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TVA COAL-GASIFICATION
COMMERCIAL DEMONSTRATION PLANT PROJECT
VOLUME 2
BASIS OF STUDY
ASSESSMENTS AND PROCESS SELECTION

FINAL REPORT
November 1, 1980



Foster Wheeler Energy Corporation

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PROCESS PLANTS DIVISION



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VOLUME 2 BASIS OF STUDY, ASSESSMENTS AND PROCESS SELECTIONS
TABLE OF CONTENTS

Section

1.0	Introduction
2.0	Basis of Study
2.1	Overview
2.1.1	Design Features
2.1.2	Coal-Handling Facilities
2.1.3	Buildings and other Support Structures
2.1.4	Coal-Handling Facilities
2.1.5	Other Support Facilities
2.2	TVA Design Criteria
2.3	Foster Wheeler Supplement
3.0	Assassments and Process Selections
3.1	Gas Cleanup
3.2	Oxygen Purity
3.3	Coal Washing
3.4	Sulfur Recover/
3.5	Equipment Drives
3.6	Flant Reliability, Spares & Gas Storage
3.7	Gas Delivery Pressure
3.8	Effects of Scale
3.9	Load Changes

SECTION 1.0INTRODUCTION

The Tennessee Valley Authority (TVA) is considering the design, construction, and operation of a commercial scale coal gasification facility to produce a clean, medium BTU fuel gas (MBG). The project includes all process and support systems required to convert approximately 20,000 tons per day of Kentucky #9 bituminous coal, as fed to the gasifiers, into MBG equivalent to about 300 billion BTU per day.

The first phase of the proposed project involves conceptual design, environmental and siting studies and economic analyses of commercial plants emphasizing the following gasification technologies:

- Babcock & Wilcox entrained flow gasifier
- Lurgi dry ash gasifier
- BGC/Lurgi slagging gasifier
- Texaco entrained flow gasifier
- Koppers Totzek entrained flow gasifier

Foster Wheeler's effort in this Phase I study was divided into the following major tasks:

- Establish a consistent basis for study
- Assess processing alternatives and recommend process selection
- Develop baseline conceptual designs for commercial plants based on each of the five gasification processes
- Prepare a preliminary environmental assessment for each of the five baseline design cases
- Develop preliminary economics and sensitivity analyses for each of the five design cases
- Recommend areas where further study and engineering evaluations are required

Foster Wheeler's work concerning basis of study and assessments/process selection is summarized in this volume.



2.1 OVERVIEW

2.1.1 Design Features

- Plant and unit size: The plant will consist of four or more identical modules, each with a capacity of approximately 5,000 tons of coal per day as delivered to the gasifiers. The exact size of each of the modules is to be recommended by the Contractor, based on the requirement that the coal throughput would permit whole numbers of gasifiers (two or more) and based on standard size availability of other critical plant components. The exact number of plant modules will be recommended by the Contractor based on technical and economic considerations.
- Construction schedule: Modules are to be constructed sequentially on a schedule that permits convenient procurement of equipment and efficient use of labor. Ancillary facilities which do not economically allow sequential expansion are initially sized and built to serve the needs of the entire plant. Such facilities include coal receiving and handling, slag and ash storage, and water treatment.
- Equipment selection: In the absence of any significant economic advantages, high capital investment with corresponding low operating cost is preferred over the case of low capital investment with corresponding high operating cost.
- Plant life: Each module shall be designed to operate for 20 years beyond its startup date. The plant hardware shall be assigned no salvage value.
- Operating requirement: The module shall be designed to operate at 100 percent capacity for 90% of the time.
- Electric power: Electricity, as required for the plant, will be furnished by TVA at the plant boundary at a medium voltage level (4.16 kV, 6.9 kV, and/or 13.8 kV) from independent sources. The plant requires power for safe shutdown in case of TVA power failure. The plant design includes an emergency power system, consisting of a 5000 KW diesel generator set.



- Transportation access: The site is accessible only by barge and road. Specifications for barge and road receiving facilities are described in Section 2.1.2. It shall be assumed that the site will not receive nor ship materials by rail. This however, does not preclude the possible use of a rail system for materials handling within the plant boundary.

In addition to handling the necessary coal traffic, the barge facility shall be arranged and sized to receive other plant consumables (e.g. char, limestone, chemicals, catalysts, etc.) and to ship immediately salable byproducts (e.g. sulfur).

2.1.2 Coal-Handling Facilities

- System has been designed for coal receipts by barge and sized to handle total coal needs of the plant, plus truck receiving facilities capable of receiving five percent of total needs of the plant.
- Receiving capacities are based on two shifts per day (14 hours), five days per week operation.
- All equipment components are designed for continuous operation.
- Facilities are in accordance with requirements of NFPA and OSHA.
- Auxiliary systems are provided as required for environmental compliance.
- Service air and water are provided throughout system.
- Barge receiving facilities: Dock provided with elevator type unloader(s); barge pulling and positioning system; barge clean-out system; and sufficient room for mooring twenty-four loaded and twenty-four empty jumbo (195 ft. x 35 ft.) barges.
- Surge hoppers are provided to enable "smoothing" of receiving rate. Belt scales are provided from surge to breakers.
- Breakers or crushers for handling as delivered coal.
- Conveying from breakers to dead storage and direct to live storage is provided.
- Dead storage area is sized to accommodate a 90-day supply of coal, with pile height to suit stocking-out and reclaim operations by tractor scrapers. Pile sides angle or repose to be 35° from horizontal, (maximum).



- Tractor-scraper reclaim from dead storage is used.
- Enclosed live storage is provided for 56-hour plant operation.
- Tramp-iron removal facilities and belt-weighing scales are provided for coal being reclaimed from live and dead storage.
- Dual systems are provided for other required systems such as sampling, etc.
- Coal-handling system controls are as follows:
 - a) Total system control from one location.
 - b) Receiving and reclaiming systems provided with separate controls.
 - c) Normal interlocks included in each system.
 - d) Barge unloader controlled locally.

