig	13,311 lbs/hr
Electricity	536.5 KW
Fuel Gas	0.9 MMSCFD
Cooling Water	1,282 gpm

Note: Fuel gas consumption is based on a heating value of 1000 Btu/SCF.

10.5 Water Summary

Table 10.5 below presents the water balances for the Naphtha Hydrotreater. This balance is per train.

Table 10.5

WATER SUMMARY

Water Inlet	lb/hr
Feed	128
Produced in Reactors	567
Wash Water	9,386
TOTAL	10,081
Water Outlet	
Sour Water to NH_3 Recovery	10,033
Hydrogen Purge	24
Low Pressure Gases to Plant 6	24
Products	0
TOTAL	10,081
Sour Water to NH_3 Recovery, gpm	20

11.0 Plant 5 (Gas Oil Hydrotreater)

11.0 Design Basis Criteria and Considerations

The Gas Oil Hydrotreater is designed as a two-train unit to increase the overall plant reliability.

Feeds to the Gas Oil Hydrotreater are the distillate sidestream from the atmospheric tower and the overhead liquid and the upper sidestream from the vacuum tower of the Coal Liquefaction (Plant 2). The three products (350-450°F, 450-650°F, and 650-850°F) will be separated in a fractionator on the back-end of the unit.

Characteristics of the three feeds from Coal Liquefaction are presented in Table 11.1 below:

Table 11.1

Feeds to the Gas Oil Hydrotreater

	Atmospheric Sidestream	Vacuum Overhead	Vacuum Upper SS
Feed Rate, BPSD	19,582	34	23,842
Gravity, °API	17.9	21.0	9.0
Product Cuts (wt.%)			
C_6-350°F	1.7	4.4	
350-450°F	24.7	50.5	0.1
450-650°F	58.0	45.1	37.3
650-850°F	15.6		62.6

Design criteria for the Gas Oil Hydrotreater is based on the following:

Reactor Inlet Pressure, psig	2,600
Reactor Outlet Pressure, psig	2,500
Hydrogen Partial Pressure, psia (outlet of reactor)	1,800
Reactor Inlet Temperature, °F	600
Reactor Outlet Temperature, °F	750
Chemical Hydrogen Consumption, SCF/B	1,080
LHSV, V/V/hr	1.0
Catalyst Type	NiMo

Characteristics of the products from the Gas Oil Hydrotreater are given in Table 11.2 below:

Table 11.2

Products from the Gas Oil Hydrotreater

	350-450°F	450-650°F	650-850°F
Gravity, °API	29.9	17.3	10.5
Distillation, TBP °F			
IBP	340	414	640
5% (wt.)	348	463	645
10%	358	469	665
30%	376	531	676
50%	399	576	717
70%	414	619	759
90%	433	645	815
95%	449	655	835
EP	465	665	838
Sulfur wppm	20	20	20
Nitrogen wppm	500	500	500

Design Considerations

Three final products (350-450°F, 450-650°F, and 650-850°F) were combined for hydrotreating in the same unit because of the reasons below:

- The reactions conditions recommended are approximately the same for the two heavier products and the lighter product is relatively small.
- Significant savings in capital cost and operating cost are realized by hydrotreating in a common unit even though final fractionation is more complex.
- The initial fractionation on the Coal Liquefaction Plant can be made as rough cuts and thus save capital and operating costs.

The reaction system was designed with parallel reactors and only one in series. Because of the reaction conditions, the reactors were designed with two catalyst beds with an interbed hydrogen quench.

The separation system was designed with high and low pressure systems, each with hot and cold separators. There was no incentive for a warm separator as in Plant 2. The letdown from the high pressure to the low pressure systems was done through an expander turbine to improve the efficiency of the process. The expander was designed to drive the reactor feed booster pump.

It was very difficult to fractionate all three products on the same tower; therefore, the 450°F- material was taken overhead in the fractionator to a small outboard stabilizer tower to make the front-end cut on the 350-450°F product. There was no problem fractionating the 450-650°F as a sidestream product or the 650-850°F as a bottoms product.

11.1 Process Description and Process Flow Diagram

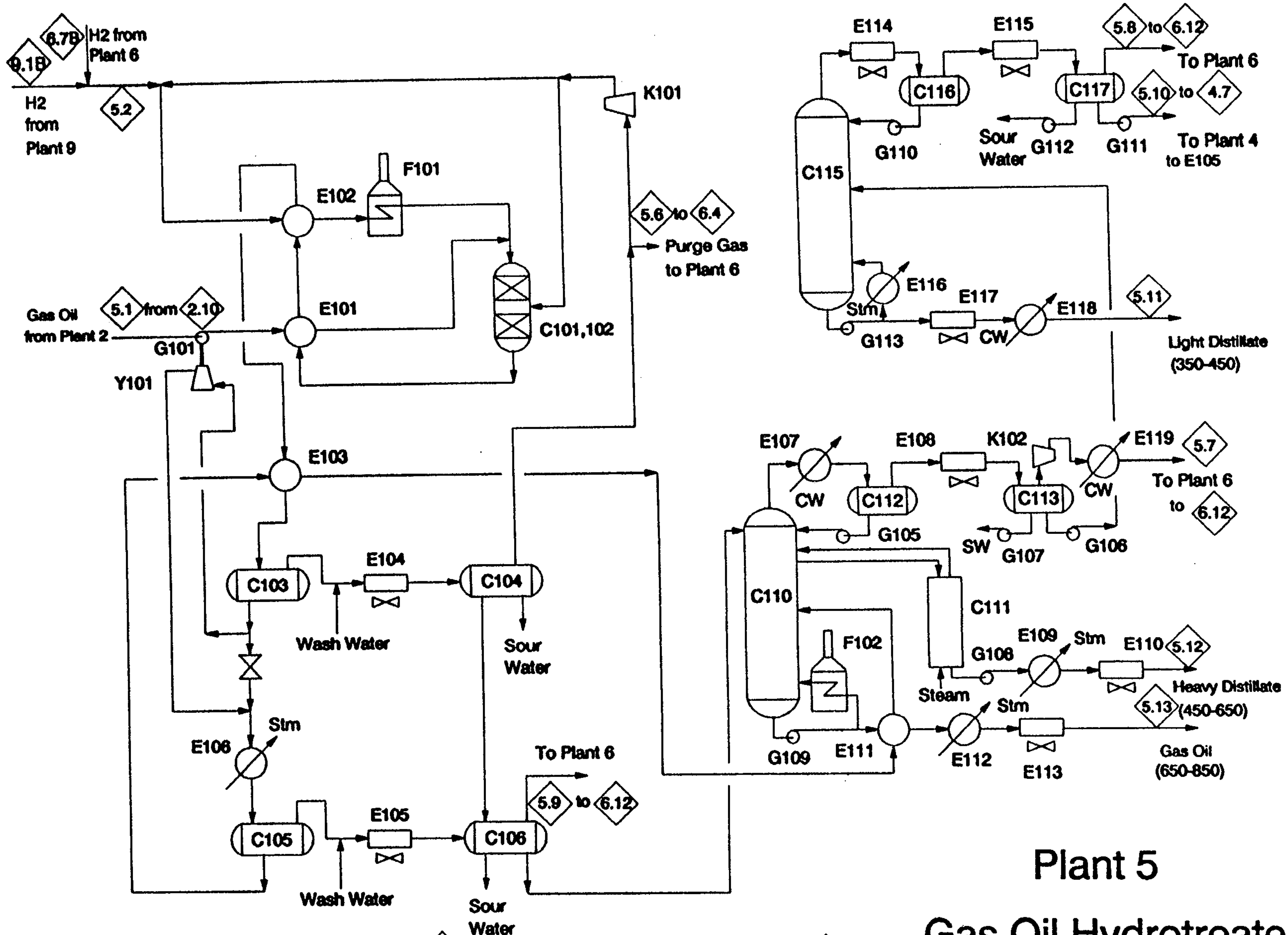
This Plant consists of two sections. These are: 1) Reaction and Separation section and 2) Product Distillation section. Figure 11.1, the Process Flow Diagram, schematically depicts the process. The process description is broken down in two parts according to the above two sections.

Reaction and Separation Section

Cold feed is pumped from intermediate tankage through heat exchange with the sidestream product of the atmospheric tower of the Coal Liquefaction Plant. Hot feed is pumped from intermediate tankage directly to the Gas Oil Hydrotreater and is mixed with the preheated cold feed. The mixed feed is pumped up to reactor pressure in the reactor charge pump (G101) driven by an hydraulic expander turbine. The mixed feed is sent through heat exchange with the reactor effluent in E101. Recycle and makeup hydrogen are mixed and heated first in heat exchange with the reactor effluent (E102) and then in a fired heater (F101).

The hydrogen is then mixed with the heated feed prior to entering the fixed bed reactors (C101 and C102), which operate at an inlet pressure of 2600 psig. The reactor effluent passes through sequential heat exchange with the cold feed (E101), hydrogen gas (E102), and the hot low pressure separator liquid (E103) before being sent to the hot high pressure separator (C103). The vapor from the hot high pressure separator is cooled in an air-fin exchanger (E104) and is sent to the cold high pressure separator (C104). Wash water is injected into the reactor effluent line prior to the air-fin cooler for control of ammonium salts. Hydrogen rich vapor from the cold high pressure separator is compressed by K101 and returned to the front end of the unit to mix with makeup hydrogen.

A purge hydrogen stream is sent to Plant 6 to remove light hydrocarbons from the system. Sour water is removed from the cold high pressure separator to be eventually sent for ammonia recovery. The hydrocarbon liquid is sent to the cold low pressure separator (C106). The liquid from the hot high pressure separator is sent to an hydraulic expander turbine (Y101) driving the reactor charge pump before going to the hot low pressure separator (C105). The hot low pressure separator liquid passes through heat exchange with the reactor effluent (E103) and is then sent to product distillation. The hot low pressure separator vapor is cooled in an air-fin exchanger (E105) and sent to the cold low pressure separator (C106). The vapor from this separator is sent to Plant 6, and the liquid is sent to product distillation. Sour water is removed from the cold low pressure separator to be eventually sent for ammonia recovery.



Notes: (1) Combined wash water streams are represented by 5.3.
(2) Combined sour water streams are represented by 5.5.
(3) Combined steam flow rate is represented by 5.4.

Plant 5

Gas Oil Hydrotreater

Figure 11.1 Process Flow Diagram

Product Distillation

Liquid from the hot low pressure separator is preheated in exchange with the product fractionator bottoms in E111 and charged as the lower feed to the fractionator (C110). Liquid from the cold low pressure separator is the upper feed to the fractionator. The overhead of the tower has three products: an overhead vapor stream which is sent to Plant 6; an unstabilized liquid product; and a water stream which is recycled to the wash water system. The overhead liquid is sent to an outboard stabilizer (C115) for correcting the initial boiling point of the 350-450°F product. The overhead of the stabilizer has a vapor product which is sent to Plant 6 and a liquid product which is sent to the Naphtha Hydrotreater (Plant 4) product distillation section. The stabilized 350-450°F product is sent to final storage. The sidestream off the main fractionator is steam-stripped (C111) to correct the initial boiling point, cooled, and sent to finished product storage (450-650°F). The bottoms product is sent to heat exchange with the lower feed to the fractionator (E111), cooled in E113, and sent to finished product storage (650-850°F).

11.2 Material Balance

The material balance for this plant is shown in Table 11.1. The plant consists of two trains. The material balance shown in the table is for one train only.

PLANT 5
MATERIAL BALANCE

TABLE 11.3

PLANT INPUT

Stream No.	5.1	5.2	5.4	5.3		
COMPONENTS	TT PLT 2 GAS OIL #/hr	H2 FROM #9 & #6 #/hr	STEAM #/hr	WASH WATER #/hr	TOTAL INPUT #/hr	RXN DELTA #/hr
N2	0	73			73	
H2	0	9245			9245	-5146
H2O	0	0	2500	18191	20691	2355
H2S	0	0			0	115
NH3	0	0			0	879
CO	0	0			0	
CO2	0	0			0	
C1	0	24			24	4092
C2	0	0			0	3827
C3	0	0			0	5937
C4	0	0			0	8391
C5-350	2745	0			2745	17956
350-450	39645	0			39645	11494
450-650	158385	0			158385	-8036
650-850	134745	0			134745	-45965
850-1000	0	0			0	4101
1000+	0	0			0	
PHENOLS	0	0			0	
MEA		0			0	
UNCONVERTED COAL		0			0	
ASH		0			0	
TOTAL	335520	9341	2500	18191	365552	0
TEMPERATURE (F)	299	100	650	*		
PRESSURE (psig)	50	2690	50	*		

PLANT OUTPUT

Stream No.	5.6	5.7-5.9	5.15	5.5	5.10	5.11	5.12	5.13	
COMPONENTS	H2 PURGE #/hr	TT VAP TO PLT 6 #/hr	SW VAP TO PLT 6 #/hr	TT SOUR WATER #/hr	STB OH TO PLT 4 #/hr	LT DIST #/hr	HVY DIST #/hr	GAS OIL #/hr	TOTAL OUTPUT #/hr
N2	51	22							73
H2	3313	784	2	0					4099
H2O	396	674	3	21880			94		23047
H2S	1	4		110					115
NH3	28	57	2	793					879
CO		0							0
CO2		0							0
C1	2935	1175	3	2	1				4116
C2	1937	1867	1	1	2		19		3827
C3	2102	3784	1	1	49				5937
C4	1944	5992	1		455				8391
C5-350	1166	6590	1	5	9573	3353	14		20701
350-450	11	85		71	763	45660	4549		51139
450-650		1		13		925	133147	16263	150349
650-850		0					12319	76461	88780
850-1000		0					2	4099	4101
1000+		0							0
PHENOLS		0							0
MEA		0							0
UNCONVERTED COAL		0							0
ASH		0							0
TOTAL	13884	21034	12	22876	10843	49938	150144	96823	365553
TEMPERATURE (F)	135	*		135	135	110	135	135	
PRESSURE (psig)	2690	*		-	10	80	20	20	

* This stream represents a combination of input/output streams of the plant; a single temperature and pressure cannot be reported.

11.3 Major Equipment List

Table 11.4 includes the major equipment list for Plant 5 (Gas Oil Hydrotreater).

Table 11.4 Major Equipment List

PLANT 5 -- GAS OIL HYDROTREATER

Reactors and Vessels

Equipment No.	Equipment Description
5.1- C101	Hydrotreater Reactor
5.1- C102	Hydrotreater Reactor
5.1- C103	Hot High Pressure Separator
5.1- C104	Cold High Pressure Separator
5.1- C105	Hot Low Pressure Separator
5.1- C106	Cold Low Pressure Separator
5.1- C107	Wash Water Drum
5.1- C108	Sour Water Drum
5.1- C109	H2 Recycle Comp K.O. Drum
5.2- C112	Fractionator Ovhd Accum (Hot)
5.2- C113	Fractionator Ovhd Accum (Cold)
5.2- C114	Frac Ovhd Vent Comp K.O. Drum
5.2- C116	Stabilizer Ovhd Accum (Hot)
5.2- C117	Stabilizer Ovhd Accum (Cold)

Towers

Equipment No.	Equipment Description
5.2- C110	Fractionator
5.2- C111	AGO Product Stripper
5.2- C115	Distillate Product Stabilizer

Fired heaters

Equipment No.	Equipment Description
5.1- F101	Hydrogen Preheater
5.2- F102	Fractionator Reboiler

Table 11.4 Major Equipment List - continued

Heat Exchangers

Equipment No.	Equipment Description
5.1- E101	Reactor Effl/Feed Preheater
5.1- E102	Reactor Effl/H2 Preheater
5.1- E103	Reactor Effl/HLPS Liquid
5.1- E104	Hot HP Sep Vapor Cooler
5.1- E105	Hot LP Sep Vapor Cooler
5.1- E106	Hot HP Sep Liq Steam Gen
5.2- E107	Frac Ovhd Cond
5.2- E108	Frac Ovhd Vapor Cond
5.2- E109	AGO Prod Steam Gen
5.2- E110	AGO Prod Cooler
5.2- E111	Frac Bottoms/Feed Ex
5.2- E112	Frac Bottoms Steam Gen
5.2- E113	Frac Bottoms Cooler
5.2- E114	Distillate Stabilizer Cond
5.2- E115	Dist Stabil Ovhd Vap Cooler
5.2- E116	Distillate Stabil Reboiler
5.2- E117	1st Distillate Prod Cooler
5.2- E118	2nd Distillate Prod Cooler
5.2- E119	Frac Ovhd Vent Comp Aftercooler

Compressors

Equipment No.	Equipment Description
5.1- K101	Recycle Hydrogen Compressor
5.2- K102	Fractionator Ovhd Vent

Table 11.4 Major Equipment List - continued

Pumps/Expander

Equipment No.	Equipment Description
5.1- G101	Reactor Charge Pump
5.1- G102	Low Press Wash Water Pump
5.1- G103	High Press Wash Water Pump
5.1- G104	Sour Water Drum Outlet Pump
5.2- G105	Fractionator Reflux Pump
5.2- G106	Dist Stabilizer Feed Pump
5.2- G107	Frac Ovhd Sour Water Pump
5.2- G108	AGO Product Pump
5.2- G109	Fractionator Bottoms Pump
5.2- G110	Dist Stabilizer Reflux Pump
5.2- G111	Stabilizer Ovhd Liquid Prod Pump
5.2- G112	Stabilizer Ovhd Sour Water Pump
5.2- G113	Distillate Prod Pump
5.1- Y101	Hot HP Sep Liquid Expander

11.4 Utility Summary

The utility requirement per train for this plant is summarized in Table 11.5.

Table 11.5

Utility Summary

Steam, lbs/hr		
	150 psig	—16,932 lbs/hr
	600 psig	32,524 lbs/hr
Electricity		1090 KW
Fuel Gas		1.9 MMSCFD
Cooling Water		4,114 gpm

11.5 Water Summary

Table 11.6 below presents the water balances for the Gas Oil Hydrotreater.

Table 11.6

WATER SUMMARY (Total for 2 Trains)

Water Inlet	lb/hr
Produced in Reactors	4,710
Wash Water	40,742
Steam	5,000
TOTAL	50,452
Water Outlet	
Sour Water to NH_3 Recovery	43,760
Recycled to Wash Water	4,360
Hydrogen Purge	791
Low Pressure Gases to Plant 6	1,353
Products	188
TOTAL	50,452
Wash Water Summary	
Requirements	40,742
Separator Water Recycled	4,360
Makeup Wash Water	36,382
Sour Water to NH_3 Recovery, gpm	88

12. Plant 6 (Hydrogen Purification)

12.0 Design Basis Criteria and Considerations

A centralized hydrogen purification unit is being proposed. The unit will have two major sections:

1. Recovery of hydrogen from high pressure purge gas from the five coal liquefaction plants and gas oil hydrotreater.
2. Recovery of hydrogen from lower pressure purge gas from five coal liquefaction plants, the gas plant and hydrotreaters.

The first section will take sour purge gas from the coal liquefaction plants and gas oil hydrotreater. The hydrogen-rich gas will be water-washed and amine-treated to remove ammonia, carbon dioxide, and hydrogen sulfide. Part of the scrubbed gas will be recycled to plant 2 and the rest will be sent to membrane units for hydrogen recovery. This process was selected because the pressure of the hydrogen product is much higher than for a pressure swing absorber, thus saving recompression costs (both capital and operating). The hydrogen product at a minimum purity level of 99.0 mol% and H_2 product from PSA unit will be compressed to required pressure. The nonpermeate product will be sent to the PSA unit.

The low pressure gas from the overhead of the Plants 2, 3, 4 and 5 is treated and compressed and then sent to the pressure swing adsorber. Pressure swing absorption will be used for this service because of the high recovery of hydrogen and because the feed gas pressure is consistent with the point of maximum recovery of hydrogen. The hydrogen product at a minimum of 99.0 mol% purity will be compressed and combined with the hydrogen product from the membrane unit. The light hydrocarbon tail gas stream will be sent to the Gas Plant.

The original flow scheme for the Hydrogen Purification Plant was for the compression and treating of the low pressure gases be done in the Gas Plant with the absorber/deethanizer overhead stream being sent to Plant 6 for hydrogen recovery in the pressure swing absorption (PSA) unit. Because the low pressure gases contained a high percentage of hydrogen and the resultant vapor-liquid equilibrium considerations, it was impossible to obtain an acceptable recovery of propane and butanes in the absorber/deethanizer. Therefore, the flow scheme was switched to place the PSA unit before the absorber/deethanizer. This then caused the move of the compression and treating of the low pressure gases to Plant 6.

The high pressure hydrogen purge streams were sent to a membrane unit for hydrogen recovery rather than a PSA unit because the membrane can better accept a high pressure feed stream, resulting in a higher pressure hydrogen product. This saved hydrogen compression costs. The PSA unit was preferred in the low pressure service because of the higher hydrogen recovery and lower pressure drop in the system. This also saved hydrogen compression costs.

The first stage hydrogen compression on the PSA product exited at the outlet pressure of the membrane unit. The two hydrogen streams were combined before being compressed to the required pressure. This saved capital costs.

12.1 Process Description and Process Flow Diagram

The Hydrogen Purification Plant has two sections: (1) recovery of purified hydrogen from high pressure purge gas streams in membrane separation units, and (2) recovery of purified hydrogen from low pressure offgas streams in pressure swing absorption (PSA) units. These are designated as Plant 6.1 and 6.2, respectively. Figure 12.2 is the process flow diagram for Plant 6.1 where Figures 12.2 and 12.3 combined show the process flow diagram for Plant 6.2.

Plant 6.1 High Pressure Purge Gas by Membrane

Purge Gas Scrubbing. The portion of the cold high pressure separator vapor which is purged from the Coal Liquefaction Plants is combined with H_2 product from gas oil hydrotreater and is sent to Plant 6. The sour gases are sent first to a water wash tower (C104) to remove ammonia. The ammonia water is sent to Plant 38 for the recovery of anhydrous ammonia. The gas stream exiting the water wash tower is sent to the amine treater (C109) for the removal of carbon dioxide and hydrogen sulfide. The rich amine is sent to a common amine regeneration system located in the low pressure recovery section of the plant. The capacity for the water wash and amine treater towers are provided in two parallel trains for reliability.

Part of the treated hydrogen purge stream is recycled back to coal liquefaction and the rest is sent to the membrane unit.

Membrane Separation Unit. The scrubbed high pressure purge gas is sent to membrane separation units (V101) for hydrogen recovery. This process was selected because the pressure of the hydrogen and nonpermeate products is much higher than for a pressure swing absorber, thus saving recompression costs (both capital and operating). The hydrogen product at a minimum purity level of 99.0 mol% and the purified hydrogen streams from the PSA unit are combined and compressed in the second stage hydrogen compressor (K106), driven by an electric motor. The discharge of the second stage compressor is cooled in an air-fin exchanger (E111). Part of the discharge is sent to plant 4 as part of H_2 makeup. The other part is combined with the treated H_2 purge stream and is compressed to 3300 psig in K-108 and recycled to plant 2. The rest of E111 discharge is sent to K107 and is compressed to the pressure required by plant 5. The K107discharge stream is cooled in an air-fin exchanger (E112) and with cooling water (E113) before exiting the plant to the makeup hydrogen system. The nonpermeate product is sent to the PSA units for recovery of residual hydrogen.

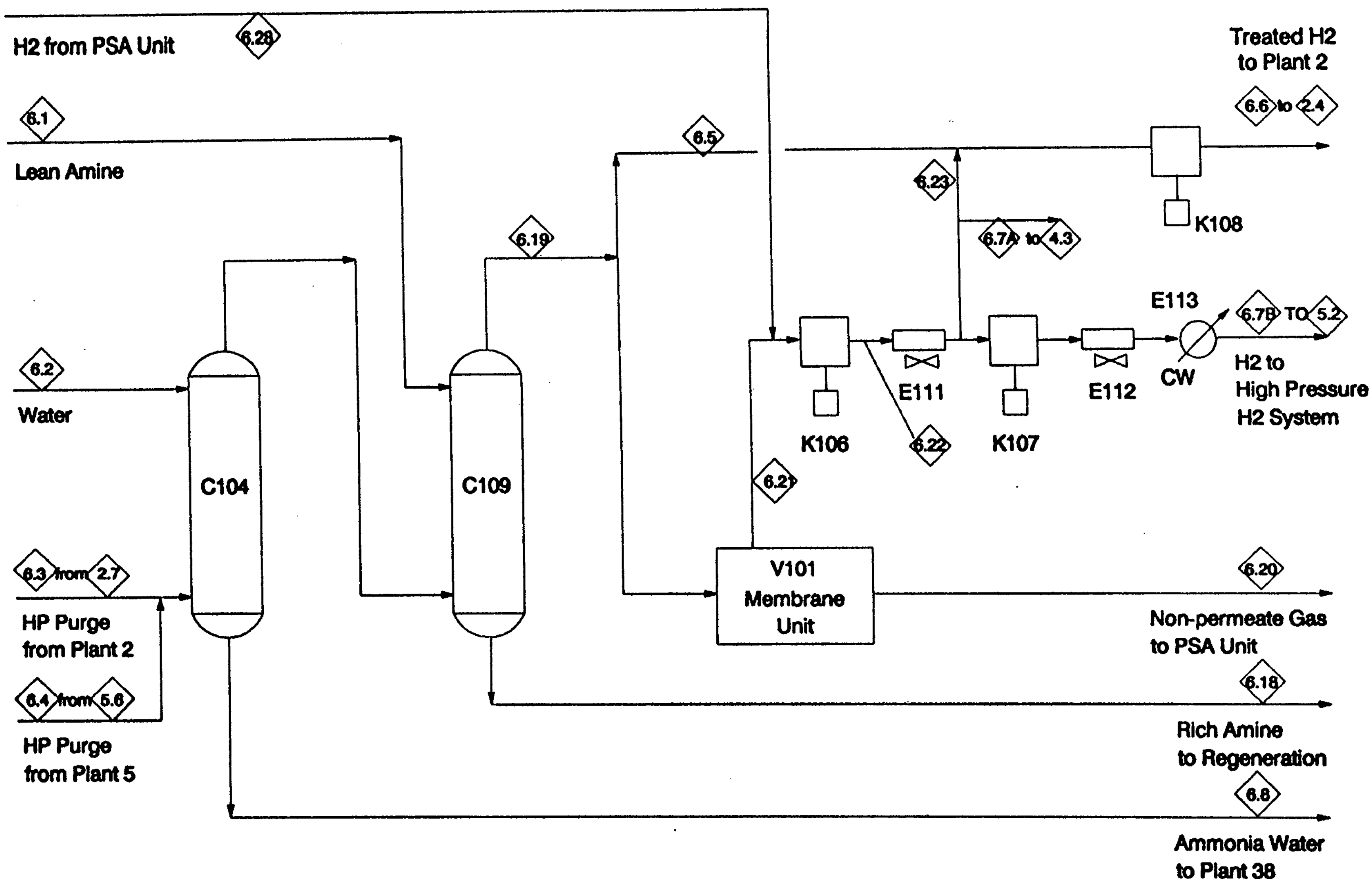
Plant 6.2 Low Pressure Offgas by PSA

Low Pressure Gas Compression. Miscellaneous gas streams are recovered from Coal Liquefaction (Plant 2), Gas Plant (Plant 3), the Naphtha Hydrotreater (Plant 4), and the Gas Oil Hydrotreater (Plant 5). The low pressure gases (less than 180 psig) are sent to the Low Pressure Gas Compressor (K101). The compressor will have a suction pressure of 5 psig and an intermediate suction of 40 psig. The discharge of the first stage compressor will be cooled in air-fin exchanger E101. The second stage compressor (K102) discharge will be at 250 psig. The discharge from the compressor will be cooled in air-fin exchanger E102.

Low Pressure Gas Scrubbing. Gases after compression are sent first to a water wash tower (C103) to remove ammonia. The ammonia water is sent to Plant 38 for the recovery of anhydrous ammonia. The gas stream exiting the water wash tower is sent to the amine treater (C106) for the removal of carbon dioxide and hydrogen sulfide. The rich amine is sent to a common amine regeneration system. The capacity for the water wash and amine treater towers are provided in two parallel trains for reliability.

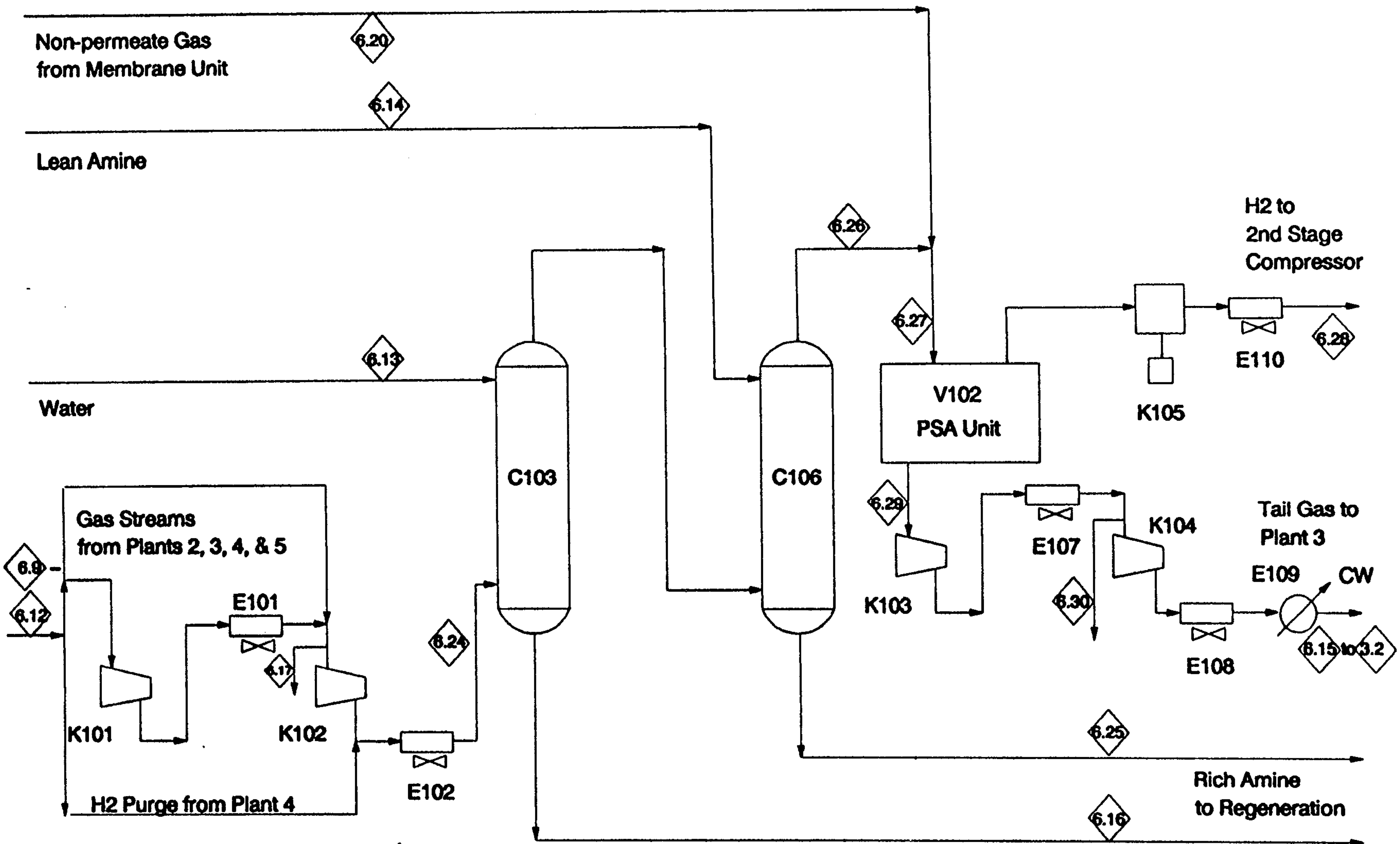
Pressure Swing Absorption Unit. This section of the plant is fed the low pressure gas (200 psig) from the overhead stream from the amine treater and the membrane nonpermeate product. Pressure swing absorption (PSA) will be used for this service because of the high recovery of hydrogen. The hydrogen product is at a minimum purity of 99.0 mol %. The product from the PSA unit (V102) is compressed from 190 psig to 500 psig in motor-driven reciprocating compressors (K105). This first stage of compression takes the lower pressure stream up to the product pressure of the membrane separation unit. The first stage discharge is cooled in an air-fin exchanger (E110) and will be sent to Plant 6.1 for further compression. The PSA tail gas stream will be compressed to 200 psig in a two-stage compressor (K103 and K104), cooled by intercooler E107 and aftercoolers E108 and E109, and sent to the Gas Plant for recovery of propane and heavier hydrocarbons.

Amine Regeneration. As shown in Figure 12.3, rich amine streams from the high pressure and intermediate pressure amine treaters are collected in the combined amine separator (C111). Flashed acid gases are sent to the Sulfur Plant (Plant 11). The rich amine is filtered (Y101) and preheated with the lean amine bottoms in exchanger E105 before entering the amine regenerator C112. The overhead acid gas stream is sent to Plant 11. The regenerated lean amine is cooled in E105 and enters the regenerated amine surge drum (C114). Lean amine is then cooled in exchanger E106, filtered (Y102 and Y103), and pumped back to the intermediate pressure amine treater by pump G101 and to the high pressure amine treater by pump G102.



Plant 6.1
High Pressure H2 Recovery

Figure 12.1 Process Flow Diagram



NOTE: 6.9 from 2.8 ; 6.10 from 3.3
6.11 from 4.1 + 4.2a + 4.2b
6.12 from 5.7 + 5.8 + 5.9 + 5.15

Plant 6.2
Low Pressure H2 Recovery

Figure 12.2 Process Flow Diagram

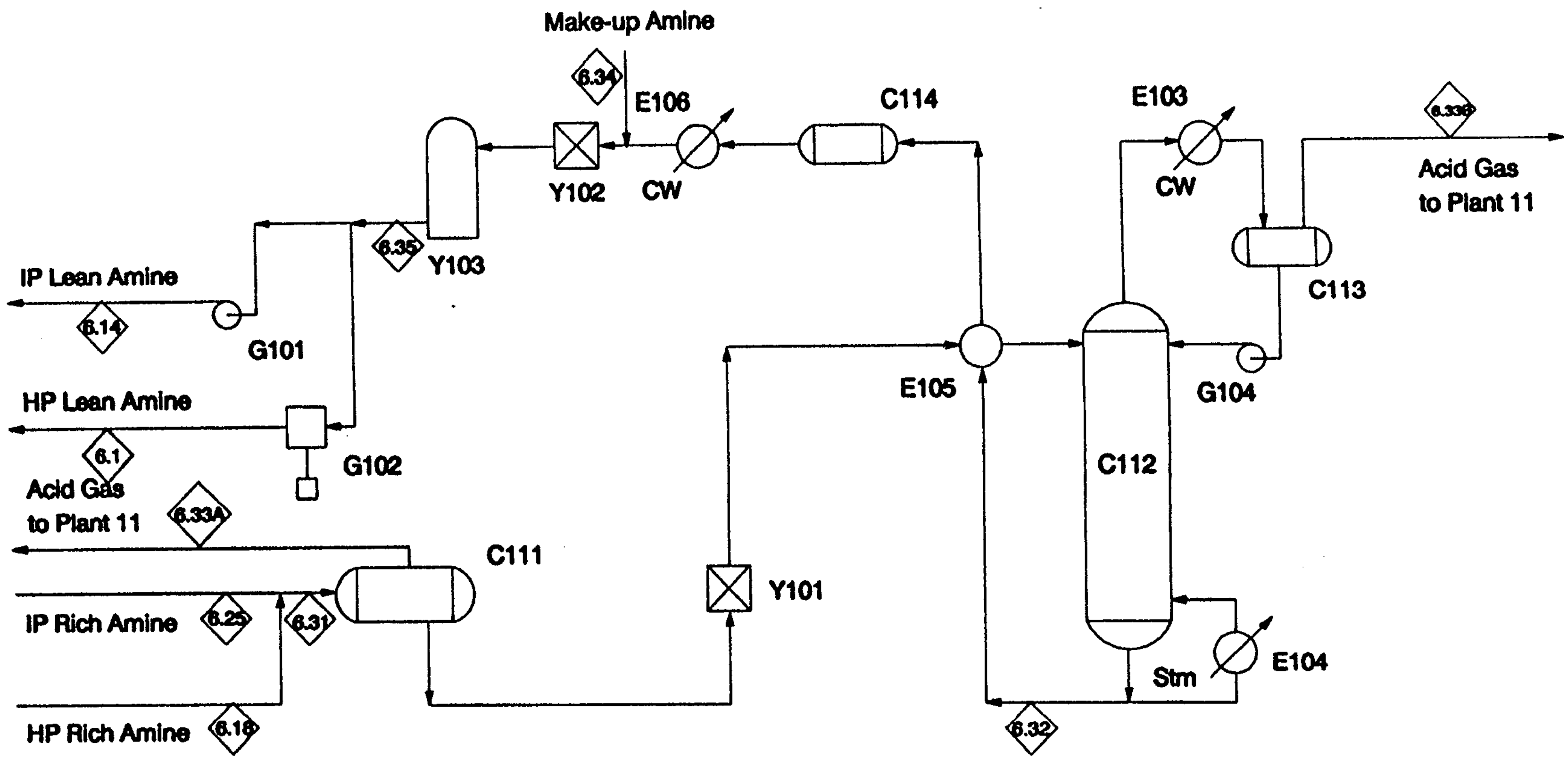


Figure 12.3 Process Flow Diagram

Plant 6.2
Amine Regeneration

12.2 Material Balance

The material balance for the plant is shown in Table 12.1.

Table 12.1 Material Balance

PLANT INPUT

Stream No.	6.3	6.4	6.2+6.13	6.9	6.10	6.11	6.12	6.34	
COMPONENTS	PLT 2 H2 PURGE #/hr	PLT 5 H2 PURGE #/hr	WATER #/hr	PLT 2 TO PSA #/hr	PLT 3 TO PSA #/hr	PLT 4 TO PSA #/hr	PLT 5 TO PSA #/hr	MAKEUP AMINE #/hr	TOTAL INPUT #/hr
N2	1024	102		439	0	20	44		1627
H2	17210	6626		4496	0	2008	1572		31912
H2O	1535	792	30000	4835	48	95	1353	1831	40487
H2S	10110	2		10212	115	0	8		20447
NH3	2415	56		202	0	90	116		2879
CO	677	0		290	0	0	0		967
CO2	865	0		401	0	0	0		1266
C1	18116	5870		12589	10	1097	2356		40038
C2	10035	3874		12179	135	745	3736		30704
C3	6575	4204		15160	780	847	7570		35136
C4	805	3888		8906	1710	3654	11984		30947
C5–350	425	2332		26743	1198	862	13180		44740
350–450	0	22		268	0	0	170		460
450–650	0	0		24	0	0	2		26
650–850	0	0		0	0	0	0		0
850–1000	0	0		0	0	0	0		0
1000+	0	0		0	0	0	0		0
PHENOLS	0	0		0	0	0	0		0
MEA	0	0		3	0	0	0	35	38
UNCONVERTED COAL	0	0			0	0	0		0
TOTAL	69791	27768	14000	96748	3994	9418	42090	1866	281675
TEMPERATURE (F)	130	134	*	*	*	*	*		
PRESSURE (psig)	2970	2450	*	10	10	10	10		

PLANT OUTPUT

Stream No.	6.8+6.16	6.7 **	6.6	6.17	6.15	6.33 ***	6.30	
COMPONENTS	NH3 H2O #/hr	H2 PROD TO #4,5 #/hr	TT H2 REC TO #2 #/hr	PSA COMP KO LIQ #/hr	PSA TAIL GAS to #3 #/hr	ACID GAS #/hr	TAIL GAS COMP KO LIQ #/hr	TOTAL OUTPUT #/hr
N2		0	349		1279	0		1627
H2		9230	20429		2253	0		31912
H2O	34938	0	508	255	3037	1749		40487
H2S	1583	0	8		44	18812		20447
NH3	2879	0	0		0	0		2879
CO	0	0	210		757	0		967
CO2	73	0	0		0	1193		1266
C1		54	7505		32476	0	3	40038
C2		0	4308		26384	0	13	30704
C3		0	3338		31749	0	49	35136
C4		0	1453		29375	0	118	30947
C5–350		0	853	5	41098	3	2781	44740
350–450		0	7	4	104	0	345	460
450–650		0	0		26	0		26
650–850		0	0		0	0		0
850–1000		0	0		0	0		0
1000+		0	0		0	0		0
PHENOLS		0	0		0	0		0
MEA		0	3		35	0		38
UNCONVERTED COAL		0	0		0	0		0
ASH		0	0		0	0		0
TOTAL	39473	9284	38970	264	168617	21757	3309	281675
TEMPERATURE (F)	*			126	95	*		
PRESSURE (psig)	*			45	210	*		

* This stream represents a combination of input/output streams; a single temperature/pressure cannot be reported.

** 6.7 = 6.7A + 6.7B

*** 6.33 = 6.33A + 6.33B

12.3 Major Equipment List

The major equipment list for the plant is shown in Table 12.2.

Table 12.2 Major Equipment List

PLANT 6 -- HYDROGEN PURIFICATION

Reactors and Vessels

Equipment No.	Equipment Description
6.2-C101	1st Stg PSA Feed Comp K.O.
6.2-C102	2nd Stg PSA Feed Comp K.O.
6.2-C105	IP Amine Trt Abs Inlet Sep
6.2-C107	IP Amine Trt Abs Outlet Sep
6.1-C108	HP Amine Trt Abs Inlet Sep
6.1-C110	HP Amine Trt Abs Outlet Sep
6.2-C111	Combined Amine Btms Sep
6.2-C113	Amine Regen Ovhd Accum
6.2-C114	Regen Amine Surge Drum
6.2-C116	1st Stg PSA TG Comp K.O.
6.2-C117	2nd Stg PSA TG Comp K.O.
6.2-C118	1st Stg H2 Comp K.O.
6.1-C119	2nd Stg H2 Comp K.O.
6.1-C120	3rd Stg H2 Comp K.O.
6.2-C115	Makeup Amine Tank
6.2-C103	IP Water Wash Tower
6.1-C104	HP Water Wash Tower
6.2-C106	IP Amine Absorber
6.1-C109	HP Amine Absorber
6.2-C112	Amine Regenerator

Heat Exchangers

Equipment No.	Equipment Description
6.2-E101	PSA Feed Comp Intercooler
6.2-E102	PSA Feed Comp Aftercooler
6.2-E103	Amine Regen Condenser
6.2-E104	Amine Regenerator Reboiler
6.2-E105	Lean/Rich Amine Exchanger
6.2-E106	Lean Amine Cooler
6.2-E107	PSA TG Comp Intercooler
6.2-E108	PSA TG Comp 1st Aftercooler
6.2-E109	PSA TG Comp 2nd Aftercooler
6.2-E110	1st Stg H2 Comp Intercooler
6.1-E111	2nd Stg H2 Comp Intercooler
6.1-E112	3rd Stg H2 Comp 1st Afterclr
6.1-E113	3rd Stg H2 Comp 2nd Afterclr

Table 12.2 Major Equipment List - continued

Compressors

Equipment No.	Equipment Description
6.2-K101	1st Stg PSA Feed Compressor
6.2-K102	2nd Stg PSA Feed Compressor
6.2-K103	1st Stg PSA TG Compressor
6.2-K104	2nd Stg PSA TG Compressor
6.2-K105	1st Stage H2 Compressor
6.1-K106	2nd Stage H2 Compressor
6.1-K107	3rd Stage H2 Compressor
6.1-K108	H2 Recycle Compressor

Pumps

Equipment No.	Equipment Description
6.2-G101	IP Amine Pump
6.2-G102	HP Amine Pump
6.2-G103	Amine Makeup Pump
6.2-G104	Amine Regen Reflux Pump

Miscellaneous

Equipment No.	Equipment Description
6.2-Y101	Rich Amine Mech Filter
6.2-Y102	Lean Amine Mech Filter
6.2-Y103	Lean Amine Carbon Filter
6.1-V101	Membrane Unit
6.2-V102	PSA Unit

12.4 Utility Summary

The utilities for this plant are summarized in Table 12.3.

Table 12.3

Utility Summary

	Plant 6.1	Plant 6.2
Steam (50 psig)	4042 lbs/hr	55,431 lbs/hr
Cooling Water	1400 gpm	5262 gpm
Water (Makeup)	28 gpm	32 gpm
Electricity	24,103 KW	19,841 KW

12.5 Water Summary

The following table presents the water summary for Plant 6.