2.1.3 Buildings and other Support Structures

- Listed below are the buildings (or structures) that will constitute the support facilities for the main plant. For the buildings listed, the following facilities are included:
 - a) Toilets and lockers
 - b) Lunch room
 - c) Janitorial supply and storage room
 - d) Utilities equipment room
 - e) Office space or room
 - f) Material or equipment storage room
 - g) Assembly or conference room(s)

The consideration for selecting which facilities are to be provided in each building include: (a) Time and motion efficiency of workers (relative location of building), (b) Size of work force, and (c) Physical, sociological and psychological needs of the work force.

2.1.4 Coal-Handling Facilities

- Barge Unloading and Crew Building
 - a) Control booth
 - b) Warming room
- Breaker and Sample Building
 - a) Lab



- Hopper Building
- Transfer Stations
- Conveyors
- Conditioner Building
 - a) Electrical control room
 - b) Lab
- Coal Silos
 - a) Silo-filling houses
- Surge Hopper (at barge unloading facility only).
- Coal Handling Maintenance and Storage Facilities
 - a) Yard Maintenance Building (utility building)
 - 1. Equipment repair shop
 - 2. Storage for repair parts
 - 3. Fuel storage
 - b) Storage facilities for conveyor parts

2.1.5 Other Support Facilities

- Access Control Portal
 - a) Entrance area
 - b) Control room
- Office Building
 - a) Reception area
 - b) Administrative offices
 - c) Technician offices, library, and computer rooms
 - d) Field engineers' offices
- Service Bay
 - a) Shops
 - b) Power stores
 - c) Medical unit
 - d) Classrooms
- Lab Building
 - a) Test labs
- Outage Facility
 - a) Storage for outage parts



- General Yard Storage Area (open)
- Air Separation Building
- Scale House at Weigh Station
- Sulfur Loading Facility
 - a) Operator's shelter
- Intake Pumping Station
- Sewage Treatment Plant
- Fire Protection Equipment Housing
- Fire Protection Deluge Building
- Environmental Data Station
 - a) Instrument room
- Visitor Facility
 - a) Display area
 - b) Reception office
- General Site
 - a) Fencing
 - b) Yard lighting
 - c) Planting and seeding
 - d) Driveways and walks
 - e) Parking lot
- Steam Generation Facility
 - a) Boiler
 - b) Control room
 - c) Particulate removal
 - d) Labs (air and water quality)
- Water Treatment Plant
- Waste Treatment Plant
- Cooling Tower Blowdown
- Potable Water Plant

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DESIGN CRITERIA
FOR
CONCEPTUAL DESIGNS & ASSESSMENTS
OF TVA'S COAL GASIFICATION
DEMONSTRATION PLANT

Tennessee Valley Authority

March 1980

DESIGN CRITERIA
for
Conceptual Designs and Assessments
of TVA's Coal Gasification
Demonstration Plant

Tennessee Valley Authority
March 1980

FOREWARD

This document establishes a consistent basis for the conceptual designs and assessments of the coal gasification plant. It does not necessarily reflect any decisions by TVA regarding ultimate plant location, configuration, product mix, or process selection. In the event of conflict between criteria and assumptions previously presented to the contractors, the information presented herein shall take precedence.

TABLE OF CONTENTS

	Page
Foreword	i
Summary	iv
Section 1 Plant Design Considerations	1-1
1.1 Design Features	1-1
1.2 Coal Receiving and Handling Facilities	1-2
1.3 Building and Other Support Structures	1-3
1.4 Codes, Standards, and Other Design Considerations	1-6
Section 2 Characteristics of Raw Materials and Specifications for Plant Outputs	2-1
2.1 Coal Characteristics	2-1
2.2 Product Gas Specifications	2-3
2.3 Byproducts Specifications	2-3
2.4 Byproducts Disposition	2-3
Section 3 Site Information	3-1
3.1 Introduction	3-1
3.2 Annual Extreme and Average Site Conditions	3-2
3.3 Monthly Average Site Conditions	3-3
3.4 Percent Occurrence of Wind Speed Classes	3-3
3.5 Percent Occurrence of Wind Direction	3-3
3.6 River Water Analysis	3-4
Section 4 Environmental Control Guidelines	4-1
4.1 Air Emissions	4-1
4.2 Liquid Effluents	4-1
4.3 Solids and Liquids Impoundment	4-1
Section 5 Assumptions for Economic Assessments	5-1
5.1 Module Life	5-1
5.2 Module Service Factor	5-1
5.3 Capital and Operating Cost Estimates	5-1
5.4 Financing	5-1
5.5 Economic Evaluation Rate	5-1
5.6 General and Administrative Expenses	5-1
5.7 Schedule of Cash Flow by Year	5-1
5.8 Product Gas Cost	5-1
5.9 Coal Cost	5-1
5.10 Limestone Cost	5-1
5.11 Coke Cost	5-1
5.12 Byproduct Credit	5-1
5.13 Excess Electricity Sales	5-1

TABLE OF CONTENTS
(continued)

		Page
Section 5	Assumptions for Economic Assessments (continued)	
	5.14 Coal Fines Sales	5-1
	5.15 Land Cost	5-1
	5.16 Clearing and Grubbing Cost	5-1
	5.17 Excavation Cost	5-1
	5.18 Fill Cost	5-1
	5.19 Electric Power Cost	5-1
	5.20 Construction Labor	5-2
	5.21 Operating Labor	5-2
	5.22 Maintenance Labor	5-2
	5.23 Escalation Rates	5-2
	5.24 Sensitivity Analysis	5-2
	5.25 Byproduct Values for Sensitivity Analysis	5-2
Table		Page
2.1	Paradise Steam Plant Peabody No. 9 Deep Mine Raw Plant Feed	2-4
2.2	Float and Sink Analysis	2-5
2.3	Float and Sink Analysis	2-6
2.4	Analysis of Coal Sample Used in Washability Tests	2-7
5.1	Electric Power Cost	5-3
5.2	Construction Labor Cost	5-4
5.3	Operating Labor Cost	5-5
5.4	Maintenance Labor Cost	5-8
5.5	TVA Power Plant Operations Manning	5-9
5.6	TVA Power Plant Maintenance Manning	5-10
5.7	Escalation Rates	5-11
5.8	Sensitivity Analysis Applied to Cost of Gas	5-12
5.9	Byproduct Values for Sensitivity Analysis	5-13