	High Pres. Water Wash	Compr. Suction K.O. Liquid	Inter. Pres. Water Wash
Makeup Water, gpm	28	0	32
Sour Water, lb/hr			
H_2O	14,671	255	20,267
H_2S	780	-	803
NH_3	2471	-	408
CO_2	64	-	9
HC	0	3318	0
Total	17,986	3573	21,487
Sour Water, gpm	33	5	43
Disposition of Sour Water	Plant 38	Plant 4	Plant 38

13. Plant 8 (Critical Solvent Deashing Unit - ROSE-SR)

13.0 Design Basis, Criteria and Considerations

Feed Rate, BPSD 50,800

Component	Composition, wt% Feed [(1)]	Ash Concentrate [(2)]
Ash	17.60	40.50
Carbon	73.80	51.66
Hydrogen	6.21	3.54
Nitrogen	0.78	0.83
Sulfur	1.61	3.47
Total	100.00	100.00

(1) Contains unconverted coal 9.8 wt% and solvent 0.9 wt%
(2) Contains unconverted coal 24.2 wt% and solvent 3.1 wt%

13.1 Process Description and Process Flow Diagram

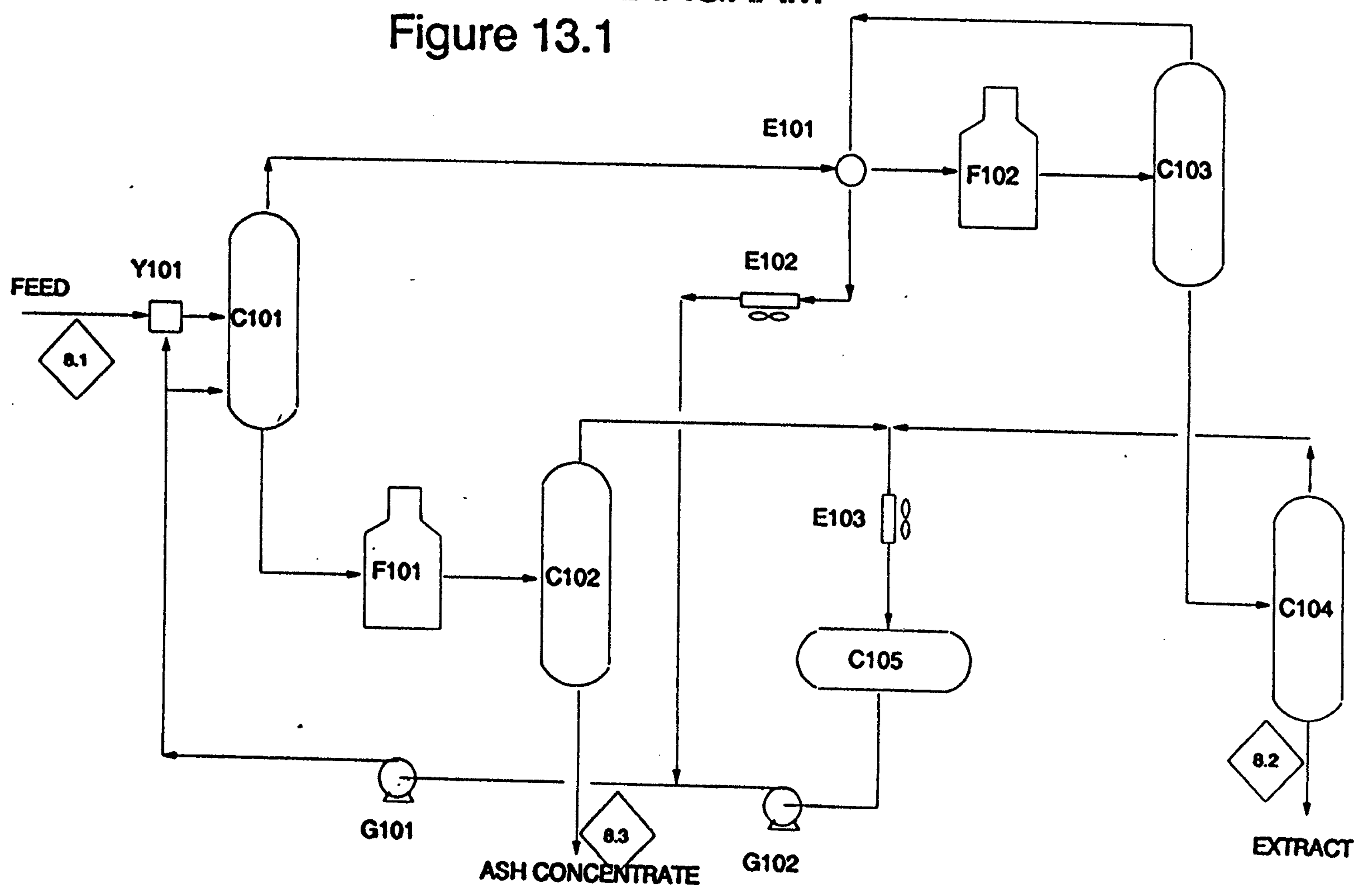
The simple operation of the ROSE-SR process is illustrated schematically on the Process Flow Sketch of Plant 8. The feedstock is charged through a mixer Y101, where it is contacted with light hydrocarbon solvent at elevated temperature and pressure. A portion of the solvent is mixed with the feed in Y101, and the remainder of the solvent enters through a separate solvent inlet that provides countercurrent solvent flow. The mixture passes to the asphaltene separator, C101, where the heavy asphaltene/solids fraction drops out of solution and is withdrawn as a liquid from the bottom of the separator. The asphaltene and residual solvent pass through heater F101 to flash tower C102, where the solvent is flashed and stripped from the product and the asphaltene product comes out as ash concentrate.

The solvent-deashed oil phase flows from the top of C101 through heat exchanger E101, where it is heated by the circulating solvent before entering heater F102. The temperature is elevated above the critical temperature of the solvent before the solvent-deashed oil mixture enters the deashed oil separator.

At this increased temperature, the deashed oil (DAO) is virtually insoluble in the solvent. Therefore, a phase separation occurs in the DAO separator, C103. The DAO phase is withdrawn from the bottom of C103 and stripped of its solvent in tower C104. The ROSE extract thus produced comes out of tower C104. The substantially DAO-free supercritical solvent provides a significant part of the enthalpy needed to increase the temperature of the overhead stream from C101 to the conditions in C103. This circulating solvent flows overhead from C103 through exchanger E101, giving up heat to the solvent-DAO solution from C101. The circulating solvent is further cooled in exchanger E102 before returning to C101 via the solvent circulation pump G101. In many applications, E102's process duty is negligible, though the exchangers are sized to handle operational upsets such as an increased charge temperature. E102 is overdesigned to enable the ROSE unit to handle upsets in charge temperature. The solvent circulation pump G101 develops only the differential pressure necessary to overcome the separator system pressure drop.

The small amount of solvent dissolved in the heavy phases from vessels C101 and C103 (between 7% and 15% of the total extraction solvent) is recovered by conventional stripping in C102 and C104. The recovered solvent is condensed in E103 and then collected in the solvent surge drum C105. This small flow of solvent is pumped by the recycle solvent pump G102 into the large flow of high-pressure circulating solvent upstream of the solvent circulation pump, G101.

PROCESS FLOW DIAGRAM
Figure 13.1



PLANT 8
ROSE PROCESS

13.2 Material Balance

The overall material balance for the Critical Solvent Deashing Unit is presented in Table 13.1.

Table 13.1

Material Balance

	lbs/hr			
Stream No.	8.1	8.2	8.3	
	Feed	Extract	Ash Concentrate	
H_2O	0	0	0	
850°F-	380	380	0	
850-1000°F	10,095	1,765	8,330	31 %
1000°F+	642,440	536,585	105,855	
Unconverted Coal	90,965	----	90,965	69 %
Ash	163,510	----	163,510	
Total	907,390	538,730	368,660	
Tons/day	10,889	6,465	4,424	

13.3 Major Equipment List

The major equipment list for the plant is shown in Table 13.2 below:

Table 13.2

Equipment Number	Equipment Description
8 - C101	Asphaltene Separator
8 - C102	Flash Tower
8 - C103	DAO Separator
8 - C104	Solvent Stripper
8 - C105	Solvent Surge Drum
8 - E101	Solvent/DAO Solution Exchanger
8 - E102	Solvent Cooler
8 - E103	Solvent Condenser
8 - G101	Solvent Circulation Pump
8 - G102	Recycle Solvent Pump
8 - F101	Ash Concentrate Heater
8 - F102	Extract Heater
8 - Y101	Feed/Solvent Mixer

Due to the proprietary nature of this unit, Kerr-McGee did not provide the above major equipment list and/or major equipment sizes. The above list is a non-confidential version prepared by Bechtel for the reader's convenience.

13.4 Utility Summary

The utility summary for this plant is shown in Table 13.3.

Table 13.3

Utility	Units	Rate
Steam Consumption - 150 psig	(lbs/hr)	25,419
Fuel Gas - Fired	(MMBtu/hr) (MMSCFD)	212 5.1
Electricity	(kW)	4,363

Note: Fuel gas rate in MMSCFD is based on a gross heating value of 1,000 Btu/SCF.

13.5 Water Summary

There is essentially no water used in this process, however, stripping steam produces 25,419 lbs/hr of waste water.

14. Plant 9 (Hydrogen Production by Coal Gasification)

14.0 Design Basis, Criteria and Considerations

Plant 9 is designed to gasify 4,424 tons/day of Ash Concentrate from ROSE-SR Plant and 6,127 tons/day of MF coal. Syngas from the gasifiers is used to produce 416 MMSCFD of 99.9 volume percent hydrogen product as required by the coal liquefaction and hydrotreating plants. The hydrogen poor purge stream from this gasifier is used as medium Btu fuel gas.

Feed Streams

- 4,424 TPD ash concentrate from Plant 8 (ROSE-SR)
- 6,127 TPD clean MF coal from Plant 1 (Coal Preparation)
- 8,962 TPD of 99.5 volume percent oxygen from Plant 10 (Air Separation)

Feed analysis and compositions are presented in Table 14.1.

Product Streams

- 416 MMSCFD of 99.9 volume percent hydrogen
- 94 MMSCFD of medium Btu fuel gas
- 27 MMSCFD of acid gas containing H_2S as feed to Sulfur Plant (Plant 11)
- 414 MMSCFD of stripped CO_2 waste gas for discharge to the atmosphere
- 2812 TPD (dry basis) of slag and soot to landfill disposal

Product compositions are presented in Table 14.2.

Gasifier Yields

- For ash concentrate, 33.4 SCF of CO + H_2 and 8 SCF of CO_2 per pound of dry and ash-free feed
- For clean coal, 31.5 SCF of CO+H_2 and 7.2 SCF of CO_2 per pound of dry and ash-free feed
- Average about 97% carbon conversion to syngas. Unconverted carbon is rejected with slag

TABLE 14.1

GASIFICATION FEED STREAMS

Illinois No.6 Coal

Ultimate Analysis, %(wt.), Dry Basis

Carbon	71.0
Hydrogen	4.8
Nitrogen	1.4
Sulfur	3.2
Oxygen	8.1
Ash	11.5
HHV, Btu/lb	10,951

Ash Mineral Analysis, % (wt.)

SiO_2	49.8
Fe_2O_3	17.6
Al_2O_3	19.2
TiO_3	1.0
CaO	6.3
MgO	1.0
SO_3	2.9
K_2O	2.0
Na_2O	0.5
P_2O_5	0.2
Undetermined	-0.5

Ash Concentrate

Composition, % (wt.)

850-1000°F	2.3
1000°F+	28.7
Unconverted Coal	24.7
Ash	44.3

HHV, Btu/lb 8,457

Ultimate Analysis, % (wt.), Dry Basis

Carbon	48.3
Hydrogen	2.9
Nitrogen	1.1
Sulfur	3.1
Oxygen	0.3
Ash	44.3

TABLE 14.2

PRODUCT DATA

	H_2 Product Mol %	Medium Btu Gas Mol%	H_2S-Rich Off-Gas Mol.%	CO_2 Off-Gas Mol %
H_2O	0.0	0.0	0.45	1.81
H_2	99.90	85.11	2.79	0.43
CO	0.0	6.95	2.19	0.09
CO_2	0.0	2.33	64.49	89.91
$AR+N_2$	0.10	5.16	1.51	7.75
C_1	0.0	0.45	0.01	0.01
H_2S	0.0	0.0	28.50	0.0
COS	0.0	0.0	0.06	0.0
Total				
#-Mol/hr	45,676	10,309	3,007	45,444
MMSCFD	416	94	27	414

14.0.1 Technology Selection

Two coal gasification technologies, the Texaco and Shell processes, have been evaluated for synthesis gas production in Plant 9. The block flow diagrams for these two processes are presented on Figure 14.1.

The major differences between these two processes are as follows:

	Texaco	Shell
1.	Feed Type	
	Molten coal or coal slurry	Dried and pulverized coal to 70-90% through 200 mesh
2.	Operating Conditions	
	High pressure process. Pressure up to 1100 psig	Low to medium pressure process. Pressure up to 400 psig
3.	Gasifier Construction	
	Vertical cylindrical pressure vessel with refractory lining	Horizontal ellipsoidal vessel. Gasifier shell has a double-walled construction, the inner shell is refractory lined
4.	Heat Removal Methods	
	Direct quench syngas with water in the gasifier and gas scrubber	Recover heat by generating steam in the waste heat boilers
5.	Shift Conversion	
	Produce sufficient steam in the syngas for shift reaction	Steam injection is required for shift reaction to occur
6.	Hydrogen Production	
	Produces higher H_2/CO ratio (0.75) syngas. Less shift conversion required.	Produces low H_2/CO ratio (0.43) syngas which requires more shift catalyst for shift conversion to reach the same H_2 production.

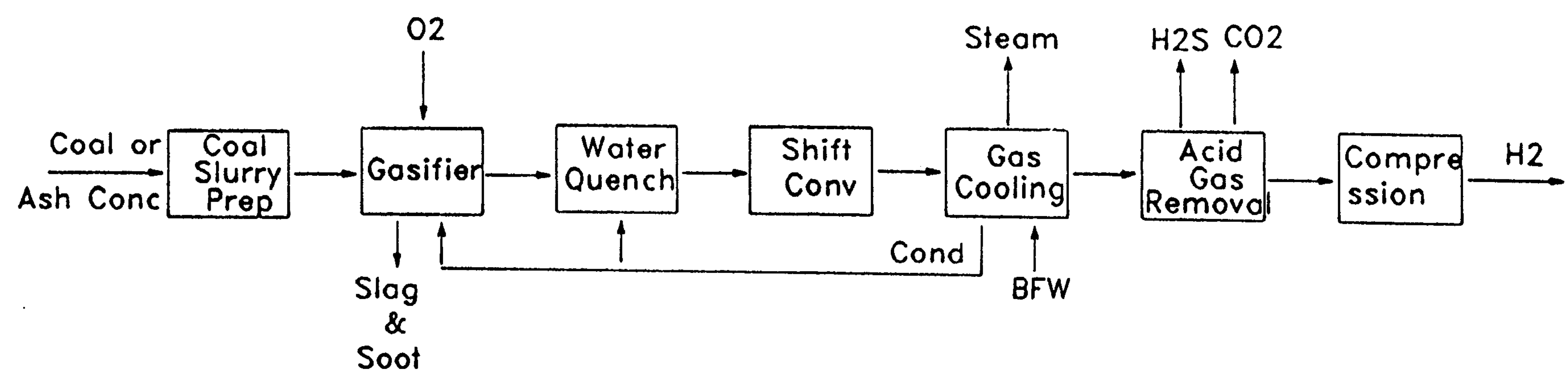
	Texaco	Shell
7.	Horsepower for Hydrogen Compression	
	Higher pressure process. Less horse-power required	Low pressure process requires more horsepower for hydrogen compression
8.	Water Requirement	
	Relatively poor quality condensate from the gas cooling sections can be used for direct quenching of syngas	More boiler feed water of higher quality is required for steam generation in the gasifier and waste heat boiler
9.	Ability to Handle ROSE bottoms	
	Its pilot plant had demonstrated that the ash-containing residues obtained from the H-Coal liquefaction plant were process efficiently	No pilot plant result has been reported for H-Coal liquefaction residues
10.	Commercial Plant Experience	
	Several large size commercial plants have been operated success-fully since 1980's	Only small demonstration plants have been built

For high pressure hydrogen production, Texaco's gasification process has lower capital and utility costs, and has showed it can process H-Coal liquefaction vacuum tower bottoms, as well as ROSE-SR ash concentrate, successfully. (Reference: Texaco report DE-84-013199, February 1984). Therefore, the Texaco technology is recommended for this project as the gasification process for Plant 9.

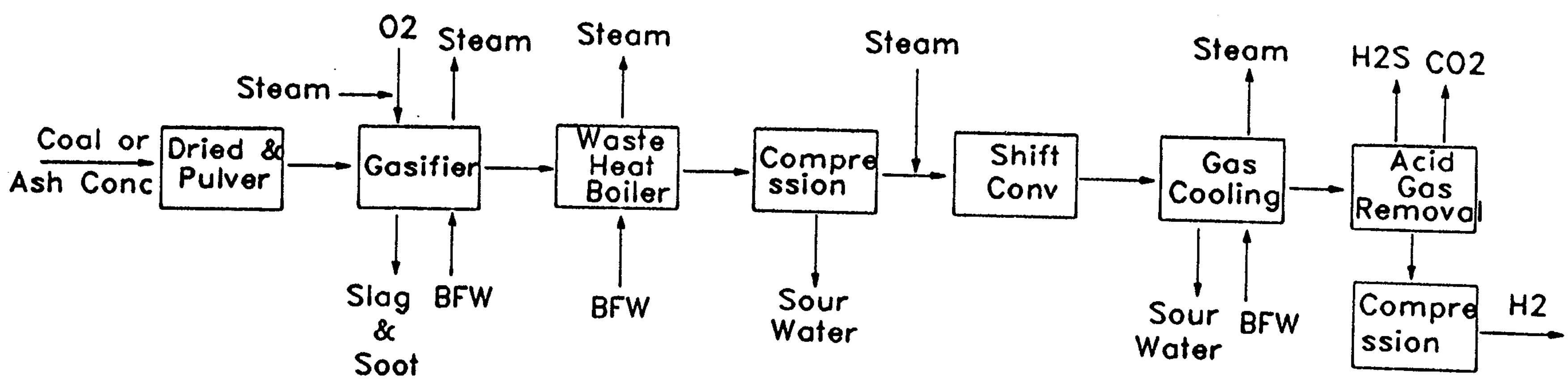
FIGURE 14.1

Block Flow Diagram – Hydrogen Production by Coal Gasification

Scheme 1 – Texaco Gasification



Scheme 2 – Shell Gasification



14.0.2 Design Considerations

The limiting size for the gasifier reactors was taken as 2,200 TPD of coal or ash concentrate. This is larger than any reactors which have been designed and built to date; however, Texaco has stated that designing this size of gasifier is achievable.

The shift reaction was designed to maximize hydrogen production. The shift reactor vessels were limited in size to 16 feet outside diameter for shop fabrication and transportation considerations.

The RECTISOL process was chosen because it has been commercially proven in this service. The process can separate carbon dioxide from hydrogen sulfide as a relatively pure stream for venting to the atmosphere. The hydrogen sulfide is sent to the Sulfur Plant. PSA is required after RECTISOL to achieve the required hydrogen purity.

14.1 Process Description and Process Flow Diagram

Plant 9 consists of seven trains (5 operating and 2 spare) of gasification, five trains of shift conversion and gas cooling, five trains of gas purification, and six trains of hydrogen compression. Process flow diagrams are presented on Flowsheets shown in Figures 14.2 through 14.4 as follows.

Fig. No.	Plant Section No.	Plant Name
14.2	Slurry Preparation and Gasification	9.1
14.3	Shift Reactor and Gas Cooling	9.2
14.4	Hydrogen Purification and Compression	9.3/9.4/9.5

Slurry Preparation and Gasification

Ash concentrate from Plant 8 (ROSE-SR) and coal from Plant 1 (Coal Cleaning and Handling) are fed with water and a special additive to a grinding mill. The slurry from the grinding mill discharges to a slurry run tank.

The slurry feed from slurry feed pumps and oxygen feed from the Plant 10 (Air Separation Plant) combine at the gasifier burner and enter the gasification chamber. This chamber operates at about 950 psig and 2500°F. About 92-95% of the carbon in the coal is converted to gaseous products, and the ash in the coal is slagged.