SECTION 1

PLANT DESIGN CONSIDERATIONS

1.1 Design Features

- 1.1.1 Plant and unit size: The plant will consist of four or more identical modules, each with a capacity of approximately 5,000 tons of coal per day as delivered to the gasifiers. The exact size of each of the modules is to be recommended by the Contractor, based on the requirement that the coal throughput would permit whole numbers of gasifiers (two or more) and based on standard size availability of other critical plant components. The exact number of plant modules will be recommended by the Contractor based on technical and economic considerations, and on the stipulation that the total plant capacity shall be approximately 20,000 tons of coal per day as delivered to the gasifiers
- 1.1.2 Construction schedule: Modules are to be constructed sequentially on a schedule that permits convenient procurement of equipment and efficient use of labor. Ancillary facilities which do not economically allow sequential expansion shall be sized and built to serve the needs of the entire plant. Such facilities include coal receiving and handling, slag and ash storage, water treatment, and any others which, subject to TVA concurrence, the Contractor deems appropriate.
- 1.1.3 Equipment selection: In the absence of any significant economic advantages, high capital investment with corresponding low operating cost is preferred over the case of low capital investment with corresponding high operating cost.
- 1.1.4 Plant life: Each module shall be designed to operate for 20 years beyond its startup date. The plant hardware shall be assigned no salvage value.
- 1.1.5 Operating requirement: Each of the modules shall be designed to operate at a stream factor of 90 percent. In systems which have surge capacity, redundant or spare components may be eliminated if those components can be repaired or replaced well within the total system's surge capacity time using normal plant maintenance personnel. In such case the cost of needed spare parts inventory shall be included in the working capital estimate.
- 1.1.6 Electric power: Electricity, as required for the plant, will be furnished by TVA at the plant boundary at a medium voltage level (4.16 kV, 6.9 kV, and/or 13.8 kV) from independent sources. If the plant requires power for safe shutdown in case of TVA power failure, the plant design shall include an emergency power system tailored for that purpose.

- 1.1.7 Transportation access: The site is accessible only by barge and road. Specifications for barge and road receiving facilities are described in section 1.2. It shall be assumed that the site will not receive nor ship materials by rail. This however, does not preclude the possible use of a rail system for materials handling within the plant boundary.

1.2 Coal Receiving and Handling Facilities

- 1.2.1 The system shall be designed for coal receipts by barge and sized to handle total coal needs of the plant, plus truck receiving facilities capable of receiving 5 percent of total needs of the plant.
- 1.2.2 In addition to handling the necessary coal traffic, the barge facility shall be arranged and sized to receive other plant consumables (e.g., char, limestone, chemicals, catalysts, etc.) and to ship immediately salable byproducts (e.g., sulfur). The requirements for additional barge loading facilities not included in the base-case design needed to ship slag and other future salable byproducts shall be identified by the contractors.
- 1.2.3 The following is provided as a guideline only: The barge dock shall be provided with elevator-type unloader(s), barge pulling and positioning system, barge cleanout system, and sufficient room for mooring 24 loaded and 24 jumbo (195 x 35 ft) barges.
- 1.2.4 The river can handle barge traffic consisting of tows up to 3 barges abreast. All channels are a minimum of 200' wide and 11' deep. The dam locks can accommodate up to 9 jumbo (35' x 195') barges per lift or 8 jumbo barges plus a tow boat per lift. Other vessels may be tied to the lock wall during lift operations. A normal tow consists of 15 jumbo barges plus a tow boat. Maximum load per jumbo barge is 1,500 tons of coal. Maximum draft of each fully loaded barge is 9'.
- 1.2.5 Receiving capacities shall be based on two shifts per day (14 hours), five days per week operation.
- 1.2.6 All equipment components shall be designed for continuous operation.
- 1.2.7 All facilities shall be in accordance with requirements of NFPA and OSHA.
- 1.2.8 Auxiliary systems shall be provided as required for environmental compliance.
- 1.2.9 Service air, water, and steam (if needed for thawing) shall be provided throughout the system.
- 1.2.10 A surge hopper shall be provided to enable "smoothing" of receiving rate with belt scales provided from surge to breakers.

- 1.2.11 Breakers or crushers shall be provided for handling as-delivered coal.
- 1.2.12 Conveying from breakers to dead storage and direct to live storage shall be provided.
- 1.2.13 Dead storage area shall be sized to accommodate a 90-day supply of coal, with pile height to suit stocking-out and reclaim operations by tractor scrapers. Maximum angle of repose for pile sides shall be 45° from horizontal. Assume 70 lb/cu ft as density of compacted coal in dead storage.
- 1.2.14 Tractor-scraper reclaim from dead storage shall be used.
- 1.2.15 Enclosed live storage shall provide for 56-hour plant operation.
- 1.2.16 Tramp-iron removal facilities and belt-weighing scales shall be provided for coal being reclaimed from live and dead storage.
- 1.2.17 Dual systems shall be provided for other required systems such as sampling, etc.
- 1.2.18 Coal-handling system control requirements are as follows:
 - a. Total system control from one location.
 - b. Receiving and reclaiming systems provided with separate controls.
 - c. Normal interlocks included in each system.
 - d. Barge unloader controlled locally.