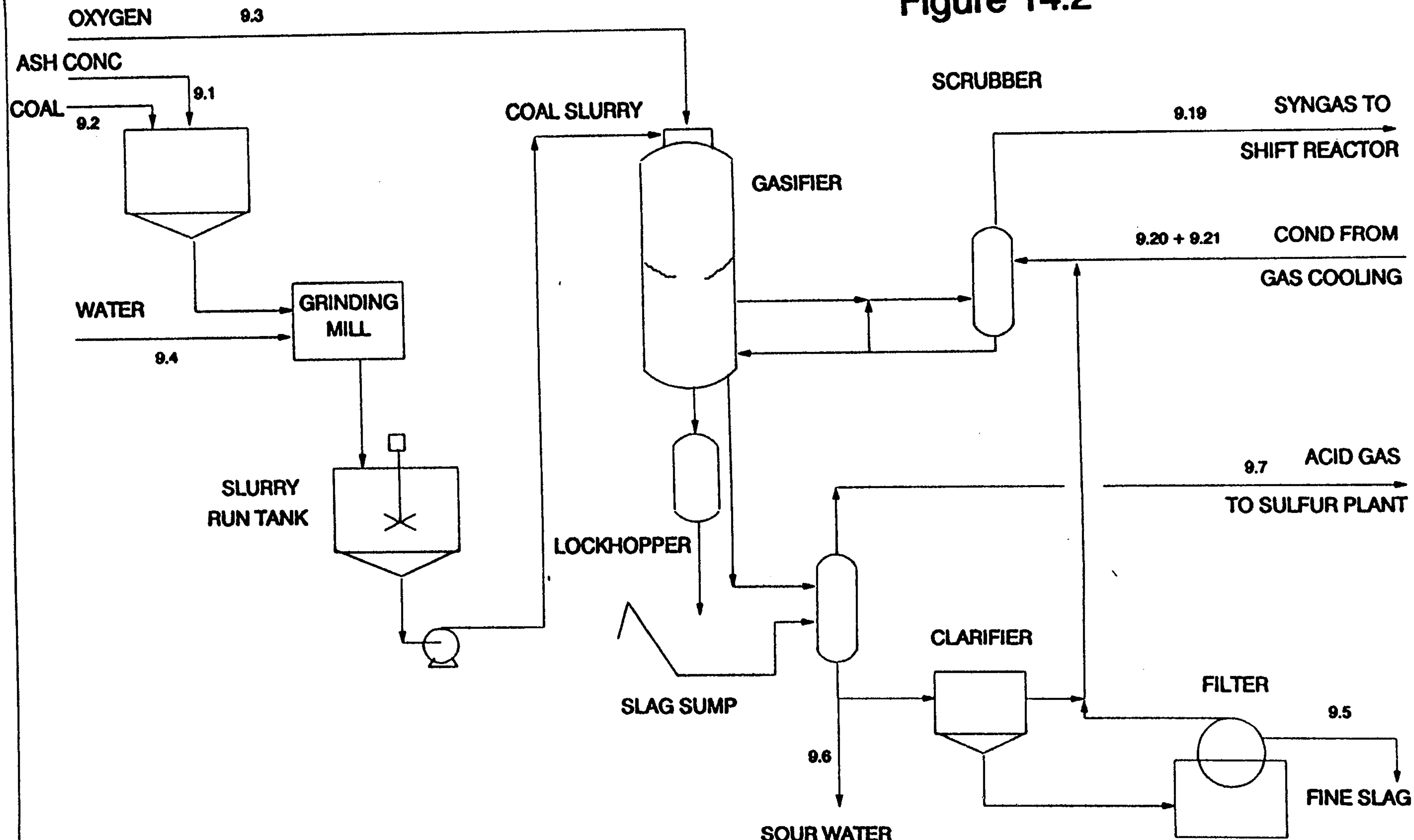
The gas with the molten slag is quenched in water at the bottom of the chamber. The slag is collected in a slag lock hopper connected to the base of the gasifier. Slag accumulations in the lock hopper are dumped into a sump. Slag dumped into the pit settles to the bottom. From the sump, the partly clarified soot slurry is pumped to a flash system to removed H_2S and other odorous material. The settled slag is dragged over the incline of the drag conveyor.

Synthesis gas leaving the gasifier is further cleaned in a water scrubber to remove the residual slag particles.

Water and condensate collected in the gas cooling system and in the soot water system are recycled to the scrubber.

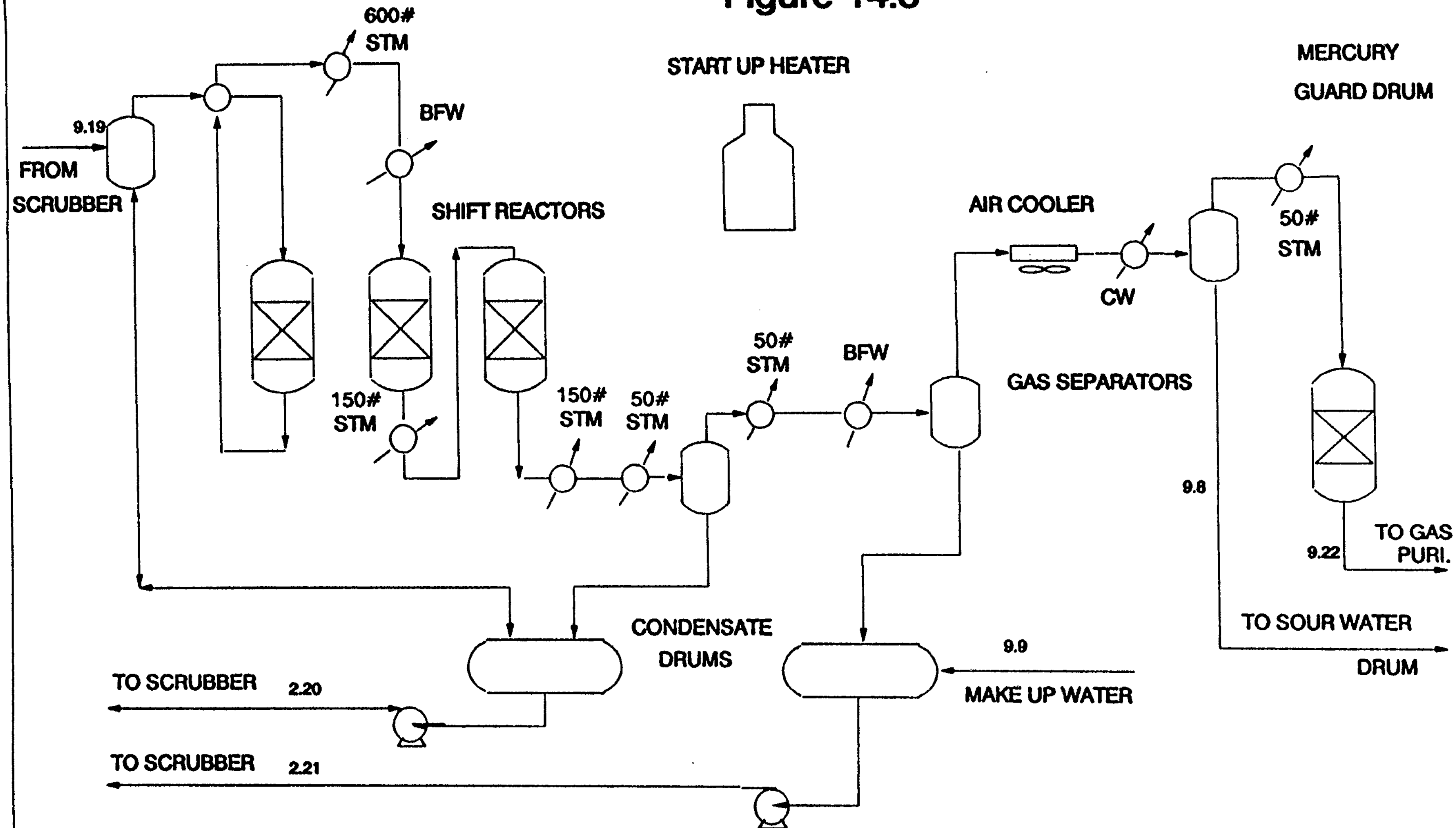
The size of the gasifier is limited by economic considerations and experience factors. Seven gasifiers of the largest design are included in the base-line design, including two spares to ensure continuous operation at the design capacity.

Figure 14.2



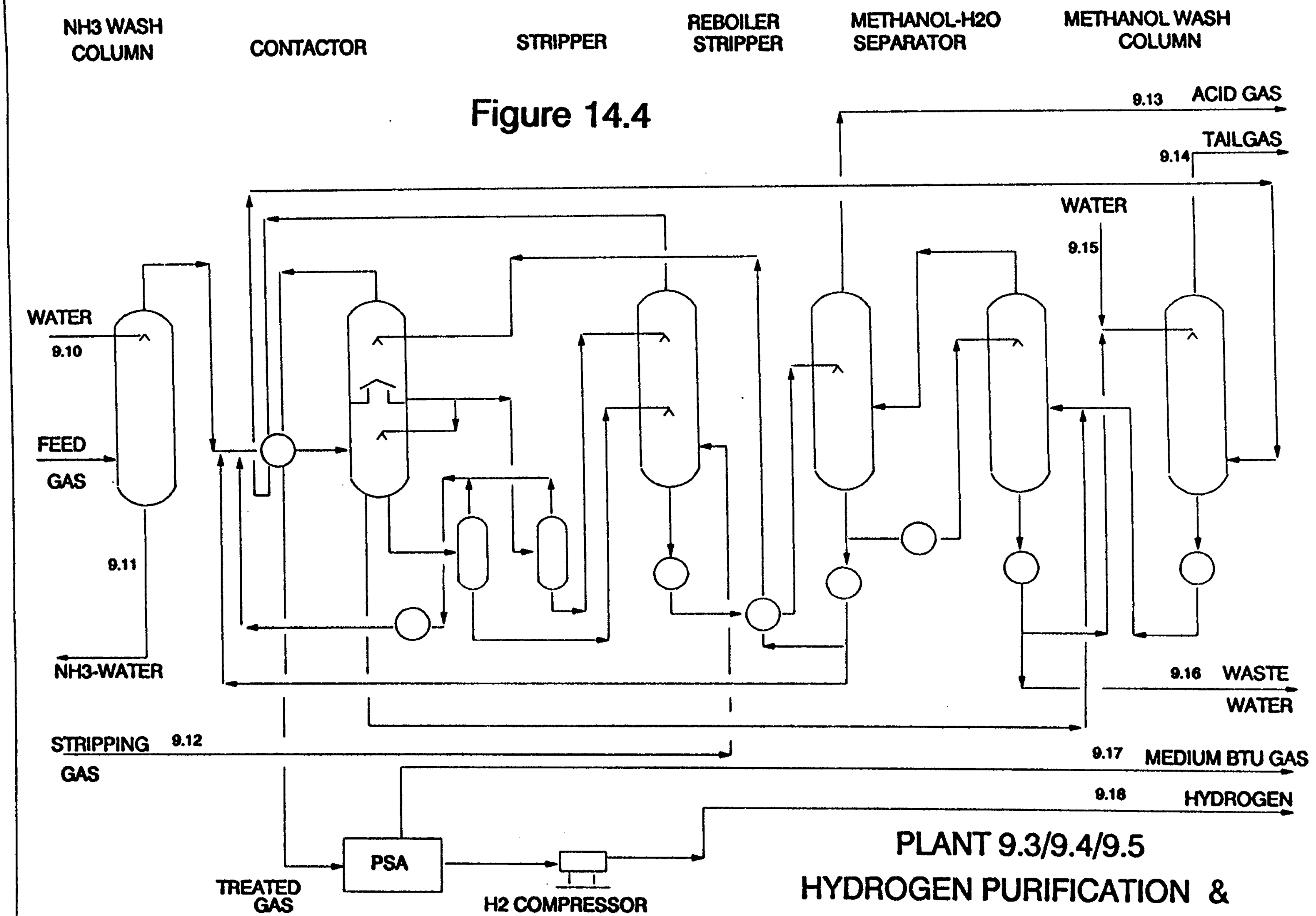
PLANT 9.1
SLURRY PREPARATION &
GASIFICATION

Figure 14.3



PLANT 9.2
SHIFT REACTOR &
GAS COOLING

Figure 14.4



PLANT 9.3/9.4/9.5
HYDROGEN PURIFICATION & COMPRESSION

Shift Reactor and Gas Cooling

The clean syngas from the gas scrubber goes to the shift converter where about 97-98% of the carbon monoxide is converted to hydrogen. The shift reaction is highly exothermic. A high temperature favors fast reaction rates but results in unfavorable equilibrium conditions. Initial stages will be designed to achieve high reaction rates at the highest allowable outlet temperature. The maximum allowable outlet temperature must be below the catalyst sintering point and within the limits for practical vessel design. The last reactor temperatures are kept low to obtain the low CO level in the product gas. The inlet temperature to each reactor stage must be kept above the water dew point to avoid catalyst being deactivated in the presence of liquid water.

A system with a multiple reactor design has been developed with the number of stages and trains, and operating conditions being optimized for minimum cost. To accommodate the required capacity, five trains each with three reaction stages were needed as limited by maximum reactor vessel diameter.

The catalyst chosen is either BASF's K8-11 or Haldor Topsoe's type SSK. Both catalysts operate in the sulfide form, thereby permitting handling of sulfur-bearing gas.

The heat of reaction is removed by a steam generator producing 600 psig steam from the first stage, a steam generator producing 150 psig steam from the second stage, and steam generators producing 150 psig and 50 psig steam from the third stage.

The shifted gas is further cooled against boiler feedwater, air, and cooling water and flows to the Hydrogen Purification section for acid gas removal. Process condensate is separated in the hot and warm receivers and pumped to the gas scrubbing and quench sections of the plant.

Hydrogen Purification and Compression

The Rectisol process removes H_2S and CO_2 by physical absorption of acid gas in methanol at relatively low temperature.

The feed gas is first prewashed by water in a water wash tower to remove NH_3 in the gas stream. The feed gas then enters at the bottom of a two-stage contactor where it is contacted with the methanol solution. In the first stage some of the CO_2 and practically all of the H_2S and COS are removed. In the second stage, which is above the first stage, the bulk of the CO_2 is removed.

The rich solvent which contains acid gas is regenerated by pressure reduction in the flash drums. The rich solvent is withdrawn from the flash drums and enter a stripper where the bulk of the CO_2 is stripped from the rich solvent by using nitrogen as the stripping gas.

The methanol leaving the stripper is completely regenerated in a reboiler regeneration column. The lean methanol withdrawn from the bottom of this column is fed to the top of the two-stage contactor. The acid gas, which contains all of the H_2S in the feed gas and some CO_2, is stripped from this column and sent to sulfur plant.

The fully regenerated methanol from the bottom of the reboiler regeneration column contains all the water in the feed gas. To purge this water, a slip stream is charged to the methanol-water separator column where dry methanol is driven overhead. The overhead vapor is fed to the H_2S stripper. The bottoms from the methanol-water separator is water. A part of this stream is cooled and used for washing traces of methanol vapor from the tail gas (CO_2) stream. The balance is waste water.

The treated gas from the two-stage contactor flows to a pressure swing adsorption unit where the hydrogen is upgraded to a purity of 99.9%. The hydrogen is then compressed and delivered to the coal liquefaction and hydrotreating plants.

14.2. Material Balance

The detailed material balance for the plant is shown in Table 14.3 where as the overall material balance is shown in Figure 14.5.

TABLE 14.3
PLANT 9
MATERIAL BALANCE

PLANT INPUT

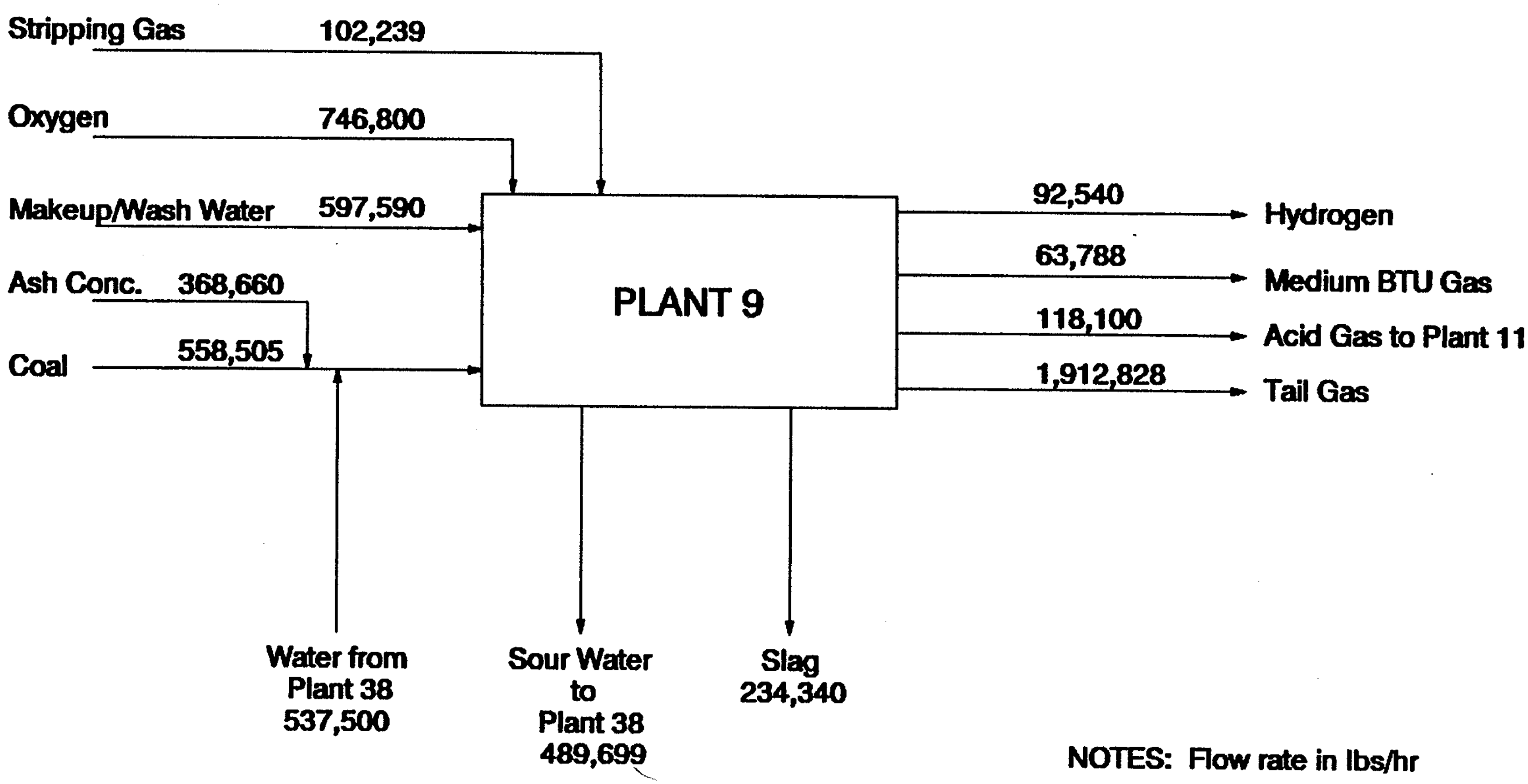
STREAM NO.	9.1	9.2	9.3		9.12	
	ASH CONC	CLEAN COAL	O2	MU STM & WATER	STRIP. GAS	TOTAL INLET
COMPONENTS	#/hr	#/hr	#/hr	#/hr	#/hr	#/hr
H2O		47898		1135090	0	1182988
N2			3269		102239	105508
O2			743531		0	743531
H2						0
CO						0
CO2						0
H2S						0
COS						0
NH3						0
C1						0
850+	114185					114185
UC COAL	90965	452032				542997
ASH	163510	58576				222086
SLAG						0
TOTAL	368660	558505	746800	1135090	102239	2911295

879268

PLANT OUTPUT

STREAM NO.	9.18	9.17		9.14		9.5	
	H2 PROD	MED BTU GAS	ACID GAS	TAIL GAS	SOUR WATER	SLAG	TOTAL OUTLET
COMPONENTS	#/hr	#/hr	#/hr	#/hr	#/hr	#/hr	#/hr
H2O	0	0	244	14790	482877	0	497911
N2	1280	14893	1268	98665	0	0	116106
O2	0	0	0	0	0	0	0
H2	91260	17549	168	390	5	0	109372
CO	0	20056	1845	1120	0	0	23021
CO2	0	10554	85329	1797790	5837	0	1899510
H2S	0	0	29136	7	285	0	29428
COS	0	0	105	0	0	0	105
NH3	0	0	4	0	695	0	699
C1	0	736	1	66	0	0	803
850+	0	0	0	0	0	0	0
UC COAL	0	0	0	0	0	0	0
ASH	0	0	0	0	0	222086	222086
SLAG	0	0	0	0	0	12254	12254
TOTAL	92540	63788	118100	1912828	489699	234340	2911295

BASELINE



NOTES: Flow rate in lbs/hr

Figure 14.5 OVERALL MATERIAL BALANCE

14.3 Major Equipment List

The major equipment for this plant is listed in Table 14.4.

Table 14.4

EQUIPMENT LIST

PLANT 9 -- COAL GASIFICATION & SHIFT CONVERSION

REACTORS AND VESSELS

Equipment No.	Equipment Description
9.1-C101	Gasifier
9.1-C102	Slag Lock Hopper
9.1-C103	Gas Scrubber
9.2-C104	No.1 Gas Separator
9.2-C105	First Stage Shift Reactor
9.2-C106	Second Stage Shift Reactor
9.2-C107	No.2 Gas Separator
9.2-C108	No.3 Gas Separator
9.2-C109	Third Stage Shift Reactor
9.2-C110	Final Gas Separator
9.2-C111	Mercury Guard Drum
9.1-C112	Flash Condensate Separator
9.2-C113	Hot Condensate Drum
9.2-C114	Warm Condensate Drum
9.1-C115	Soot Slurry Flash Drum
9.1-C116	Vacuum Pump Separator
9.1-C118	Vacuum Flash Drum
9.1-C119	Filtrate Receiver
9.1-C121	Blowdown Drum
9.1-C123	Burner CW Gas Separator
9.1-C124	Flash Water Drum
9.1-D101	Mill Discharge Tank w/Mixer
9.1-D102	Additive Storage Tank
9.1-D103	Water Supply Tank
9.1-D104	Slurry Run Tank w/Mixer
9.1-D105	Gray Water Tank
9.1-D106	Filter Feed Tank w/Mixer
9.1-D107	Burner Cooling Water Tank
9.1-D108	Vent Scrubber Tank

Table 14.4 Equipment List - continued

Towers

Equipment No.	Equipment Description
9.1-C117	Flash Gas Quencher
9.1-D120	Vent Air Separator
9.1-D122	Vent Air Quencher

Heat Exchangers

Equipment No.	Equipment Description
9.1-E101	Mill Water Heater
9.1-E102	Flash water Cooler
9.1-E103	Make up Water Cooler
9.1-E104	Burner CW Cooler
9.2-E105	Shift Reactor Preheater
9.2-E106	No.1 Steam Generator
9.2-E107	No.2 Steam Generator
9.2-E108	No.5 Steam Generator
9.2-E109	No.5 Gas Cooler
9.2-E110	Shift Gas Air Cooler
9.2-E111	Gas Final Cooler
9.2-E112	Hg Guard Preheater
9.2-E113	No.4 Steam Generat
9.1-E114	Soot Slurry Steam Generator
9.1-E115	Soot Slurry Cooler
9.1-E116	Flash Gas Cooler
9.1-E117	Gray Water Preheater
9.1-E118	Flash Quench Cooler
9.1-E119	Vent Air Quench Cooler
9.1-E120	Steam Blowdown Cooler
9.2-E121	Shift Gas Cooler
9.2-E122	No.3 Steam Generator

Table 14.4 Equipment List - continued

Pumps

Equipment No.	Equipment Description
9.1-G101	Mill Slurry Pump
9.1-G102	Slurry Feed Pump
9.1-G103	Gas Scrubber Circulation Pump
9.2-G104	Hot Condensate Pump
9.2-G105	Warm Condensate Pump
9.2-G106	Gray Water Pump
9.1-G107	Soot Slurry Pump
9.1-G108	Gray Water Charge Pump
9.1-G109	Slag Fine Pump
9.1-G110	Gray Water Feed
9.1-G111	Flash Water Supply Pump
9.1-G112	Flash Condensate Pump
9.1-G113	Filter Feed Pump
9.1-G114	Gray Water Purge Pump
9.1-G115	Water Supply Pump
9.1-G116	Additive Transfer Pump
9.1-G117	Slag Sump Pump
9.1-G118	Slag Quench Circulation Pump
9.1-G119	Air Quencher Circulation Pump
9.1-G120	Air Scrubber Circulation Pump
9.1-G121	Burner Cooling Pump

Vacuum Pumps

Equipment No.	Equipment Description
9.1-H101	Vacuum Pump
9.1-H102	Filter Vacuum Pump

Air Blowers

Equipment No.	Equipment Description
9.1-K101	Cake Blower
9.1-K102	Vent Air Blower

Table 14.4 Equipment List - continued

Fired heaters

Equipment No.	Equipment Description
9.1-F101	Shift Reactor Start Up Heater

Miscellaneous

Equipment No.	Equipment Description
9.1-T101	Slag Sump Conveyor
9.1-Y101	Rod Mill w/Trommel
9.1-Y102	Start Up Ejector
9.1-Y103	Gas Scrubber Ejector
9.1-Y104	Clarifier
9.1-Y105	Soot Filter
9.1-Y106	Coal Slurry Screen

14.4 Utility Summary

Table 14.5 below presents a summary of the utilities required for Plant 9.