1.3 Buildings and Other Support Structures

Listed below are the buildings (or structures) that will constitute the support facilities for the main plant. Contractor may change this list to suit his plant design. For the buildings listed, the Contractor shall consider the inclusion of the following facilities:

- a) Toilets and lockers
- b) Lunchroom
- c) Janitorial supply and storage room
- d) Utilities equipment room
- e) Office space or room
- f) Material or equipment storage room
- g) Assembly or conference room(s)

The consideration for selecting which facilities are to be provided in each building include: (a) time and motion efficiency of workers (relative location of building), (b) size of work force, and (c) physical, sociological, and psychological needs of the work force.

- 1.3.1 **Coal Handling Facilities**
 - 1.3.1.1 **Barge Unloading and Crew Building**
 - a. Control booth
 - b. Warming room
 - 1.3.1.2 **Breaker and Sample Building**
 - a. Lab
 - 1.3.1.3 **Hopper Building**
 - 1.3.1.4 **Transfer Stations**
 - 1.3.1.5 **Conveyors**
 - 1.3.1.6 **Coal Washing Building (if required)**
 - a. Electrical control room
 - b. Lab
 - 1.3.1.7 **Coal Silos**
 - 1.3.1.8 **Surge Hopper (at barge unloading facility only)**
 - 1.3.1.9 **Coal Handling Maintenance and Storage Facilities**
 - a. Yard maintenance building (utility building)
 - 1. Equipment repair shop
 - 2. Storage for repair parts
 - 3. Fuel storage
 - b. Storage facilities for conveyor parts
- 1.3.2 **Other Support Facilities**
 - 1.3.2.1 **Access Control Portal (guard house)**
 - a. Entrance area
 - b. Control room
 - 1.3.2.2 **Office Building**
 - a. Reception area
 - b. Administrative offices
 - c. Technician offices, library, and computer rooms
 - d. Field engineers' offices
 - 1.3.2.3 **Service Bay**
 - a. Shops
 - b. Power stores
 - c. Medical unit
 - d. Classrooms

- 1.3.2.4 Lab Building
 - a. Test labs
- 1.3.2.5 Facility for Major Plant Outage or Overhaul Event
 - a. Storage for outage parts
 - b. Assembly and organization of outage work force
- 1.3.2.6 General Yard Storage Area (open)
- 1.3.2.7 Air Separation Building
- 1.3.2.8 Scale House at Weigh Station
- 1.3.2.9 Sulfur Loading Facility
 - a. Operator's shelter
- 1.3.2.10 Intake Pumping Station
- 1.3.2.11 Sewage Treatment Plant
- 1.3.2.12 Fire Protection Equipment Housing
- 1.3.2.13 Fire Protection Deluge Building
- 1.3.2.14 Environmental Data Station
- 1.3.2.15 Visitor Facility
 - a. Display area
 - b. Reception office
- 1.3.2.16 General Site
 - a. Fencing
 - b. Yard lighting
 - c. Planting and seeding
 - d. Driveways and walks
 - e. Parking lot
- 1.3.2.17 Steam Generation Facility
 - a. Boiler
 - b. Control room
 - c. Flue gas cleanup (if required)
 - d. Labs (air and water quality)
- 1.3.2.18 Water Treatment Plant
- 1.3.2.19 Waste Treatment Plant
- 1.3.2.20 Cooling Tower Blowdown
- 1.3.2.21 Potable Water Plant

1.4 Codes, Standards, and Other Design Considerations

Listed below are some of the codes and standards that TVA uses for its plant designs. This list is by no means complete (e.g., piping and pressure vessel codes are not included). It is furnished only as a guide to the contractors in their conceptual design work. It is expected that the contractors will select all appropriate standards and codes for the conceptual design effort and will identify in their final reports those that were used.

1.4.1 Structural and Miscellaneous Steel Material Requirements

1.4.1.1 Materials. Materials should conform to the following specifications:

- | | |
|--|--|
| a. Structural tubing | American Society for Testing and Materials (ASTM) A 501 or A 500, Grade B. |
| b. Structural pipe | ASTM A 53, Grade B, or A 36. |
| c. Structural shapes, plates, and bars (high strength) | ASTM A 441. |
| d. Structural shapes, plates, and bars | ASTM A 36. |
| e. High strength bolts | ASTM A 325 or A 490. |
| f. Erection bolts | ASTM A 307. |
| g. Anchor bolts | ASTM A 307 or ASTM A 36 depending on strength requirements. |
| h. Pipe handrail | ASTM A 120, A 36, or A 53. |
| i. Headed concrete anchors | ASTM A 108. |
| j. Weld rods | American Welding Society (AWS). |
| k. Steel grating (painted or galvanized) | ASTM A 569. |
| l. Expansion anchors | Manufacturer's standard quality. |
| m. Corrosion-resistant bolts | ASTM A 193, Grade B8, or ASTM A 320 B8. |
| n. Corrosion-resistant nuts | ASTM A 194, Grade 8. |

1.4.1.2 Loads. The following loads should be used for the design of the structural and miscellaneous steel components:

a. Live Loads.

- i. Access stairs:
 - a. Main stairs 100 lb/ft².
 - b. Miscellaneous stairs 75 lb/ft².
- ii. Ladders 200 pounds with factor of safety of 4.
- iii. Grating 100 lb/ft² plus equipment.
- iv. Platforms 100 lb/ft² plus equipment.
- v. Hatch covers Applicable floor live load.
- vi. Handrail 200-pound concentrated load at top rail for maximum condition.
- vii. Trash sluice Full of water which has weight of 62.4 lb/ft³.

b. Dead Loads.

- i. Grating (1-1/2-inch steel) 12 lb/ft².
- ii. Steel framing Calculated by design engineer.
- iii. Cable tray supports:
 - a. 18-inch wide tray 48 pounds per linear foot per tray.
 - b. 24-inch wide tray 63 pounds per linear foot per tray.
- iv. Pipes Loads furnished by mechanical piping groups.
- v. Equipment Furnished by the appropriate design groups.
- vi. Ducts Furnished by the appropriate design groups.