Table 14.5

PLANT 9 UTILITY SUMMARY

Utility	Gasification & Shift Conversion	Hydrogen Purification	Hydrogen Compression
Steam, lb/hr		0	0
600 psig	-291,800	50,000	0
150 psig	-812,900	113,000	0
50 psig	-894,300		
Boiler Feedwater, gpm	4,000	0	0
Cooling Water, gpm	33,900	10,500	8,200
Raw Water, gpm	1,118	78	0
Fuel Gas, MMSCFD	0.62	0	0
Electricity, Kw	14,660	5,900	42,650
Nitrogen, MMSCFD	0	34	0

Note: Negative values represent utility production.

14.5 Water Summary

Table 14.6 below presents the water balances for the Hydrogen Production Plant (Coal Gasification - Texaco Process).

Table 14.6

WATER SUMMARY

	GPM Inlet	GPM Outlet
Plant 9.1/9.2		
Slurry Preparation	1,075	
Makeup Water	1,118	
Sour Water to Plt 38		912
Plant 9.3		
Wash Water	78	
Waste Water to Plt 38		54
TOTAL	2,271	966

15. Plant 10 (Air Separation)

15.0 Design Basis Criteria and Considerations and Process Flow Diagram

15.1 Process Description

This is a package plant, the process flow diagram of which is shown in Figure 15.1. The air separation package plants are cryogenic units producing oxygen at 99.5% (mol) purity and nitrogen at 99.9% (mol) purity. Oxygen will be delivered at the pressure required by the selected coal gasification process. Nitrogen will be used for purging, blanketing, and other utility purposes and will be delivered to the utility system at 150 psig. Liquid nitrogen will also be produced off the Air Separation Plant and stored for use during periods of high nitrogen demand.

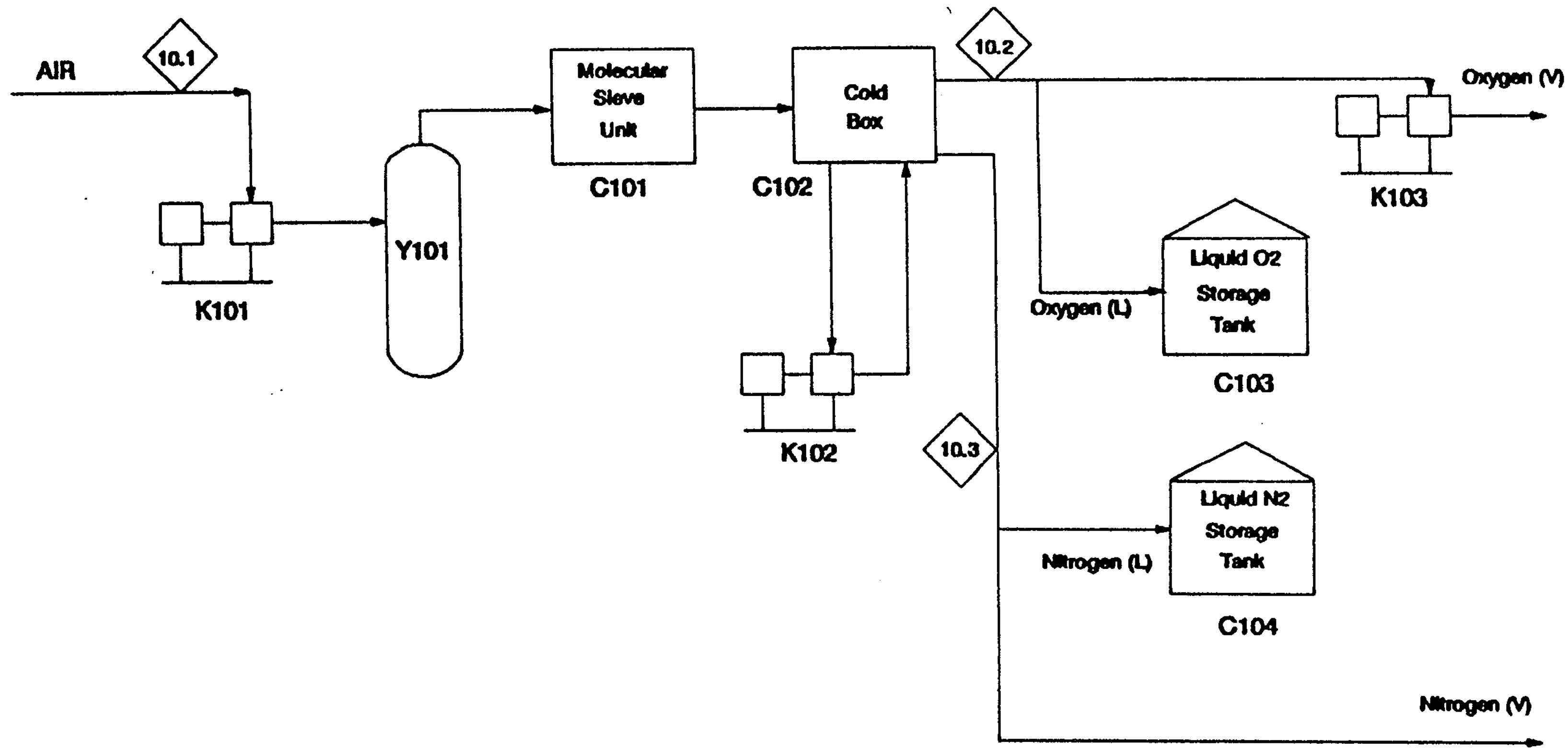
Plant 10 consists of five parallel trains each sized for 45 MMSCFD of oxygen. Each train can operate independently and uses a low pressure cycle type process. The process features an air compressor to compress the inlet air, direct contact cooling to remove the heat of compression from the air, molecular sieve units to purify the air, fractionation columns to obtain the high purity gaseous oxygen, an expander compressor to provide the refrigeration needed for the process, and an oxygen compressor to pressurize the oxygen product for use in the Coal Gasification Plant.

15.2 Material Balance

The overall material balance for this plant is shown in Table 15.1. This balance is for five trains.

PROCESS FLOW DIAGRAM

FIGURE 15.1



PLANT 10
AIR SEPARATION

TABLE 15.1

PLANT 10 (5 TRAINS)
MATERIAL BALANCE

	PLANT INPUT	PLANT OUTPUT	
STREAM NO.	10.1	10.2	10.3
COMPONENTS	AIR LBS/HR	OXYGEN LBS/HR	NITROGEN LBS/HR
OXYGEN	743531	743531	0
NITROGEN & ARGON	2463614	3269	2460345
TOTAL	3207145	746800	2460345

for 5 plants.

15.3 Major Equipment Summary

The major equipment summary is shown below in Table 15.2.

Table 15.2

Major Equipment Summary

Equipment Number	Equipment Description
10-C101	Molecular Sieve Unit
10-C102	Fractionation Tower (Cold Box)
10-C103	Liquid Oxygen Storage Tank
10-C104	Liquid Nitrogen Storage Tank
10-E101	Evaporative Cooler
10-G101	Circulation Pump
10-G102	Make-up Pump
10-K101	Inlet Air Compressor
10-K102	Expander
10-K103	Oxygen Compressor
10-Y101	Spray Cooler

15.4 Utility Summary

The only utility consumed by the Air Separation Plant is electric power. A summary is shown below:

Electricity, Kw	Per train	31,999
	Total	159,994

15.5 Water Summary

There is no water used in this process and no waste water produced.

16. Plant 11 (By-Product Sulfur Recovery)

16.0 Design Basis, Criteria and Considerations

Plant Capacity

Design capacity per train 180 LT/D

	No. of Trains	Total Capacity (LT/D)
• Operating	4	720
• Design	5	900

Yields

- Minimum sulfur recovery for Claus sulfur plant is 95%.
- Total sulfur recovery for Plant 11 is 99.9%.

Typical Sulfur Industry Specifications

• Sulfur	99.5%	minimum
• Carbon	0.2%	maximum
• Ash	50. ppm	maximum
• H2S	50. ppm	maximum
• SO_2	50. ppm	maximum
• Color	Bright yellow	
• Arsenic, Selenium, Tellurium	Commercially free	

Design Considerations

Three Claus trains were specified to be operational normally to achieve 100% on-line capacity in case one of the trains must be taken out of service on an emergency basis.

The SCOT process was selected for the tail gas treater for the Sulfur Plant because it is a proven process and is becoming an industry standard. It meets all present environmental regulations.

16.1 Process Description and Process Flow Diagram

The complete sulfur plant is comprised of three sections. These are, 1) Sulfur Plant, 2) SCOT Unit, and (3) Incinerator. The process description of each section is given below.

Sulfur Plant

The sulfur recovery plant, the description of which is shown in the Process Flow Diagram, Figure 16.1, has been designed in accordance with the principles of the classic "Claus" reaction. About one-third of the hydrogen sulfide (H_2S) in the feed is oxidized to form sulfur dioxide (SO_2) and water (H_2O). The sulfur dioxide then reacts with the remaining hydrogen sulfide to form elemental sulfur and water vapor. Ammonia (NH_3) is oxidized to nitrogen and water. In addition to the reactions noted above, a small part of the feed hydrogen sulfide dissociates to free hydrogen and elemental sulfur. Any hydrocarbons in the acid gas feed are oxidized in the thermal reactor to carbon monoxide (CO), carbon dioxide (CO_2) and water.

The acid gas and the SCOT recycle acid gas are combined and routed to the acid gas knock-out drum. The knock-out drum is designed to remove entrained sour water and condensed hydrocarbons from the acid gas streams.

The vapor stream, from the acid gas knock-out drum, is fed to the acid gas preheater, where the stream is heated to about 450°F. The hot gas is then fed to the acid gas injector, located on the front chamber of the thermal reactor.

Combustion air is supplied by the air blower to the acid gas injector to oxidize some of the H_2S to SO_2 and H_2O. The quantity of air is controlled to achieve a ratio of H_2S to SO_2 of 2:1. During this reaction, the thermal reactor first and second chamber temperatures are about 2400°F and 2200°F, respectively. The combustion air is heated to about 450°F in the air preheater to help maintain the 2200°F overall thermal reactor temperature.

The hot combustion gas from the thermal reactor second chamber enters a two-pass waste heat boiler. The boiler is designed to cool the hot gas to about 600°F and to generate 150 psig steam. The cooled gas from the boiler enters the thermal sulfur condenser. Most of the sulfur produced in the thermal reactor is condensed as the gas is cooled in the thermal condenser. The thermal condenser is a steam generator and the heat removed from the hot gases generates 50 psig steam. Liquid sulfur is separated from the gas in a separation chamber at the outlet end of the condenser. The condensed sulfur drains to a sulfur seal pot in a sulfur day tank through a steam jacketed drain leg. The seal pot maintains a liquid sulfur seal to prevent the process gas from escaping through the sulfur drain legs.

The cooled gas from the thermal condenser flows to the first reheater. The gas is heated to about 450°F by 600 psig steam.

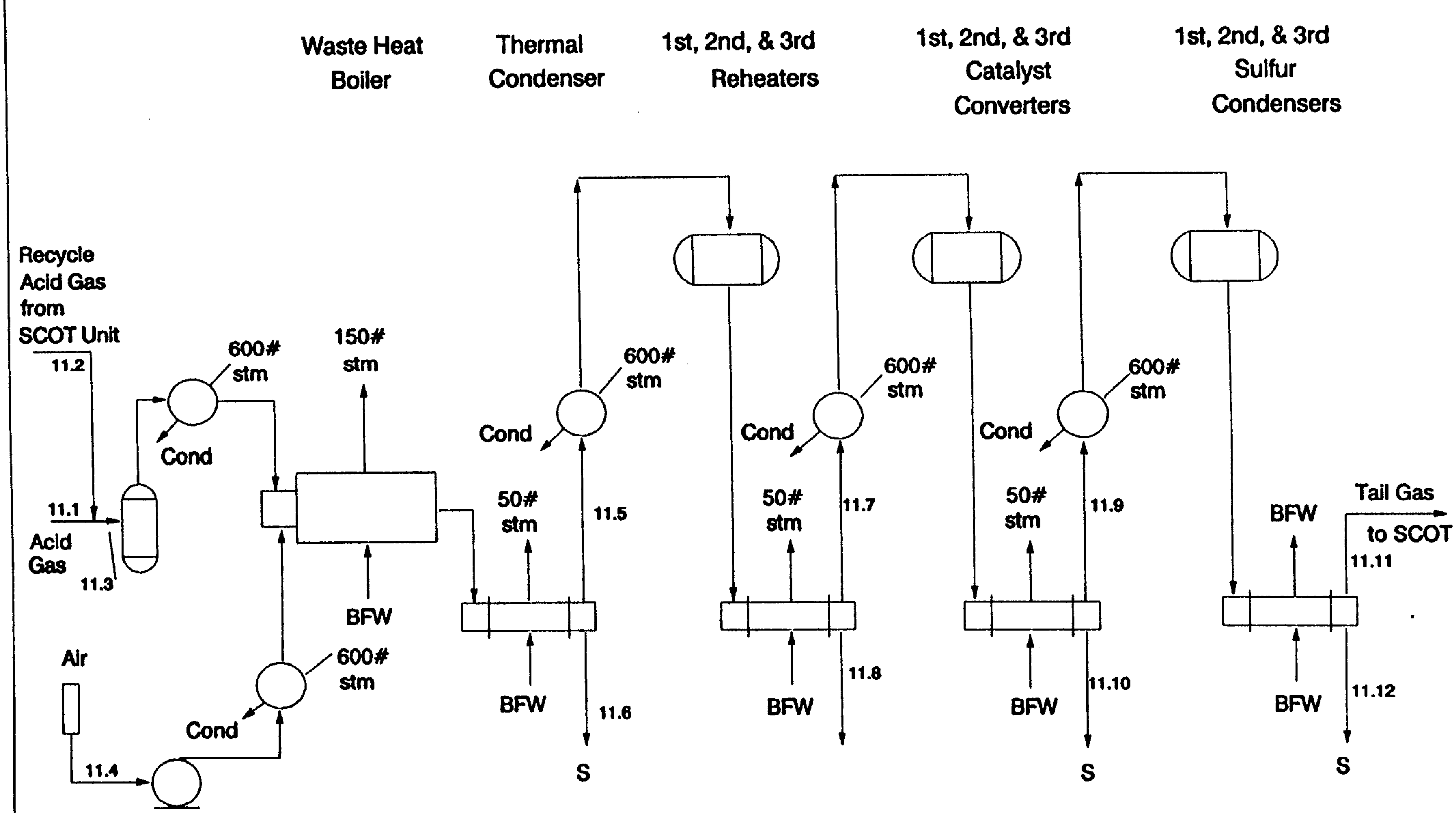


Figure 16.1 Process Flow Diagram

PLANT 11.1
SULFUR PLANT

sulfur.drw

The hot gas from the first reheater flows downward through the first catalyst converter where the reaction of SO_2 with H_2S occurs over a fixed bed of alumina catalyst. Since the reaction is exothermic, the temperature will rise across the bed. The temperature rise will be the greatest in this reactor because of the higher concentration of reactants present.

The reactor product gas, containing newly formed elemental sulfur, leaves the reactor and enters the first sulfur condenser. The bulk of the sulfur formed in the first catalyst converter is condensed and is drained to the sulfur day tank through a steam jacketed drain leg. Steam at 50 psig is generated on the shell side of the condenser.

The cooled gas from the first condenser flows to the second reheater where it is heated to about 420°F by 600 psig steam. Since the concentration of the reactants are lower in the feed to the second catalyst converter than in the feed to the first, less reaction takes place and the temperature rise across the second converter is lower.

Hot effluent gas from the second converter bed enters the second sulfur condenser. The bulk of the sulfur formed in the second converter is condensed and drained to the sulfur day tank through the third steam jacketed drain leg. Steam at 50 psig is generated on the shell side of the condenser. The cooled gas from the second sulfur condenser flows to the third reheater. This reheater has the same function as the other reheaters, and uses 600 psig saturated steam as a heating medium to heat the third catalyst converter feed to about 400°F.

The third catalyst converter is similar to the first and second catalyst converters. Since the concentration of H_2S and SO_2 is very low in the feed to the third reactor bed, a relatively small amount of sulfur is formed in this stage. The reaction that takes place in this stage results in a very small temperature rise across the catalyst bed. The sulfur formed in the catalyst converter is condensed in the third sulfur condenser and is drained to the sulfur day tank through a steam-jacketed drain leg. The cooled gas (tail gas) from the third sulfur condenser flows to the SCOT tail gas treating unit.

SCOT Unit

The Process Flow Diagram for this section is shown in Figure 16.2. The fundamental reaction in the Hydrogenation Section of the SCOT Unit are hydrogenation of the SO_2 and elemental sulfur. Most of the COS and CS_2 present in the reactor are hydrolyzed by the water vapor in the tail gas.

The sulfur plant tail gas is heated to over 500°F in the SCOT feed heater. The temperature of this stream is controlled to the desired SCOT reactor inlet temperature by means of a temperature controller, which regulates the amount of natural gas or syngas that is fed to the SCOT burner as fuel.

Syngas is added to the feed heater outlet to provide the necessary hydrogen for the hydrogenation of the SO_2 and elemental sulfur to H_2S.

The heated gases then pass downward through the reactor catalyst bed where the SO_2, elemental sulfur and other sulfur containing compounds (such as COS) are converted to H_2S, CO_2, CH_4, and water vapor. To assure the complete reduction of the sulfur compounds to H_2S, an excess above the stoichiometric requirement of hydrogen is fed to the reactor. The reactions are exothermic and cause a temperature rise across the reactor.

The hot gases leaving the SCOT reactor are cooled to about 320°F in the SCOT waste heat exchanger. The waste heat exchanger is a steam generator and the heat removed from the hot gases generates 50 psig steam.

The cooled gases leaving the waste heat exchanger are further cooled to about 100°F by contact with water in the quench tower. The gases enter the quench tower below the tower packing and flow upward, countercurrent to the cool quench water. The quench water is fed to the tower above the packing at a temperature of about 100°F. The temperature of the quench water is raised to about 150°F as it cools the gases in the column. The quench water from the bottom of the column is then pumped by the quench water pumps, through the quench water cooler, where it is cooled before being returned to the top of the tower. The quench water bleed stream is the excess water contained in the quench tower feed gas that is condensed in the quench tower.

The quench tower overhead gas stream enters the SCOT absorber below the bottom tray and flows upward through the absorber countercurrent to the MDEA solution flowing down through the absorber.

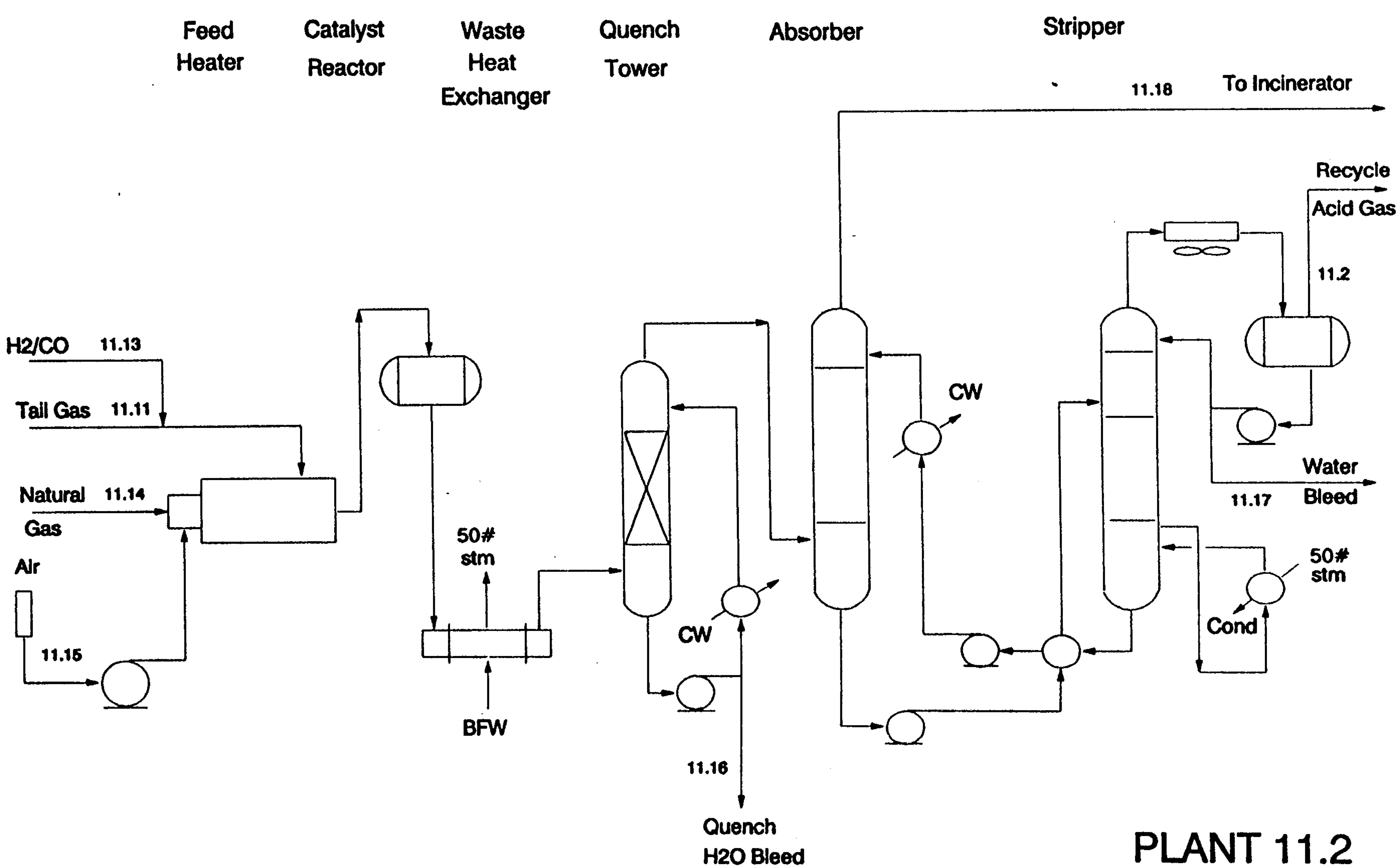
Lean MDEA solution at about 100°F is fed to the top of the absorber. The solvent flows downward through the absorber, contacting the gas stream flowing upward and absorbing nearly all of the H_2S and part of the CO_2 from the gas stream. The treated vent gas then leaves the top of the absorber at about 100°F and is directed to the incinerator.