- c. Wind Loads. All steel components exposed to normal wind loads should be designed for a basic wind at 30 feet above grade with a 100-year period of recurrence.
- d. Seismic Loads. All structural and miscellaneous steel components should be seismically designed according to the requirements of section 2312 of the Uniform Building Code.

1.4.1.3 Design Specifications and Procedures. All steel components should be designed to the following specifications and procedures:

- a. Part I of the AISC Specification for the Design, Fabrication, and Erection of Structural Steel for Buildings, February 12, 1969, should be used for the steel design as amended through June 12, 1974.
- b. American Concrete Institute Standard: Building Code Requirements for Reinforced Concrete (ACI 318-77), Section 10.14, should be used in the design for concrete bearing stresses.

REFERENCES

AISC Manual of Steel Construction (Seventh Edition).

ACI 318-77, Building Code Requirements for Reinforced Concrete.

International Conference of Building Officials, Uniform Building Code, 1976, Section 2312 (Earthquake Regulations).

1.4.2 Concrete

- 1.4.2.1 Fill Concrete. Where fill concrete is required, it should have a specified compressive strength of 2,000 lb/in.² at the age of 90 days.
- 1.4.2.2 Structural Concrete. All structural concrete not supported on forms, shores, or metal decking should have a specified compressive strength at the age of 90 days, and all structural concrete supported on forms, shores, or metal decking should have a specified compressive strength at the age of 28 days. The weight of concrete should be taken at 145 lb/ft³ in all stability calculations and as 150 lb/ft³ in all structural calculations. The minimum compressive strength required for all structural concrete is 3,000 lb/in.².

- 1.4.2.3 Bearing Stress for Structural Steel Base Plates. Concrete bearing stresses used in the design of structural steel base plates should be limited to values determined for concrete in accordance with ACI 318-77.
- 1.4.2.4 Thermal Considerations. For normal operation or any other long-term period temperatures should not exceed 150°F except for local areas, such as around penetrations, which are allowed to have increased temperatures not to exceed 200°F. For accident or any other short-term period, the temperatures should not exceed 350°F for the interior surface. However, local areas are allowed to reach 650°F from steam or water jets in the event of a pipe failure. Higher temperatures than those given above may be allowed for concrete if tests are provided to evaluate the reduction in strength and this reduction is applied to design allowables. Also, evidence should be provided which verifies that the increased temperatures do not cause deterioration of the concrete either with or without load.
- 1.4.2.5 Reinforcing Steel. Reinforcing steel should be Specification ASTM A 615, Grade 60 deformed bars.
- 1.4.2.6 Construction Codes of Practice. The following codes of practice, with indicated exceptions, will establish the standards of construction procedure:
- ACI 214, "Recommended Practice of Evaluation of Compression Test Results of Field Concrete"
 - ACI 306, "Recommended Practice for Cold-Weather Concreting"
 - ACI 318, "Building Code Requirements for Reinforced Concrete"
 - ACI 347, "Recommended Practice for Concrete Formwork"
 - ACI SP-2, "Manual of Concrete Inspection"
 - ASTM C 94, "Ready-Mixed Concrete"
 - ACI 305, "Recommended Practice for Hot-Weather Concreting"
 - ACI 2.11.1, "Recommended Practice for Selecting Proportions for Normal Weight Concrete"
 - ACI 304, "Recommended Practice for Measuring, Mixing, Transporting, and Placing Concrete"

ACI 315, "Manual of Standard Practice for Detailing Reinforced Concrete Structures"

ASME, "Boiler and Pressure Vessel Code," Section VIII Subsection B, Requirements Pertaining to Methods of Fabrication of Pressure Vessels

Part UW, Requirements for Pressure Vessels Fabricated by Welding

AWS D1.1, "Code for Welding in Building Construction"

ASTM, C 618, Class F, "Specification for Fly Ash for Use as An Admixture in Portland Cement Concrete"

ANSI N45.2.5-74, "Supplementary Quality Assurance Requirements for Installation, Inspection, and Testing of Structural Concrete and Structural Steel During the Construction Phase of Nuclear Power Plants"

REFERENCES

International Conference of Building Officials, Uniform Building Code, 1976, Section 2312 (Earthquake Regulations).

American Concrete Institute Standard: Building Code Requirements for Reinforced Concrete (ACI 318-77 and 1975 supplement).

1.4.3 Electrical Design Considerations

1. NFPA Lightning Protection Code No. 78.
2. NEC (1978) - ANSI C1.
3. NESC (1977) - ANSI C2.
4. Applicable OSHA Standards.
5. IEEE 383-1974 flame test requirements for insulated cable.
6. Use underground electric power distribution within the process area for all power lines above 125 volts.
7. IEEE Standard 141-1976 - "Electrical Power Distribution for Industrial Plants."
8. TVA Electrical Design Guide DG-E2.4.1 Auxiliary Power Systems - Performance Criteria and Application Procedures (applicable portions to be furnished by TVA to Contractor).
9. The medium voltage (4,160 volts, 6,900 volts, or 13,800 volts) system to be low resistance grounded. The low voltage (480/ 277 volts) to be solidly grounded.

10. All motors should have service factor of 115 percent.
11. Standard industry I&C criteria.

1.4.4 Fire Protection and Occupational Safety and Health Requirements

The following items shall be provided:

1. Two 100-percent capacity fire pumps which take suction from a lake or river. The pumps shall be designed in accordance with National Fire Protection Association (NFPA) Standard 20. Motor-operated, self-cleaning strainers with 1/32-inch mesh shall be provided in the discharge line for each pump.
2. A looped yard piping system that is capable of supplying water from the fire pumps to fire hydrants and building fire suppression systems. The fire hydrants shall be located in accordance with NFPA Standard 24. Equipment houses shall be provided at each hydrant.
3. A class III standpipe system in all buildings exceeding one story or 50 feet in height. Hose racks shall be equipped with 100 feet of 1-1/2 inch single jacket lined fire hose and 1-1/2 inch variable fog nozzles (suitable for class A, B, and C fires). The standpipe system shall be designed in accordance with NFPA Standard 14.
4. Fixed water spray or preaction sprinkler systems as dictated by economic and life safety considerations. These systems shall be designed in accordance with NFPA Standards 15 and 16, respectively. Other fixed systems, such as foam or gaseous suppression systems, should be provided for hazards where a water suppression system is not suitable.
5. Portable fire extinguishers located in accordance with NFPA Standard 10. The standard extinguisher shall be a 20-pound cartridge-operated type with a monoammonium phosphate agent and a UL rating of 10A:60B:C. Where dry chemical extinguishers may cause cleanup problems with electrical equipment, a 17 pound stored pressure Halon 1211 extinguisher with a UL rating of 3A:80B:C shall be provided. Wheeled extinguishers shall not be used.
6. Fire emergency equipment room. The room shall be centrally located.
7. Fire detection system. System design shall be in accordance with NFPA Standards 72D and 72E.
8. Two separate intakes for river water shall be provided to ensure that a single component failure will not interrupt the fire water supply system.

The plant shall be designed to comply with the following:

1. 29CFR 1910, Occupational Safety and Health Standards.
2. NFPA Standard 10J, Life Safety Code.

SECTION 2

CHARACTERISTICS OF RAW MATERIALS AND SPECIFICATIONS FOR PLANT OUTPUTS

2.1 Coal Characteristics

The plant will receive coal by barge and truck under a TVA contract that specifies responsibility for coal delivery up to an unloading point at or within the plant boundary. The coal will be a Kentucky No. 9 seam, deep mine, delivered unwashed. The coal properties are described below.

2.1.1 General Information

	<u>Mean</u> ¹	<u>Standard Deviation</u> ²
H.H.V., as Received, Btu/lb	10.980	547.6
Total Moisture, wt %	9.564	1.878
Inherent Moisture, wt %	3.25	0.75
Bulk Density, as Received, lb/ft ³	45.	-
Free Swelling Index ³	(3.0-6.5)	-
Grindability Index	59.	5.117

2.1.2 Proximate Analysis, Dry, wt %

Volatile Matter	37.54	1.878
Fixed Carbon	46.63	1.604
Ash	15.83	3.086
TOTAL	100.00	

2.1.3 Ultimate Analysis, Dry, wt %

Carbon	67.31	2.794
Hydrogen	4.757	.2409
Nitrogen	1.529	.001326
Oxygen	6.343	1.085
Sulfur	4.100	.4858
Ash	15.83	3.086
Chlorine	.1310	.05965
TOTAL	100.000	

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

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

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

2.1.4 Ash Analysis, wt %

SiO ₂	045.94	003.245
Al ₂ O ₃	017.31	001.041
Fe ₂ O ₃	022.29	004.845
CaO	005.486	001.358
MgO	001.164	000.2098
SO ₃	003.264	001.485
Na ₂ O	000.5570	000.2065
K ₂ O	002.336	000.3225
TiO ₂	000.7417	000.1621
Undetermined (by difference)	<u>000.9113</u>	-
TOTAL	100.0000	

2.1.5 Ash Fusion Temperature (°F)

Reducing Atmosphere		
Initial deformation	1,968	29.14
Softening	2,031	50.60
Fluid	2,154	143.3
Oxidizing Atmosphere		
Initial deformation	2,289	32.93
Softening	2,381	33.85
Fluid	2,474	49.72

2.1.6 Forms of Sulfur, % wt in Dry Coal

Sulfate	000.06143	000.03719
Pyritic	002.517	000.4980
Organic	001.528	000.1300
TOTAL	<u>4.10043</u>	

2.1.7 Top Coal Size, as Received: 8"

2.1.8 Sieve Analysis¹

<u>SIZE</u> <u>U.S. SIEVE</u>	<u>WT %</u> <u>LESS THAN</u>
200	2.5
100	3.5
60	5.0
40	6.5
20	10.0
10	17.0
1/8"	22.0
1/4"	33.0
1/2"	48.0
1"	67.0
1-1/2"	78.0
2"	87.0
3"	96.0

¹Analysis based on a 3" x 0" coal sample taken at the exit of a Bradford Breaker. Analysis was performed in association with coal washability tests described in paragraph 2.1.9.

2.1.9 Coal Washability

The following tables provide information relevant to the washability of the design coal. The data is the result of screen analysis (Table 2.1) and float/sink analysis (tables 2.2 to 2.3) of a 3" x 0" sample of Kentucky No. 9 deep mine coal taken at the discharge of a Bradford Breaker. The analysis was performed in April 1976 in support of TVA's study to determine the feasibility of erecting a coal washing plant at the Paradise Steam Plant. While the chemical analysis of the sample coal does not match the design coal precisely (see Table 2.4) the similarity does appear sufficient to permit an assessment of the feasibility of coal washing for improving the performance of the gasification plant.

2.2 Product Gas Specifications (as delivered to the plant battery limits)

Pressure:	600 psig minimum
Temperature:	120°F maximum
Higher Heating Value:	285 Btu/SCF minimum
Total Sulfur:	200 ppm maximum
Total Moisture:	7 lbm/MMSCF maximum
Chemical Composition:	Within the constraints described above, the composition of the gas at the plant fence may be established solely by the coal gasification and gas cleanup processes.

2.3 Byproducts Specifications

Sulfur Byproduct:	Commercial grade, "Bright Sulfur"
Ammonia Byproduct:	Anhydrous commercial grade
Phenol Byproduct:	Phenol commercial grade
Other Byproducts:	Hydrocarbons such as naphtha, light oil and tars require hydrotreating for stability if used as petroleum products.

2.4 Byproduct Disposition (processed to a sellable form with sufficient site storage to accommodate likely shipping interruptions).

Sulfur Byproduct:	Prill form
Ammonia Byproduct:	Loaded from cryogenic storage sphere to barge or tank truck
Phenol Byproduct:	Loaded from API storage tank to tank truck
Ash Byproduct:	Retained in storage pit at site for duration of plant operation, i.e., 20 years
Other Byproducts:	Naphtha, oil, and tar will be stored and loaded from storage to truck.