A vent gas analyzer is provided to advise the operator of the SCOT Unit clean up efficiency, H_2S and CO_2, and the amount of excess hydrogen in the vent gas.

Rich solvent is withdrawn from the absorber bottom and is pumped to the SCOT stripper through the lean/rich exchanger. The rich solvent enters the column above tray 5 (from the top). The stripper is heated by a reboiler using 50 psig steam.

The stripper overhead vapor is cooled to about 120°F in the reflux condenser (an air cooler), and collected in the reflux drum. The gas from the top of the reflux drum contains H_2S, CO_2, and H_2O and is routed to the sulfur plant as acid gas feed.

The lean solvent leaves the stripper bottom and flows through the lean/rich exchanger. Lean solvent from the exchanger is pumped to the SCOT absorber by the lean solvent pumps. The lean solvent stream from the pump discharge is cooled to about 100°F in the lean solvent cooler and is fed to the top of the absorber.



PLANT 11.2

SCOT UNIT

Figure 16.2 Process Flow Diagram

Incinerator

The Process Flow Diagram for this section of the plant is shown in Figure 16.3. The incinerator uses a natural draft burner to provide the heat and air necessary for the combustion of the effluent gases. The design combustion temperature is maintained by burning natural gas or syngas as fuel. The SCOT vent gas is routed below the incinerator burner to achieve maximum turbulence and contact with the oxygen (air) at high temperature.

During normal operations, the sulfur plant tail gas flows to the SCOT unit and only the SCOT vent gas is combusted in the incinerator. The design conditions for proper combustion of the SCOT vent gas is 25% excess air at a temperature of about 1200°F. The combustion products are discharged to the atmosphere through a 280 foot stack. A stack gas analyzer is provided to monitor the SO_2 in the stack.

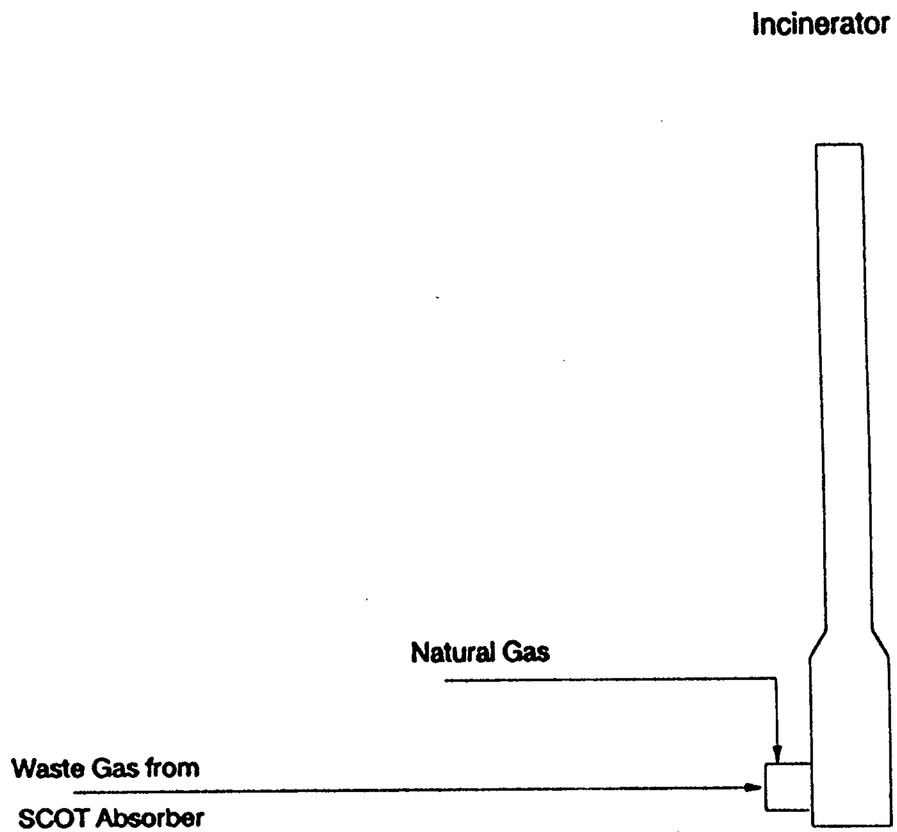


Figure 16.3 Process Flow Diagram

INCINERATOR
PLANT 11.3

16.2 Material Balance

Material balance for the complete plant is included in Table 16.1.

TABLE 16.1
PLANT 11
MATERIAL BALANCE

PLANT INPUT

COMPONENTS	PLANT 6 ACID GAS LBS/HR	PLANT 9 ACID GAS LBS/HR	PLANT 38 ACID GAS LBS/HR	AIR LBS/HR	TOTAL INPUT LBS/HR
H2O	1749	244	5195	2733	9921
H2	0	168	5		173
H2S	18812	29136	17695		65643
COS	0	105	0		105
NH3	0	4	13		17
CO	0	1845	0		1845
CO2	1193	85329	6400		92922
N2	0	1268	0	251530	252798
O2	0	0	0	75770	75770
C1	0	1	8		9
C5-350	3	0	0		3
PHENOLS	0	0	321		321
HC	0	0	12033		12033
SULFUR PROD	0	0	0		0
TOTAL	21757	118100	41670	330033	511560

PLANT OUTPUT

COMPONENTS	SULFUR LBS/HR	GAS TO INCIN. LBS/HR	SOUR WATER LBS/HR	TOTAL OUTPUT LBS/HR
H2O		781	61160	61941
H2		555		555
H2S		63		63
COS		105		105
NH3		17		17
CO		1845		1845
CO2		132514		132514
N2		252798		252798
O2		0		0
C1		0		0
C5-350		0		0
PHENOLS		0		0
HC		0		0
SULFUR PROD	61722	0		61722
TOTAL	61722	388678	61160	511560

16.3 Major Equipment List

The major equipment list for this plant is included in Table 16.2.

Table 16.2 Major Equipment List

PLANT 11-- SULFUR AND SCOT PLANT

Reactors and Vessels

Equipment No.	Equipment Description
11.1-C101	Acid Gas KO Drum
11.1-C102	Combustion Chamber
11.1-C103	First Catalyst Converter
11.1-C104	Second Catalyst Converter
11.1-C105	Third Catalyst Converter
11.2-C106	SCOT Feed Heater
11.2-C107	SCOT Catalyst Reactor
11.2-C111	Reflux Accumulator

Towers

Equipment No.	Equipment Description
11.2-C108	Quench Tower
11.2-C109	SCOT Absorber
11.2-C110	SCOT Stripper

Heat Exchangers

Equipment No.	Equipment Description
11.1-E101	Acid Gas Preheater
11.1-E102	Air Preheater
11.1-E103	First Reheater
11.1-E104	Second Reheater
11.1-E105	Third Reheater
11.1-E106	Waste Heat Boiler
11.1-E107	Thermal Condenser
11.1-E108	First Sulfur Condenser
11.1-E109	Second Sulfur Condenser
11.1-E110	Third Sulfur Condenser
11.2-E111	Waste Heat Exchanger
11.2-E112	Quench Water Cooler
11.2-E113	Lean Solvent Cooler
11.2-E114	Lean/Rich Solvent Exchanger
11.2-E115	Overhead Condenser
11.2-E116	Reboiler

Table 16.2 Major Equipment List - continued

Pumps

Equipment No.	Equipment Description
11.2-G101A/B	Quench Water Pump
11.2-G102A/B	Rich Solvent Pump
11.2-G103A/B	Lean Solvent Pump
11.2-G104A/B	Reflux Pump
11.4-G105A/B	Solvent Charge Pump
11.4-G106	Solvent Sump Pump
11.4-G107A/B	Sulfur Transfer Pump (Sump Type, Steam Jacketed)
11.4-G108A/B/C	Sulfur Loading Pump (Steam Jacketed)

Air Blowers

Equipment No.	Equipment Description
11.1-K101	Sulfur Plant Air Blower
11.2-K102	SCOT Air Blower

Miscellaneous

Equipment No.	Equipment Description
11.1-F101	Acid Gas Burner
11.2-F102	Tail Gas Burner
11.3-F103	Incinerator Burner
11.3-Y101	Incinerator
11.2-Y102	Quench Filter
11.2-Y103	Lean Solvent Filter
11.2-Y104	Solvent Sump Filter
11.4-D101	Solvent Storage Tank
11.4-D102	Sulfur Day Tank
11.4-D103	Sulfur Storage Tank
11.4-D104	Solvent Sump

16.4 Utility Summary

The individual plant utility summaries for the Sulfur Plant and SCOT unit are tabulated below in Table 16.5.

Table 16.3 Utility Summary

	CW	Steam Rate, lbs/hr				Fuel Gas	Electr	H2
Units	gpm	50#	150#	600#	900#	MMSCFD	Kw	MMSCFD
Sulfur	0	-64859	-186541	59956	0	0	1991	1.6
SCOT	8126	119327	0	0	0	1.7	1071	0
Total	8126	54468	-186541	59956	0	1.7	3062	1.6

Note: Negative values represent utility production

16.5 Water Summary

The Sulfur Plant uses three gallons per minute (gpm) of raw water and produces 72 gpm of sour water which is sent to Ammonia Recovery, Plant 38.

17. Plant 19 (Relief and Blowdown Facilities)

17.0 Design Basis, Criteria and Considerations

Plant 19 is for the collection and flaring of relief and blowdown discharges from all applicable plants. It includes all the plant flare subheaders, the main flare headers, and the flare stacks. Relief and blowdown subheaders are provided for Coal Liquefaction (Plant 2), the Gas Plant (Plant 3), the Naphtha and Gas Oil Hydrotreaters (Plants 4 and 5, respectively), Hydrogen Purification (Plant 6), Solvent Recovery (Plant 8), Hydrogen Production from Coal Gasification (Plant 9), Sulfur Recovery (Plant 11), and Wastewater Treatment (Plant 34).

Two separate main headers are provided for high and low pressure relief valves. Valves with set pressures below 300 psig relieve to the low pressure header while valves with set pressures equal to or above 300 psig relieve to the high pressure header. Both high and low pressure headers are connected to a common knockout drum which is connected to the two main flare stacks (one operating and one spare).

One dedicated flare header with a separate flare stack is provided for flaring H_2S emissions from Plant 34.

A preliminary analysis of the various causes of over pressure has been made for each plant to determine which contingency governs flare header size. The various causes of over pressure considered were:

- Electrical Power Failure
- Cooling Water Failure
- Total Instrument Air Failure
- Control Valve Failure
- Inadvertent Valve Closing/Opening
- Plant Fire
- Reflux Failure
- Other Failure

The various header sizing contingencies are given in Tables 17.1, 17.2 and 17.3 on the following pages. The contingency which required the greatest header size was used for sizing. These tables summarize the contingencies selected as the basis for subheader and main header sizing.

The high pressure header accepts discharge from valves with set pressures of 300 psig or above. It has a maximum total superimposed plus built up back pressure of 75 psig which may be applied to any relief valve.

The low pressure header accepts discharge from valves with set pressures below 300 psig. For the low pressure header, the maximum total back pressure is 15 psig.

Balanced bellows-type relief valves will be used in those cases where back pressure exceeds 10% of the relief valve setting. Conventional relief valves are to be used where the back pressure is less than 10% of the valve set pressure.

This system analysis is preliminary and serves as a guide only, not a final design.

Vapor relief loads to the flare system are minimized by relieving to atmosphere where practicable without jeopardizing plant safety. In general, API RP -520, "Recommended Practice for the Design and Installation of Pressure-Relieving Systems in Refineries," and API RP -521, "Guide for Pressure Relief and Depressuring Systems," are followed.

TABLE 17.1

HIGH PRESSURE FLARE HEADER SIZING CONTINGENCIES

Plant No.	Plant Name	Contingency Used as Basis for Sizing Subheader
4	Naphtha Hydrotreater	Power Failure
5	Gas Oil Hydrotreater	Power Failure
8	Solvent Recovery	Blocked Discharge
9	H_2 Production by Coal Gasification	Blocked Discharge

TABLE 17.2

LOW PRESSURE FLARE HEADER SIZING CONTINGENCIES

Plant No.	Plant Name	Contingency Used as Basis for Sizing Subheader
2	Coal Liquefaction	Power Failure
3	Gas Plant	Cooling Water Failure
4	Naphtha Hydrotreater	Cooling Water Failure
5	Gas Oil Hydrotreater	Power Failure
6	H_2 Purification	Blocked Discharge
11	Sulfur Recovery	Blocked Discharge

TABLE 17.3

MAIN HEADER SIZING CONTINGENCIES

Flare Header	Contingency Used as Basis for Sizing Main Header
Low Pressure	Power Failure (Plants 2 and 5)
High Pressure	Blocked Discharge (Plant 9)
H_2S (Plant 34)	Cooling Water Failure

17.1 Plant Description and Block Flow Diagram

Refer to Figure 17.1 for the block flow diagram for this plant. Each plant is equipped with the necessary flare headers to collect over pressure relief valve discharges, operational (startup and shutdown) discharges, and emergency vents. These unit headers tie into the appropriate main flare headers. Subheader and main header sizes are provided in the Major Line List.

The main headers run down the central pipe rack corridor and combine in the primary flare knockout drum (19-C101). Emission of H_2S from Plant 34 is an exception since it goes directly to the H_2S flare header, 19-F102, bypassing the knockout drum.

Inside the knockout drum, the relieving fluid is separated into a vapor stream, which is directed to the main flare stack and burned, and a liquid stream which is collected and pumped back to the crude unit surge tank for redistillation.

Two pumps, one motor-driven (19-G101) and the other turbine-driven (19-G102), are provided to pump back the recovered liquid. Level switches automatically start/stop either pump so pumping is maintained as required.

From the knockout drum, vapor is transported to the primary flare stacks (19-F101A and B) which include one operating stack and one spare.

Flare Stack Details

Each flare stack has refractory lined tips, a flame front generator that provides ignition to the pilot gas burner located atop each flare stack, and molecular seals to prevent air from entering the system. In addition, each primary flare feeder will be purged with fuel gas.

The primary flare features smokeless operation, achieved by an automatically controlled injection of steam. The steam provides smoke suppression by promoting more intimate mixing of flared material and the oxygen in the air to provide more complete combustion.

Knockout Drum Details

The primary flare knock-out drum is equipped with steam heating lances. These lances are used to:

- prevent material from freezing in winter conditions.
- provide means to vaporize light ends that may not be desirable to recover.

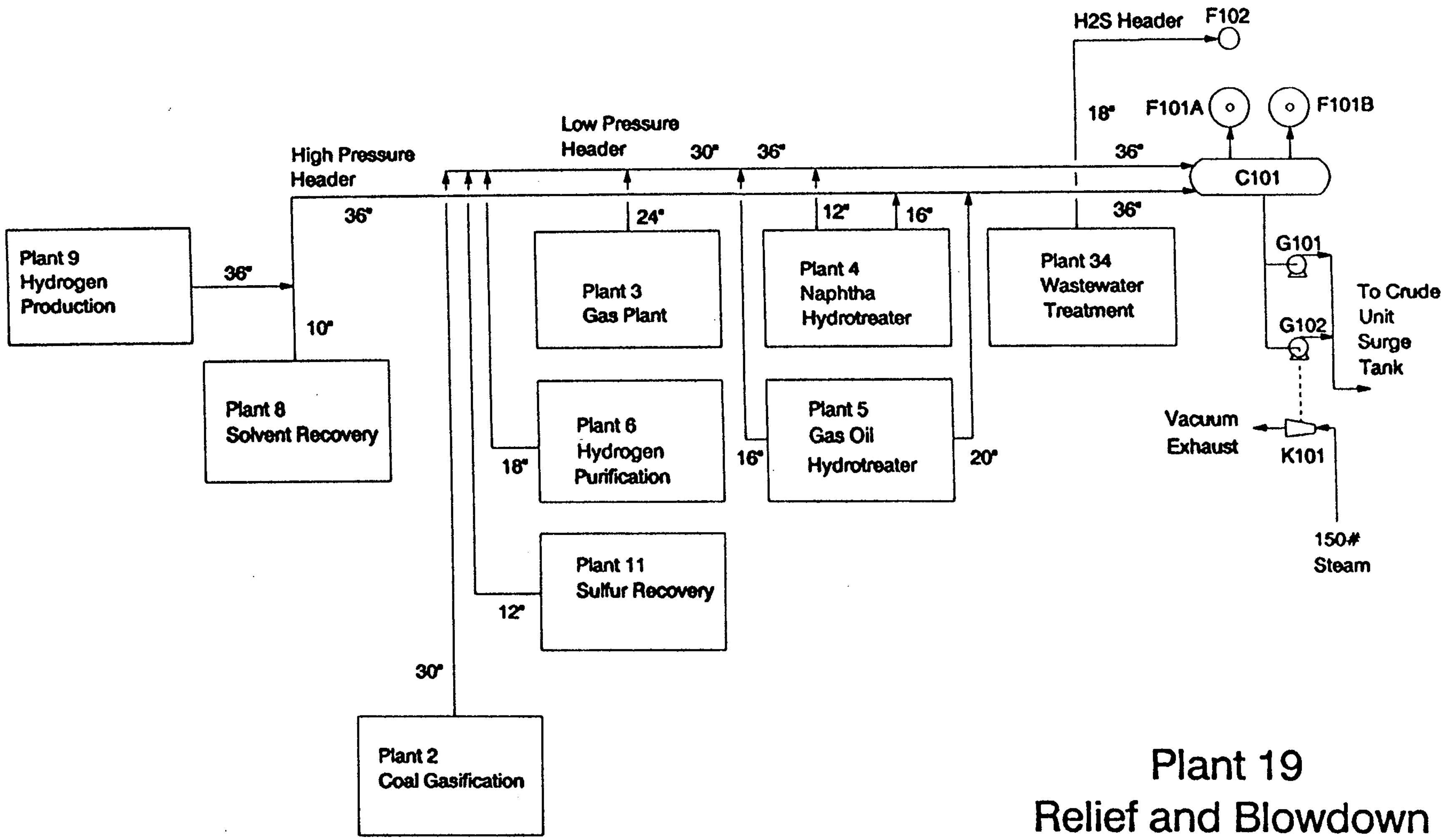


Figure 17.1 Block Flow Diagram

17.2 Major Lines List

Line Contents	From	To	Size (in.)
Atmospheric Tower and WHPS Overheads	Plant 2	Low Pressure Header	30
Stripper and Depropanizer Overheads	Plant 3	Low Pressure Header	24
OHPS Overhead	Plant 4	High Pressure Header	16
Fractionator Overhead	Plant 4	Low Pressure Header	12
HHPS Overhead	Plant 5	High Pressure Header	20
HLPS and Fractionator Overheads	Plant 5	Low Pressure Header	16
Compressor K102 Discharge	Plant 6	Low Pressure Header	18
Solvent Stream	Plant 8	High Pressure Header	10
Synthesis Gas	Plant 9	High Pressure Header	36
H_2S Acid Gas	Plant 34	H_2S Header	18
Low Pressure Discharges (< 300 psig)	Low Pressure Header	Primary Flare	36
High Pressure Discharges (>= 300 psig)	High Pressure Header	Primary Flare	36

18. Plant 20 (Tankage)

18.0 Design Basis, Criteria and Considerations

Plant 20 provides storage and delivery equipment for products and intermediates.

Products include liquid hydrocarbon fractions from the Gas Oil Hydrotreaters (Plants 4 and 5), propane and mixed butanes from the Gas Plant (Plant 3), sulfur from Sulfur Recovery (Plant 11), ammonia from the Ammonia Plant (Plant 38), and phenols from the Phenol Plant (Plant 39).

Intermediates include hydrocarbon fractions from Plant 2 and sour water feed for Ammonia Removal (Plant 38). The hydrocarbon fractions from Plant 2 are feed for the Naphtha and Gas Oil Hydrotreaters (Plants 4 and 5) and the Purge and Flush Oil System (Plant 36).

Plant 20 includes tanks, pumps, heat exchangers, and miscellaneous equipment required for product and intermediate storage and delivery.

Product storage

Thirty days storage is provided for the 4 liquid hydrocarbon fractions, 15 days for propane, mixed butanes, sulfur, ammonia, and phenol. Storage times were chosen to allow for variations in production and shipping rates. The 15 days of propane and ammonia storage consist of 10 days of refrigerated storage at atmospheric pressure and five days of pressurized storage at ambient temperature. Both types of storage are provided since refrigerated storage for propane and ammonia is more cost effective, and pressurized storage is required for shipment in pressurized containers at ambient temperature.