TABLE 2.1

Paradise Steam Plant
 Peabody No. 9 Deep Mine
 Raw Plant Feed

Lab No. 61-7008		SCREEN ANALYSIS											April 1976			
SIZE		DRY BASIS			Retd on Screen in Col 2			Pass. Screen in Col 1			CUMULATIVE RESULTS					
Retained on	% Wt.	% Ash	% Sul	Btu	% Wt.	% Ash	% Sul	Btu	% Wt.	% Ash	% Sul	Btu	% Wt.	% Ash	% Sul	Btu
3"	Rd x 1-1/2"	Rd	21.9	12.78	4.98	12685.	21.9	12.78	4.98	12685.	100.0	14.93	5.45	12265		
1-1/2"	Rd x 3/8"	RH	37.9	11.63	4.81	12835.	59.8	12.05	4.87	12780.	78.1	15.53	5.58	12147		
3/8"	Rd x 28 Mesh		32.2	18.68	6.62	11589.	92.0	14.37	5.48	12363.	40.2	19.21	6.31	11498.		
28	Mesh x 0		8.0	21.36	5.07	11134.	100.0	14.93	5.45	12265.	8.0	21.36	5.07	11134.		

TABLE 2.2 FLOAT & SINK ANALYSIS

SPECIFIC GRAVITY	CUM. RECOVERY (FLOAT)														CUM. REJECT (SINK)				
	SINK	1	2	3	4	5	6	7	8	9	10	11	12	13	14	% Ash	% Sol.	% Ash	% Sol.
1.40	Composite 3" Rd x 28 Mesh = 92.0% of Raw Plant Feed																		
1.50	3" Rd x 1-1/2" Rd = 21.9% of Raw Plant Feed																		
1.60	1-1/2" Rd x 3/8" Rd = 37.9% of Raw Plant Feed																		
1.70	3/8" Rd x 28 Mesh = 32.2% of Raw Plant Feed																		
1.70	28 Mesh x 0 = 8.0% of Raw Plant Feed																		
-	79.7	7.51	3.30	13553.	79.7	7.51	3.30	13553.	100.0	14.38	5.49	12361.	100.0	14.38	5.49	12361.	14.38	5.49	12361.
1.40	6.0	18.38	6.23	11754.	85.7	8.27	3.51	13427.	20.3	41.34	14.09	7683.	20.3	41.34	14.09	7683.	41.34	14.09	7683.
1.50	2.6	24.97	7.44	10673.	88.3	8.76	3.62	13346.	14.3	50.98	17.37	5975.	14.3	50.98	17.37	5975.	50.98	17.37	5975.
1.60	1.1	25.82	10.60	9764.	89.4	9.02	3.71	13302.	11.7	56.75	19.58	4931.	11.7	56.75	19.58	4931.	56.75	19.58	4931.
1.70	10.6	59.55	20.51	4429.	100.0	14.38	5.49	12361.	10.6	59.55	20.51	4429.	10.6	59.55	20.51	4429.	59.55	20.51	4429.
-	82.3	7.65	3.21	13557.	82.3	7.65	3.21	13557.	100.0	12.78	4.98	12685.	100.0	12.78	4.98	12685.	12.78	4.98	12685.
1.40	6.9	19.60	5.76	11679.	89.2	8.57	3.41	13412.	17.7	36.66	13.19	3631.	17.7	36.66	13.19	3631.	36.66	13.19	3631.
1.50	2.8	29.45	5.93	10027.	92.0	9.21	3.48	13309.	10.8	47.55	17.93	6684.	10.8	47.55	17.93	6684.	47.55	17.93	6684.
1.60	1.4	31.07	11.63	9597.	93.4	9.54	3.60	13253.	8.0	53.89	22.17	5514.	8.0	53.89	22.17	5514.	53.89	22.17	5514.
1.70	6.6	58.73	24.40	4648.	100.0	12.78	4.98	12685.	6.6	58.73	24.40	4648.	6.6	58.73	24.40	4648.	58.73	24.40	4648.
-	85.1	7.56	3.39	13533.	85.1	7.56	3.39	13533.	100.0	11.63	4.81	12835.	100.0	11.63	4.81	12835.	11.63	4.81	12835.
1.40	6.4	18.25	6.71	11754.	91.5	8.31	3.62	13409.	14.9	34.89	12.94	8851.	14.9	34.89	12.94	8851.	34.89	12.94	8851.
1.50	2.5	25.84	8.76	10535.	94.0	8.77	3.76	13332.	8.5	47.41	17.63	6665.	8.5	47.41	17.63	6665.	47.41	17.63	6665.
1.60	.9	31.98	11.35	9454.	94.9	8.99	3.83	13295.	6.0	56.40	21.33	5053.	6.0	56.40	21.33	5053.	56.40	21.33	5053.
1.70	5.1	60.71	23.09	4276.	100.0	11.63	4.81	12835.	5.1	60.71	23.09	4276.	5.1	60.71	23.09	4276.	60.71	23.09	4276.
-	71.5	7.34	3.24	13579.	71.5	7.34	3.24	13579.	100.0	18.68	6.62	11589.	100.0	18.68	6.62	11589.	18.68	6.62	11589.
1.40	5.0	17.44	6.13	11826.	76.5	8.00	3.43	13464.	28.5	47.12	15.12	6597.	28.5	47.12	15.12	6597.	47.12	15.12	6597.
1.50	2.7	20.87	7.15	11280.	79.2	8.44	3.56	13390.	23.5	53.43	17.03	5484.	23.5	53.43	17.03	5484.	53.43	17.03	5484.
1.60	1.1	26.66	8.98	10207.	80.3	8.69	3.63	13346.	20.8	57.66	18.31	4732.	20.8	57.66	18.31	4732.	57.66	18.31	4732.
1.70	19.7	59.39	18.83	4426.	100.0	18.68	6.62	11589.	19.7	59.39	18.83	4426.	19.7	59.39	18.83	4426.	59.39	18.83	4426.
-	74.9	11.58	3.09	12876.	74.9	11.58	3.09	12876.	100.0	21.36	5.07	11134.	100.0	21.36	5.07	11134.	21.36	5.07	11134.
1.60	5.6	24.83	3.60	10769.	80.5	12.50	3.13	12729.	25.1	50.56	10.96	5937.	25.1	50.56	10.96	5937.	50.56	10.96	5937.
1.70	19.5	57.95	13.08	4549.	100.0	21.36	5.07	11134.	19.5	57.95	13.08	4549.	19.5	57.95	13.08	4549.	57.95	13.08	4549.