Intermediate storage

Six hours intermediate storage is provided for the liquid hydrocarbon feeds from Coal Liquefaction (Plant 2) to the Naphtha and Gas Oil Hydrotreaters (Plants 4 and 5). The 6 hour storage capacity is required to provide feedstock during plant startup (prior to Plant 2 delivery of the necessary feed). Additionally, the storage mitigates the effect on downstream plant operations due to brief interruptions in the upstream plant. Interruptions could be as a result of scheduled or unscheduled maintenance or due to operating problems.

The sour water feed (for Ammonia Removal, Plant 38) and the light and heavy flush oils (for the Purge and Flush Oil System, Plant 36) have storage times of 5, 10, and 10 days, respectively, for the same reasons discussed above (for operation during startup and to mitigate the effect of upstream plant interruptions).

Tank sizing

The following factors were considered for tank sizing:

- All tanks are sized to API 650 except for pressurized spherical tanks.
- Tanks are sized for 95% maximum working capacity.
- At least two tanks are used for each finished product to avoid the problem of running to and shipping from the same tank.
- Due to soil loading considerations, tank height is limited to 48 feet for cylindrical tanks.

18.1 Plant Description and Block Flow Diagram

An overview of this plant is shown in block flow diagrams, Figures 18.1 and 18.2.

Product storage

Storage is provided for the following products as shown in the Block Flow Diagram, Figure 18.1:

- One naphtha product (C5-350°F) from the Naphtha Hydrotreater (Plant 4)
- One light distillate product (350-450°F) from the Gas Oil Hydrotreater (Plant 5)
- One heavy distillate product (450-650°F) from the Gas Oil Hydrotreater (Plant 5)
- One gas oil product (650-850°F) from the Gas Oil Hydrotreater (Plant 5)
- Propane and mixed butanes from the Gas Plant (Plant 3)
- Molten sulfur from the Sulfur Plant (Plant 11)
- Ammonia from the Ammonia Plant (Plant 38)
- Phenols from the Phenol Plant (Plant 39)

Naphtha Product - The naphtha product (C5-350°F) is delivered from the Naphtha Hydrotreater (Plant 4) fractionator bottoms at 100°F and stored in three 217,500 barrel tanks (20-D101A, B, and C) to provide 30 days storage. Each tank is equipped with a floating roof inside a stiffened dome roof. The naphtha product is pumped via pipeline in 8 hour batches to down stream customers. The product loading pumps (20-G101A and B) are designed for 4375 gpm at 1500 psi which allows delivery of 50,000 barrels to the pipeline over an eight hour period.

Light Distillate Product - The light distillate product (350-450°F) is delivered from the Gas Oil Hydrotreater (Plant 5) at 110°F and stored in three 197,000 barrel cone roof tanks (20-D102A, B, and C) to provide 30 days storage. The light distillate product is pumped via pipeline in 8 hour batches to down stream customers. The product loading pumps (20-G102A and B) are designed for 4375 gpm at 1500 psi which allows delivery of 50,000 barrels to the pipeline over an eight hour period.

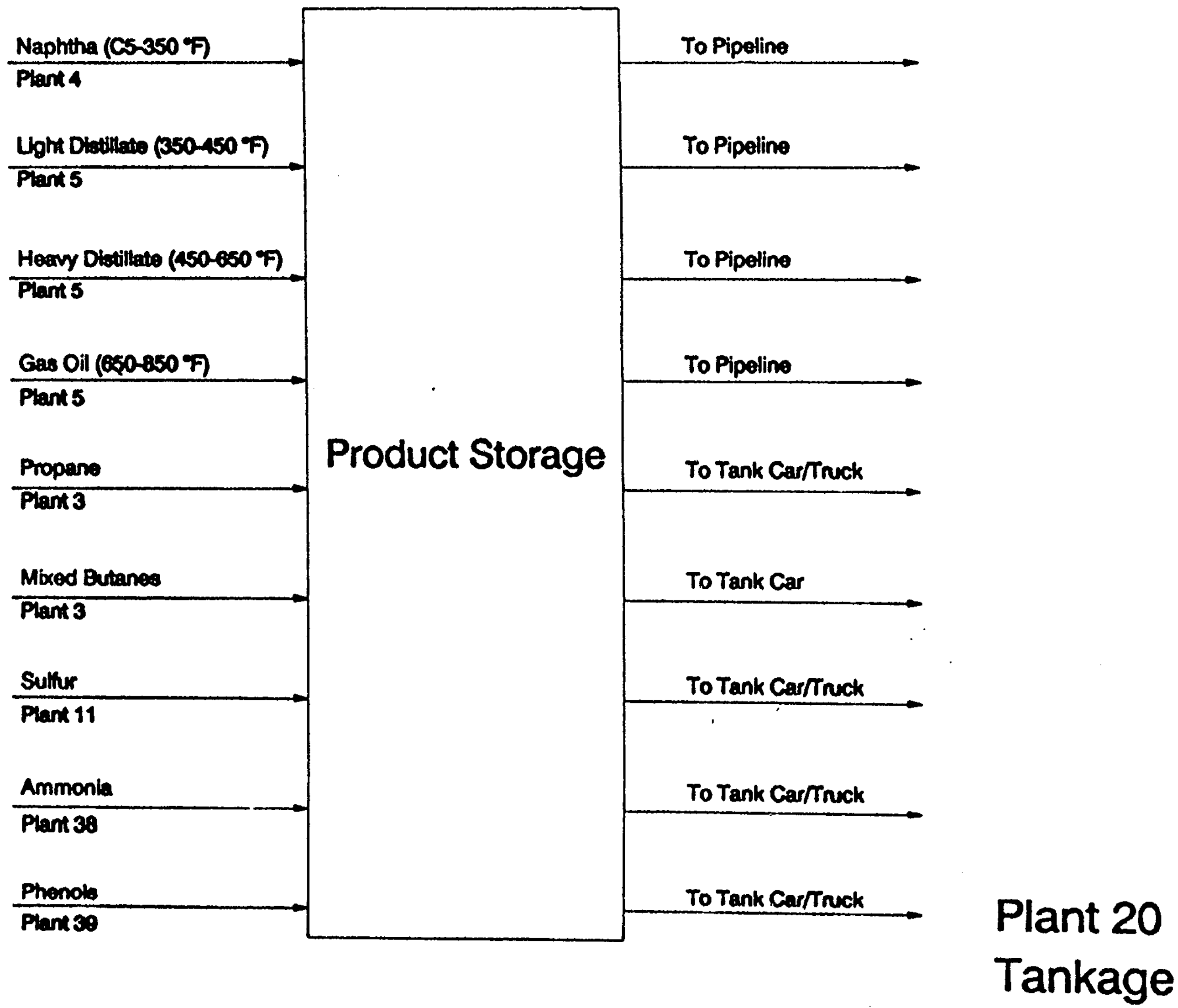
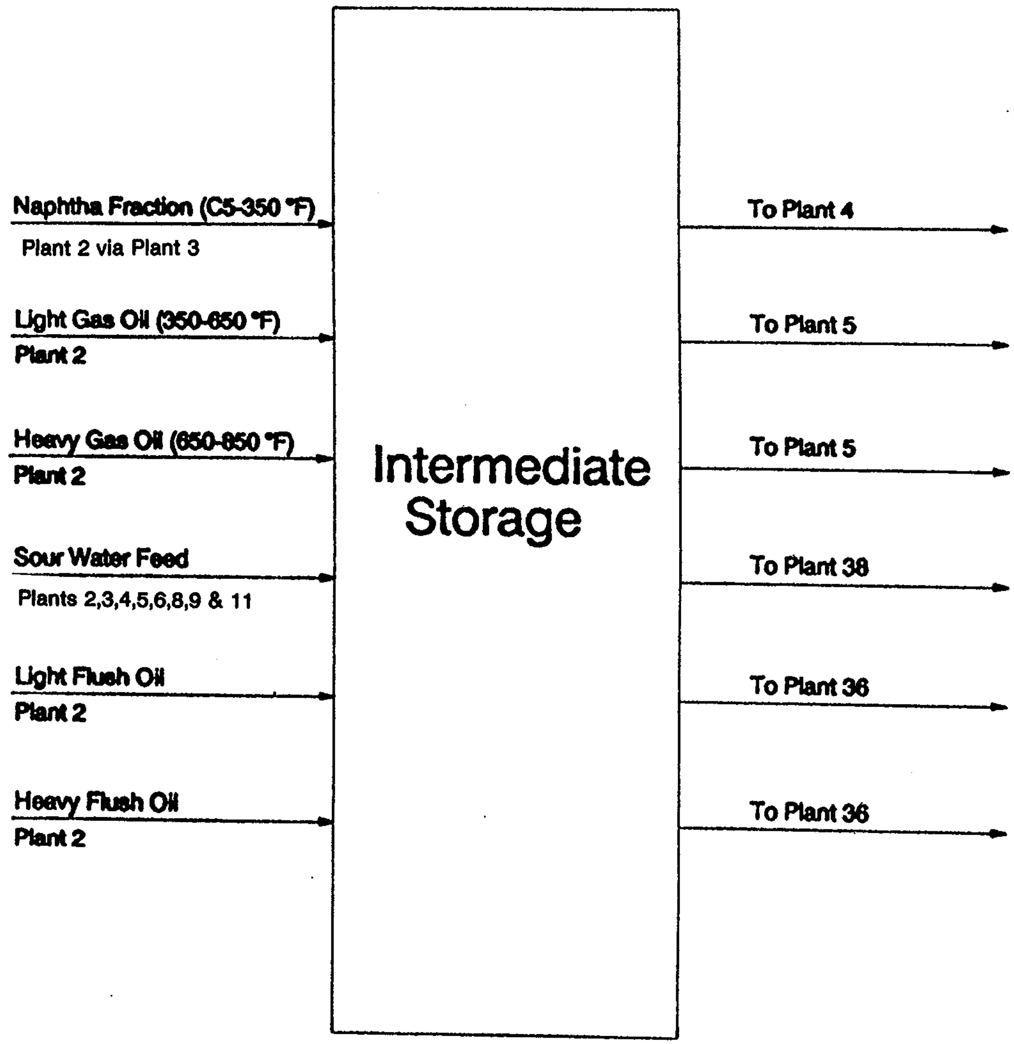


Figure 18.1 Block Flow Diagram



Plant 20
Tankage

Figure 18.2 Block Flow Diagram

Heavy Distillate Product - The heavy distillate product (450-650°F) is delivered from the Gas Oil Hydrotreater (Plant 5) at 135°F and stored in three 269,000 barrel cone roof tanks (20-D103A, B, and C) to provide 30 days storage. The heavy distillate product is pumped via pipeline in 8 hour batches to down stream customers. The product loading pumps (20-G103A and B) are designed for 4375 gpm at 1500 psi which allows delivery of 50,000 barrels to the pipeline over an eight hour period.

Gas Oil Product - The gas oil product (650-850°F) is delivered from the Gas Oil Hydrotreater (Plant 5) at 135°F and stored in three 151,100 barrel cone roof tanks (20-D104A, B, and C) to provide 30 days storage. Steam coils (20-E102A, B, and C) are provided at the bottom of these tanks to maintain the product at a pumpable temperature. The gas oil product is pumped via pipeline in 8 hour batches to down stream customers. The product loading pumps (20-G104A and B) are designed for 4375 gpm at 1500 psi which allows delivery of 50,000 barrels to the pipeline over an eight hour period.

Propane - Propane enters the plant battery limits at 100°F from Plant 3, bypassing the refrigeration unit and going directly to pressurized storage and shipping. Product in excess of shipments is cooled in a refrigeration unit to -48°F, and is stored in refrigerated tanks at atmospheric pressure.

One 25,000 barrel sphere (20-D106) provides 5 days of nonrefrigerated storage. Two 24,200 barrel tanks (20-D105A and B) provide 10 days of refrigerated storage. The tanks are equipped with a refrigeration unit and product heater (20-V103).

When needed, propane product is taken from the refrigerated tanks, heated to 66°F, and sent to pressurized storage. The propane product is then loaded onto tank cars or tank trucks for shipment (see Plant 23, Tank Car, Tank Truck Loading).

Propane product loading pumps (20-G105A and B) are designed for 2000 GPM at 290 psi.

Mixed Butanes - Mixed Butanes (nC4 and iC4) enter the plant battery limits at 100°F from Plant 3 and are sent to pressurized storage. Two 30,000 barrel spheres (20-D107A and B) provide 15 days of nonrefrigerated storage. The butane product is then loaded onto tank cars or tank trucks for shipment (Plant 23, Tank Car, Tank Truck Loading). Butane product loading pumps (20-G106A and B) are designed for 2000 GPM at 150 psi.

Sulfur - Molten sulfur from Sulfur Recovery (Plant 11) is received and stored at 300°F. Two 15,500 barrel cone roof tanks (20-D108A and B) provide 10 days of storage. Steam coils are provided in these tanks to maintain the product at a pumpable temperature. These coils include 20-E103A and B located at the bottom, 20-E104A and B roof coils, and 20-E105A and B suction heaters. The molten sulfur is then loaded onto tank cars or tank trucks for shipment (see Plant 23, Tank Car, Tank Truck Loading). Sulfur product loading pumps (20-G107A, B, and C) are designed for 1000 GPM at 75 psi.

Ammonia - Anhydrous ammonia enters the plant battery limits at 110°F from the Ammonia Plant (Plant 38) and is stored in one 15,000 barrel sphere (20-D109) which provides 5 days of storage. Additional storage consists of two 12,900 barrel refrigerated tanks (20-D110A and B) with a storage capacity of 10 days. These tanks are equipped with a refrigeration unit and product heater (20-V104) which chills the ammonia to -28°F and atmospheric pressure for storage.

Product shipment may be directly from pressurized sphere storage, or product may be taken from the refrigerated tanks, heated to 66°F, and sent to the sphere for shipment. The product is loaded onto tank cars or tank trucks for shipment (Plant 23, Tank Car, Tank Truck Loading). The loading pumps (20-G108A and B) are designed for 2000 gpm at 275 psi.

Phenols - Phenols enter the plant battery limits at 100°F from the Phenol Plant (Plant 39) and are stored in two 2,100 barrel cone roof tanks (20-D111A and B) which provide 15 days of storage. The product is loaded onto tank cars or tank trucks for shipment (Plant 23, Tank Car, Tank Truck Loading). The loading pumps (20-G109A and B) are designed for 660 gpm at 80 psi.

Intermediate Storage

Storage is provided for the following intermediates as shown in the Block Flow Diagram, Figure 18.2:

- One naphtha fraction (C5-350°F) from Coal Gasification (Plant 2) via Plant 3
- One light gas oil fraction (350-650°F) from Coal Gasification (Plant 2)
- One heavy gas oil fraction (650-850°F) from Coal Gasification (Plant 2)
- Sour water from Plants 2, 3, 4, 5, 6, 8, 9, and 11
- Light and heavy flush oils from Plant 2

Naphtha Fraction - The naphtha fraction (C5-350°F) is delivered from the Plant 2 atmospheric tower overhead via Plant 3 as lean oil for storage in one 5,000 barrel tank (20-D112) to provide 6 hours storage. The tank is equipped with a floating roof inside a stiffened dome roof. The naphtha intermediate is delivered to the Naphtha Hydrotreater (Plant 4) with pumps (20-G110A and B) designed for 510 gpm at 50 psi.

Light Gas Oil - The light gas oil (350-650°F) is delivered from a Plant 2 stream consisting of the atmospheric tower sidestripper bottoms and the vacuum tower overhead at about 385°F for storage in one 8,900 barrel cone roof tank (20-D113) to provide 6 hours storage. A steam coil (20-E106) is provided at the bottom of this tank to maintain pumping temperature and the required feed temperature into the Gas Oil Hydrotreater (Plant 5). A mixer package (20-V106) is provided to maintain intermediate homogeneity and pumpability. The light gas oil intermediate is delivered to Plant 5 with pumps (20-G111A and B) designed for 870 gpm at 50 psi.

Heavy Gas Oil - The heavy gas oil (650-850°F) is delivered from a Plant 2 vacuum tower side stream at 366°F for storage in one 5,000 barrel cone roof tank (20-D114) to provide 6 hours storage. Steam coils (20-E107 and E108) are provided at the bottom and suction nozzle, respectively, of this tank to maintain pumping temperature and the required feed temperature into the Gas Oil Hydrotreater (Plant 5). A mixer package (20-V107) is provided to maintain intermediate homogeneity and pumpability. The heavy gas oil intermediate is delivered to Plant 5 with pumps (20-G112A and B) designed for 525 gpm at 50 psi.

Sour Water Feed - Sour water from the process plants, 2, 3, 4, 5, 6, 8, 9, and 11, is accumulated in two 355,500 barrel tanks (20-D115A and B) to provide 5 days storage. Each tank is equipped with a floating roof inside a stiffened dome roof. The tank is equipped with a vapor compressor and recovery system (20-V105). A skimmer package (20-V110) removes any oil residuals from the sour water. The sour water is delivered to Ammonia Removal (Plant 38) with pumps (20-G113A and B) designed for 4000 gpm at 50 psi.

Light Flush Oil - Light flush oil, a mixture of two Plant 2 streams (the atmospheric tower side stream and vacuum tower overhead stream), is accumulated at about 385°F in one 30,100 barrel tank (20-D116) to provide 10 days storage. The tank is equipped with a floating roof inside a stiffened dome roof. A steam coil (20-E109) is provided at the bottom of the tank to maintain the flush oil at a pumpable temperature (300°F). A mixer package (20-V108) is provided to maintain intermediate homogeneity and pumpability. The light flush oil is delivered to the Purge and Flush Oil System (Plant 36) with pumps (20-G114A and B) designed for 85 gpm at 3300 psi. A precoat filter package (20-V101) is utilized to remove particulates from the oil.

Heavy Flush Oil - Heavy flush oil obtained from a Plant 2 vacuum tower side stream is accumulated at 366 ° F in two 54,400 barrel cone roof tanks (20-D117A and B) to provide 10 days storage. Steam coils (20-E110A and B) are provided at the bottom of these tanks to maintain the flush oil at a pumpable temperature (300°F). A mixer package (20-V109) is provided to maintain intermediate homogeneity and pumpability. The heavy flush oil is delivered to the Purge and Flush Oil System (Plant 36) with pumps (20-G115A and B) designed for 265 gpm at 250 psi. A precoat filter package (20-V102) is utilized to remove particulates from the oil.

18.2 Major Equipment List

Given in Table 18.1 below is the major equipment list for this plant.

TABLE 18.1
MAJOR EQUIPMENT LIST
Plant 20 - Tankage

Equipment No	Equipment Description
20-D101A,B,C	Naphtha
20-D102A,B,C	Light Distillate
20-D103A,B,C	Heavy Distillate
20-D104A,B,C	Gas Oil
20-D105A,B	Propane
20-D106	Propane
20-D107A,B	Mixed Butanes
20-D108A,B	Sulfur
20-D109	Ammonia
20-D110A,B	Ammonia
20-D111A,B	Phenols
20-D112	Naphtha
20-D113	Light Gas Oil
20-D114	Heavy Gas Oil
20-D115A,B	Sour Water Feed
20-D116	Light Flush Oil
20-D117A,B	Heavy Flush Oil

NOTES:
1. All tanks are 95% working capacity.
2. All tanks are standard size API 650 except for the spheroids.

TABLE 18.1 - continued

Major Equipment List - continued

Miscellaneous

Equipment No.	Description
20-V101	Precoat Filter, Lt Flush Oil
20-V102	Precoat Filter, Hv Flush Oil
20-V102	Propane Refrig & Heating Sys
20-V102	Ammonia Refrig & Heating Sys
20-V102	Sr Wat Compr/Vap Recv Sys
20-V102	Light Gas Oil Tank Mixer
20-V102	Heavy Gas Oil Tank Mixer
20-V102	Light Flush Oil Tank Mixer
20-V102	Heavy Flush Oil Tank Mixer
20-V102	Dephenol Feed Tank Skimmer Pkg.

Pumps

Equipment (1) No.	Equipment Description
PRODUCTS	
20-G101A,B	Naphtha
20-G102A,B	Light Distillate
20-G103A,B	Heavy Distillate
20-G104A,B	Gas Oil
20-G105A,B	Propane
20-G106A,B	Mixed Butanes
20-G107A,B,C	Sulfur
20-G108A,B	Ammonia
20-G109A,B	Phenols
INTERMEDIATES	
20-G110A,B	Naphtha Fraction
20-G111A,B	Light Gas Oil
20-G112A,B	Heavy Gas Oil
20-G113A,B	Sour Water Feed
20-G114A,B	Light Flush Oil
20-G115A,B	Heavy Flush Oil

NOTES:

1. All pumps are one operating and one spare except for the sulfur pump which is 2 operating and one spare.

TABLE 18.1 - continued

Major Equipment List - continued

Heat Exchangers

Equipment No.	Equipment Description
PRODUCTS	
20-E102A,B,C	Gas Oil Bottom Coil
20-E103A,B	Sulfur Bottom Coil
20-E104A,B	Sulfur Top Coil
20-E105A,B	Sulfur Suction Nozzle Coil
INTERMEDIATES	
20-E106	Light Gas Oil Bottom Coil
20-E107	Heavy Gas Oil Bottom Coil
20-E108	Heavy Gas Oil Suct Heater
20-E109	Light Flush Oil Bottom Coil
20-E110A,B	Heavy Flush Oil Bottom Coil

18.3 Utility Summary

Electricity = 6,758 kW
Cooling Water Circulation = 100 GPM

Steam Consumption is as tabulated below:

Description (psig)	Consumed (lb/hr)	Condensate Returned (lb/hr)
150	22,000	22,000
50	3,300	3,300

19. Plant 21 (Interconnecting Piping Systems)

19.0 Design Basis, Criteria and Considerations

Plant 21 includes the fuel gas blending and distribution system and the interconnecting process and utility piping between process plants and offsites. All above ground and underground piping systems are included except fire water piping which is included in Fire Systems (Plant 33) and plant flare headers which are included in the Flare System (Plant 19). In general, water distribution piping is underground and all other piping is located above ground on pipe racks.

Fuel gas users in the complex include process fired heaters and combustion turbine generators (CTG), the coal drying heaters. Fuel for these users must be clean gas with virtually no sulfur content so that no treatment of stack gases for sulfur removal is needed.

Two types of fuel gas are produced within the complex. One is classified as high BTU gas and the other as medium BTU gas. The high and medium BTU gases have been segregated in two separate distribution systems. Natural gas can be added to either system to meet the fuel gas requirements and to maintain consistency within each system. The gas fired equipment will burn the high BTU gas while the medium BTU gas system goes to Plant 31 to produce steam and power.

Fuel Gas System

The rates and specifications of the plant fuel gas available from two sources within the facility and natural gas are summarized in Table 19.1. The rates and compositions of the first two sources in the table are from the material balances included on the process flow diagrams for the Gas Plant (Plant 3) and Hydrogen Production by gasification (Plant 9). Natural gas composition is based on the analysis used in this complex.

The material balance and equipment design for Plant 3 and Plant 9 were developed based upon a preliminary design basis.

Interconnecting Piping

The interconnecting piping consists of all the process lines and racks connecting one process plant to another, the utility headers and the branches to each process. Pipes are sized based on pressure drop and fluid velocity considerations.

The cooling water system is routed underground and process lines and other utilities are routed on the pipe racks. All the steam, condensate and boiler feedwater lines are insulated. The headers, one for each utility service, include the following:

- 600 psig steam (superheated)
- 150 psig steam (saturated)
- 50 psig steam (saturated)
- Instrument air
- Utility air
- Utility water
- Cooling water supply
- Cooling water return
- 600 psig boiler feedwater
- Potable water
- High BTU fuel gas
- Medium BTU fuel gas
- Natural gas
- Nitrogen Gas

Storm sewer, sanitary sewer, and process wastewater lines are included in the scope of Sewers and Wastewater Treating (Plant 34).

TABLE 19.1

FUEL GAS AVAILABILITY

(All Rates are TPSD)

Stream Component	High BTU Gas From Plant 3	Medium BTU Gas from Plant 9	Natural Gas
Hydrogen	27.0	210.6	--
Carbon Dioxide	0.0	126.6	--
Carbon Monoxide	9.1	240.7	--
Methane	389.7	8.8	1819.5
Ethane	311.3	0.0	108.3
Propane	77.9	0.0	38.4
Water	36.4	0.0	--
N_2	15.3	149.6	33.4
Argon	0.0	29.1	--
O_2	0.0		207.8
TOTAL	866.7	765.4	2207.4
LHV, BTU/SCF (1)	901	259	884
MM BTU/hr (LHV)	1473	1007	3510

(1) Heating values are BTU per scf based on 379 ft^3 per pound-mole. Maximum operating pressure for high BTU gas is 75 psig and minimum pressure 50 psig at the battery limits of any user. Medium BTU gas will have an operating pressure of 300-400 psig as it exits Plant 9.

19.1 Plant Description

Fuel Gas System

There are two fuel gas systems in order to segregate high and medium BTU gases.

Medium BTU Gas

The source of the medium BTU gas is:

- o Desulfurized raw syngas from Gasification and Purification (Plant 9)

Medium BTU gas must be burned within the facility as there is no other use for it in the present concept. It is supplied to the power plant (Plant 31) to produce steam and power. Natural gas may be injected into this stream to satisfy the needs of the gas turbines in the power plant.

High BTU Gas

High BTU gas comes from the following sources:

- o Outside purchased gas (Natural Gas)
- o Desulfurized deethanizer overhead gas from the Gas Plant (Plant 3)

Outside pipeline gas is purchased only to supplement the internally produced gas or for start-up use. The deethanizer overhead gas is burned within the facility. Natural gas can be burned within the facility as needed to balance fuel usage. The high BTU gas distribution system is designed to accom- modate all of these requirements.

All of the streams first pass into a mixing drum to smooth out variations in calorific value that would result from fluctu- ations in flow rate of any of the gases. High BTU gas is to be piped to all in-plant users, with the piping designed so that all users operate entirely on high BTU gas.

Fuel Gas Blending

Two 10' x 5' vertical drums (21-C101 and 102) are provided for mixing natural gas with the medium and high BTU gases, respectively.

Interconnecting Piping

Interconnecting pipings consists of all process lines connecting the plants, the utility headers, and the branches to each plant.

The process and utility line sizes are summarized in the major line list section below.

19.2 Major Equipment Lists and Major Lines Summary

Equipment No.	Type	Description
21-C101	Drum	Natural and medium Btu gas mixing, vertical, 10 ft. X 5 ft.
21-C102	Drum	Natural and high Btu gas mixing, vertical, 10 ft. X 5 ft.

Process Lines

Line Contents	From	To	Size (in)
Naptha Product	Plant 4	Plant 20 Tank	6
Light Distillate Product	Plant 5	Plant 20 Tank	4
Heavy Distillate Product	Plant 5	Plant 20 Tank	6
Gas Oil Product	Plant 5	Plant 20 Tank	6
Propane Product	Plant 3	Plant 20 Tank	2
Mixed Butanes Product	Plant 3	Plant 20 Tank	2
Sulfur Product	Plant 11	Plant 20 Tank	3
Ammonia Product	Plant 38	Plant 20 Tank	2
Phenol Product	Plant 39	Plant 20 Tank	1
Naptha Product	Plant 20 Tank	Plant 20 Pump	22
Light Distillate Product	Plant 20 Tank	Plant 20 Pump	22

Line Contents	From	To	Size (in)
Heavy Distillate Product	Plant 20 Tank	Plant 20 Pump	22
Gas Oil Product	Plant 20 Tank	Plant 20 Pump	22
Naptha Product	Plant 20 Pump	Plant 22 Pipeline	14
Light Distillate Product	Plant 20 Pump	Plant 22 Pipeline	14
Heavy Distillate Product	Plant 20 Pump	Plant 22 Pipeline	14
Gas Oil Product	Plant 20 Pump	Plant 22 Pipeline	14
Propane Product	Plant 20 Pump	Plant 23 Tank Car/Truck	8
Mixed Butanes Product	Plant 20 Pump	Plant 23 Tank Car/Truck	8
Sulfur Product	Plant 20 Pump	Plant 23 Tank Car/Truck	10
Ammonia Product	Plant 20 Pump	Plant 23 Tank Car/Truck	8
Phenol Product	Plant 20 Pump	Plant 23 Tank Car/Truck	6
Naphtha Intermediate	Plant 2	Plant 20 Tank	6
Light Gas Oil Intermediate	Plant 2	Plant 20 Tank	6
Heavy Gas Oil Intermediate	Plant 2	Plant 20 Tank	6
Sour Water Header	Plants 2, 3, 4, 5, 9, 11	Plant 20 Tank	12

Line Contents	From	To	Size (in)
Light Flush Oil Intermediate	Plant 2	Plant 20 Tank	3
Heavy Flush Oil Intermediate	Plant 2	Plant 20 Tank	4
Naphtha Intermediate	Plant 20 Tank	Plant 2	6
Light Gas Oil Intermediate	Plant 20 Tank	Plant 2	6
Heavy Gas Oil Intermediate	Plant 20 Tank	Plant 2	6
Sour Water Header	Plant 20 Tank	Plant 38	12
Light Flush Oil Intermediate	Plant 20 Tank	Plant 2	3
Heavy Flush Oil Intermediate	Plant 20 Tank	Plant 2	4
Make-up Water	Plant 32	Plant 9	8
Raw Water	Raw Water Source	Plant 32	14
50 psig Steam	50 psig Steam Header	Plant 31	24

Utility Lines

Utility Header	Size (in)
600 psig Superheated Steam (720°F)	10
600 psig Saturated Steam	10
150 psig Saturated Steam	24
50 psig Saturated Steam	36
Instrument Air	3

Utility Header	Size (in)
Utility Air	2
Utility Water	6
Cooling Water Supply	48
Cooling Water Return	48
600 psig Boiler Feedwater	6
Potable Water	3
High Btu Fuel Gas	8
Medium Btu Fuel Gas	10
Natural Gas	6
Nitrogen Gas	26

20. Plant 22 (Product Shipping)

20.0 Design Basis, Criteria and Considerations

Plant 22 provides the pipeline and metering system for delivery of the final oil products from the hydrotreaters (Plants 4 and 5) to down stream customers.

The equipment for this plant includes the appropriate length of 20 in. schedule 40 pipe for product delivery to down stream customers and a meter for tracking the amount of product transfered for accounting and billing purposes. The meter is provided with a 16 in. proving loop for meter testing and calibration.

The pipeline is designed to carry 4375 gpm which allows 50,000 barrels of oil product to be delivered in 8 hour batches. The pressure drop should not exceed 500 psi for every 50 miles of pipe.

Dual meters are required to assure proper recording of product delivery quantities in case of single meter failure.

20.1 Plant Description

Pipeline

A 20 in. schedule 40 pipe is provided for product delivery to down stream customers. The pressure drop in the pipe is 0.15 psi per 100 ft. of pipe at the maximum flow rate of 4375 gpm.

Metering System

A dual metering system is provided for tracking the amount of product transfered for accounting and billing purposes. The meter consists of one active and one spare system. Product flow rates are measured by an in-line turbine element which transfers an electronic signal to a microprocessor. The microprocessor converts the electronic signal to digital data which is stored for future retrieval. The meter is provided with a 16 in. pipe diameter by 60 ft. long proving loop for meter testing and calibration.

Pig

A 20 in. pig is provided for periodic cleaning of residuals and debris in the pipeline.

20.2 Major Equipment List

Major Equipment List for the plant is shown in Table 20.1 below:

Table 20.1

Equipment No.	Type	Description
22-L101	Pipeline	20 in., schedule 40
22-V101	Metering System with Prover Loop	16 in. prover loop
22-V102	Pipe Cleaning Pig	For 20 in. pipe
22-V103	Pig Launcher	For launching pipe cleaning pig into the pipeline

20.3 Utility Summary

The electrical utility requirements for the product delivery pumps are included in Tankage (Plant 20).

21. Plant 23 (Tank Car/Tank Truck Loading)

21.0 Design Basis, Criteria and Considerations

The products are generally pumped from the storage tanks to the loading points at the required rate (See Table 21.1). One pump for each product delivers the required flow rate for phenol, ammonia, propane, and butane; however, two pumps are required for pumping molten sulfur. All operating pumps are provided with a spare. Loading pumps are included in Tankage (Plant 20). Nozzles are provided at both the Tank Car and Tank Truck loading racks such that any product can be loaded at two or more bays.

Each product is piped by a separate line to the loading racks, then branched to different loading nozzles.

All products are loaded at ambient temperatures (100 °F) except molten sulfur which is loaded at 300 °F.

21.1 Plant Description

Rail Tank Car Loading

Standard loading arms with telescopic nozzles and swivel joints are provided for top loading products such as phenol, and molten sulfur. For volatile products such as propane and butane, and anhydrous ammonia, loading arms with hose connections for bottom loading are used.

Tank Truck Loading

Top loading nozzles are used for phenols and molten sulfur. Nozzles with bottom loading hose connections are used for anhydrous ammonia and propane.

Piping and Valves

Piping and valves are carbon steel and conform to the required design conditions. Hydraulic shock absorbers are provided for sudden shut-offs and static neutralizing devices are sued for phenols and molten sulfur. Heat arcing and insulation are provided for molten sulfur at 300 °F. Vapor lines are provided from the tank truck and tank car loading racks in accordance with the vapor recovery system concept.

The loading nozzles have connections for bleeders and drains. Hand-operated block valves are provided just upstream of the loading nozzles and are accessible from the platforms. Connections for nitrogen purge are provided at the loading nozzles for products for which this is necessary.

Table 21.1
Truck & Tank Loading Facilities

PRODUCTS	OPERATING °F	LOADING (PSIG)	PUMP CAPACITY (GPM)
Propane (non-refrigerated)	100 °	288	2000
Butane (non-refrigerated)	100 °	150	2000
Anhydrous Ammonia (non-refrigerated)	100 °	272	2000
Phenols	100 °	80	660
Molten Sulfur	300 °	76	1000

Instrumentation

The following instrumentation is provided at each loading nozzle location:

1) Automatic excess flow shut-off control valves for emergency shut-down in case of a broken hose in the loading nozzles for all products.

2) Positive displacement meters with totalizers are provided for each product.

3) Digital counters with automatic printout are installed at the loading location.

4) Instrument air is provided at the loading racks.

Loading Platforms

These consist of structural steel, grating and handrailing. Swinging catwalks with counterweights are used for each loading bay for operator accessibility.

Tank Car Platform

The products are piped to three loading bays. The loading platform is 340 feet long, 4 feet wide and 12 feet in height. Stairs are provided at the two ends of the platform and on the sides.

Tank Truck Loading Platform

The dimensions of this platform are 10'X 8'X 10' with a 20 foot roof covering the loading area and one set of stairs, and a catwalk on either side.

Tank Truck Scale

A truck scale is provided for weighing both empty and loaded trucks.

Railroad Spur

A railroad system is provided for loading rail tank cars.

Capacity of tank cars	30,000 gallons
Capacity of tank trucks	10,000 gallons
Loading time	15 minutes
Ambient temperature	100 °F

PRODUCTS

1) Propane
2) Mixed Butanes
3) Ammonia
4) Phenol
5) Molten Sulfur

Loading Arrangement

The following factors are of primary importance:

- Butane will be shipped only in dedicated tank cars
- Propane, ammonia, and liquid sulfur will be filled in tank cars and/or dedicated tank trucks.
- Phenol will be filled in non-dedicated tank trucks.

Loading Bays

Based on the filling time required for tank cars/trucks, time required for completing documentation, line-up and position of tank cars/trucks, inclement weather, it is proposed to have:

- Three Loading Bays for tank trucks each with 2 filling positions. One bay will be dedicated to Propane (located at a safe distance from any source of flame or sparks), one for Liquid Sulfur, and one for both Ammonia and Phenol.

- Six Loading Bays for tank cars, each with one filling position. All bays are capable of filling all products.

Tank Truck Scale

A truck scale is provided for weighing empty and loaded trucks. The products, in general, are loaded per the level marker provided in tank cars and by set stop meters for tank cars. All trucks are weighed in and out. A sump pump is provided in the weighing pit to remove rain and cleaning water.

Sprinkler

A sprinkler fire water system is provided for the tank truck loading rack. The system is automatically energized in case of fire, covering the entire loading rack area with water at a minimum density of 0.25 GPM/ft^2.

Fire Hydrants and Extinguishers

Fire Hydrants and monitor are provided for the tank car loading rack in lieu of a deluge system. Portable day chemical fire extinguishes are also provided, at least at two locations for the tank truck loading rack and at fire places for the tank car loading tank.

Spill Clean-up

The loading area is paved and sloped towards the effluent drainage system. Utility steam and water provided in the rack are used for clean-up purposes.

Communication System

Communication systems are provided at both the Tank Car and Tank Truck locations and tied to the plant system. These include:

- Telephone System
- Portable Radios
- Portable Telephones

21.2 Major Equipment List

The summary of the equipment for plant 23 is presented in Table 21.2.

21.3 Utility Summary

Connected Load = 226 KW, Actual Load = 90 KW (intermittent)
Electricity = 460 V, 3 phase, 60 Hz (27 KW)
Water = 50 GPM (For Cleaning) (intermittent)
Steam, Utility = 1000 lbs/hr (for cleaning)
Nitrogen, Utility = 50,000 SCFH (for purging)

TABLE 21.2

MAJOR EQUIPMENT LIST

PLANT 23 - TANK CAR/TANK TRUCK LOADING

Compressors

Equipment No.	Equipment Description
23-K101A/B	Propane Vapor Recovery
23-K102A/B	Butane Vapor Recovery
23-K103A/B	Ammonia Vapor Recovery
23-K104A/B	Sulfur Vapor Recovery

Pumps

Equipment No.	Equipment Description
23-G101A/B	Weigh Scale Sump Pump

Miscellaneous

Equipment No.	Equipment Description
23-M101	Propane Tank Truck Loading
23-M102	Ammonia/Phenol Truck Loading
23-M103	Sulfur Tank Truck Loading
23-M104	Tank Car Loading Platform
23-V101	Tank Truck Weigh Scale
23-J101	Positive Displacement Meters

22. Plant 24 (Coal Refuse and Ash Disposal)

22. Design Basis, Criteria and Considerations

This plant is for the disposal of coal refuse from Coal Cleaning and Preparation (Plant 1) and ash or slag from the Hydrogen Production by Coal Gasification (Plant 9).

The coal refuse consists of fine and coarse material which requires separate methods of disposal. The coarse coal refuse and ash are conveyed to the coal mine via conveyor belt for disposal in land reclaimation.

The fine coal refuse material is piped as a coal-water slurry to a settling basin. The bottom of the settling basin is scraped continuously to move the fine refuse slurry to the basin shores. From the basin shores, bulldozers spread the material for air drying. After the spread material is sufficiently dry to have the consistency of a filter cake, bulldozers load the material onto the conveyor for transferral back to the mine.

The coarse coal refuse and ash are transported back to the coal mine via conveyor belt for disposal as land reclamation. The fine coal refuse is piped to a settling basin. The bottom of the settling basin is scraped continuously to move the fine refuse slurry to the basin shores. From the basin shores, bulldozers spread the material over 30 acres of land for air drying. This area of land provides up to 30 days of fine refuse storage to a maximum refuse depth of 1 ft. After the spread material is sufficiently dewatered to have the consistency of a filter cake, bulldozers load the material onto the conveyor for transferral to the mine.

22.1 Plant Description

Conveyor belt (24-T101) is provided for transferring coarse coal refuse from Coal Cleaning and Preparation (Plant 1), ash or slag from Hydrogen Production by Coal Gasification (Plant 9) back to the mine site.

The fine coal refuse material is piped as a slurry to the settling basin. The basin scraper (24-T104) operates continuously to move the fine refuse slurry material to the shores of the basin. From the basin shores, bulldozers (24-T103A and B) spread the wet material over 30 acres of land for air drying. Material which is sufficiently dewatered to have the characteristics of a vacuum filter cake, is loaded by the bulldozers onto the conveyor belt (24-T101) for transferral to the mine as land reclamation.

22.2 Major Equipment List

The major equipment list for this plant is shown in Table 22.1 below.

Table 22.1
EQUIPMENT LIST

Equipment No.	Type	Size or Capacity
24-T101	Refuse conveyor belt	8500 TPD
24-T103A,B	Bulldozer	480 yd^3/hr
24-T104	Settling Basin Scraper	1545 tons solid/day

22.3 Utility Summary

Utility requirements for processing Plant 1 coal refuse are included with the Plant 1 utility balance and are therefore not repeated here. The requirements summarized below are for Plant 9 coal ash disposal only.

Electricity, kW

Condition	Consumed
Operating	100

Table 23.1 - continued

Chemical or Catalyst	For Plant No.	Quantity Required for Start-Up	Consumption
30% Ammonia	31	7,000 lbs	936 lb/day
Sodium Sulfite	31	2,000 lbs	240 lb/day
Polymer & Chelant			
Disodium	31	1,000 lbs	144 lb/day
Phosphate			
	31	1,000 lbs	144 lb/day
Alum	32	30,000 lbs	4,680 lb/day
Polymer	32	2,000 lbs	156 lb/day
98% H_2SO_4	32	15,000 gals	15,680 lb/day
50% NaOH	32	30,000 gals	38,754 lb/day
Polymeric			
Dispersion Non-	32	7,000 lbs	1,624 lb/day
ionic Surfactant			
	32	300 lbs	24 lb/day
Chlorine	34	2000 lbs	350 lbs/day
Polymer	34	3000 lbs	450 lbs/day
Pac	34	6000 lbs	2,000 lbs/day
Phosphoric Acid as 100% H_3PO_4	38	11,0331 lb	3,460 lb/day
Dephenoliz-ation Solvent	39	7,912 lb	170 lb/day

23. Plant 25 (Catalyst and Chemical Handling)

23.0 Design Basis, Criteria and Considerations

This plant provides storage and handling for catalysts and chemicals used in all the plants. Additionally, it provides a consolidated location for tracking catalyst and chemical start-up and daily consumption requirements.

Plants requiring chemicals or catalysts include 2 (Coal Liquefaction), 3 (Gas Plant Separation), 4 (Naphtha Hydrotreater), 5 (Gas Oil Hydrotreater), 6 (Hydrogen Purification), 8 (Rose Solvent Recovery), 9 (Hydrogen Production via Coal Gasification), 11 (By-Product Sulfur Recovery), 32 (Raw, Cooling, and Service Water), 38 (Ammonia Removal), and 39 (Phenol Removal).

The equipment for this plant includes an enclosed warehouse for storing chemicals and catalysts and forklifts for transporting pallets of chemicals or catalysts into or out of the warehouse.

A warehouse is required to collect all chemicals into one area for distribution to the various plants as needed. Additionally, the warehouse is used as a temporary storage for spent catalyst that must be returned to the catalyst vendor for regeneration at the vendor's facilities.

This plant identifies all major plant catalyst and chemical requirements for startup and continuous operation.

23.1 Plant Description

A 6000 square foot chemical and catalyst warehouse (25-R101) is provided for temporary storage of chemicals. Electric forklifts (25-T101A and B) are provided for transporting pallets of chemicals or catalysts into or out of the warehouse.

23.2 Major Equipment List

Equipment No.	Type	Description
25-R101	Chemical & Catalyst Warehouse	100 ft. x 60 ft.
25-T101A,B	Forklift	Electric Motor, 25 hp

23.3 Utility Summary

Electricity = 50 kW

23.4 Chemical and Catalyst Summary

Plant 25 provides storage and handling for chemicals and catalysts used in all the plants. Table 23.1 below summarizes the start-up and consumption rates for the various chemicals or catalysts.

TABLE 23.1

Chemical or Catalyst	For Plant No.	Quantity Required for Start-Up	Consumption
Amocat-1C, 1-1/2" Extrudate	2	2,253,000 lb	68,130 lb/day
MEA	3	100 bbl	20 gal/day
Catalyst	4	80,000 lb	3 year life
Catalyst	5	490,000 lb	3 year life
MEA	6	1,200 bbl	30 gal/day
Rose Solvent	8	10,000 bbl	300 gal/day
BASF K8-11 or Haldor Topsoe SSK Catalyst	9	19,000 ft^3	3 year life
2" SS Pall Ring Packing	9	3,400 ft^3	3 year life
2" CS Pall Ring Packing	9	4,300 ft^3	3 year life
Methanol	9	2,500 bbl	200 gal/day
Claus Catalyst Kaiser S-201	11	18,400 ft^3	3 year life
SCOT Catalyst	11	6,500 ft^3	3 year life
2" SS Pall Ring Packing	11	2,500 ft^3	3 year life
MDEA	11	500 bbl50 gal/day	