TABLE 2.3
FLOAT & SINK ANALYSIS

SPECIFIC GRAVITY	CUMULATIVE RECOVERY (FLOAT) % SULFUR										CUMULATIVE REJECT (SINK) % SULFUR									
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18		
Flk	Float	% Wt.	Total	Pyrite	Sulf.	Org.	% Wt.	Total	Pyrite	Sulf.	Org.	% Wt.	Total	Pyrite	Sulf.	Org.	Org.			
Composite 3" Rd x 28 Mesh = 92.08 of Raw Plant Feed																				
-	1.40	79.7	3.30	1.42	.00	1.88	79.7	3.30	1.42	.00	1.88	100.0	5.49	3.67	.01	1.6				
1.40	1.50	6.0	6.27	4.66	.01	1.60	85.7	3.51	1.65	.00	1.86	20.3	14.09	12.53	.07	1.4				
1.50	1.60	2.6	7.44	5.96	.04	1.44	88.3	3.62	1.77	.00	1.05	14.3	17.37	15.83	.10	1.4				
1.60	1.70	1.1	10.60	9.17	.04	1.39	89.4	3.71	1.86	.00	1.85	11.7	19.58	18.02	.11	1.4				
1.70	-	10.6	20.51	18.94	.12	1.45	100.0	5.49	3.67	.01	1.81	10.6	20.51	18.94	.12	1.4				
3" Rd x 1-1/2" Rd = 21.9% of Raw Plant Feed																				
-	1.40	82.3	3.21	1.41	.00	1.00	82.3	3.21	1.41	.00	1.00	100.0	4.98	3.12	.02	1.6				
1.40	1.50	6.8	5.76	4.24	.02	1.50	89.2	3.41	1.63	.00	1.79	17.7	13.19	11.06	.08	2.0				
1.50	1.60	2.8	5.83	4.63	.03	1.17	92.0	3.48	1.72	.00	1.76	10.8	17.93	15.42	.13	2.3				
1.60	1.70	1.4	11.63	10.29	.02	1.32	93.4	3.60	1.85	.00	1.75	8.0	22.17	19.20	.16	2.6				
1.70	-	6.6	24.40	21.09	.19	3.12	100.0	4.98	3.12	.02	1.84	6.6	24.40	21.09	.19	3.1				
1-1/2" Rd x 3/8" Rd = 37.9% of Raw Plant Feed																				
-	1.40	85.1	3.39	1.45	.00	1.94	85.1	3.39	1.45	.00	1.94	100.0	4.81	2.95	.01	1.6				
1.40	1.50	6.4	6.71	5.07	.00	1.64	91.5	3.62	1.70	.00	1.92	14.9	12.94	11.52	.07	1.6				
1.50	1.60	2.5	8.76	7.15	.04	1.57	94.0	3.76	1.85	.00	1.91	8.5	17.63	16.37	.13	1.6				
1.60	1.70	.9	11.35	9.89	.05	1.41	94.9	3.83	1.92	.00	1.91	6.0	21.33	20.22	.16	.6				
1.70	-	5.1	23.09	22.04	.10	.87	100.0	4.81	2.95	.01	1.85	5.1	23.09	22.04	.18	.6				
3/8" Rd x 28 Mesh = 32.2% of Raw Plant Feed																				
-	1.40	71.5	3.24	1.37	.01	1.86	71.5	3.24	1.37	.01	1.86	100.0	6.62	4.88	.03	1.71				
1.40	1.50	5.0	6.10	4.42	.03	1.65	76.5	3.43	1.57	.01	1.85	28.5	15.11	13.69	.07	1.31				
1.50	1.60	2.7	7.15	5.61	.04	1.50	79.2	3.55	1.71	.01	1.83	23.5	17.03	15.67	.08	1.21				
1.60	1.70	1.1	8.98	7.52	.04	1.42	80.3	3.63	1.79	.01	1.83	20.8	18.31	16.97	.09	1.25				
1.70	-	19.7	18.83	17.50	.09	1.24	100.0	6.62	4.88	.03	1.71	19.7	18.83	17.50	.09	1.24				
28 Mesh x 0 = 8.01 of Raw Plant Feed																				
-	1.60	74.9	3.09	1.38	.08	1.63	74.9	3.09	1.38	.08	1.63	100.0	5.07	3.53	.11	1.				
1.60	1.70	5.6	3.60	2.49	.16	.95	80.5	3.13	1.46	.09	1.58	25.1	10.96	9.94	.19	.1				
1.70	-	19.5	13.08	12.08	.20	.80	100.0	5.07	3.53	.11	1.43	19.5	13.08	12.08	.20	.1				

Table 2.4. ANALYSIS OF COAL SAMPLE USED IN WASHABILITY TESTS

Composite 1.70 float from 3" Rd x 0ULTIMATE ANALYSIS

	<u>Dry Basis % Weight</u>
Carbon	72.99
Hydrogen	4.95
Nitrogen	1.37
Chlorine	0.12
Sulfur	3.70
Ash	9.59
Oxygen (diff.)	7.28
	<u>100.00</u>

DRY PROXIMATE

%Ash	9.59
%Volatile	40.40
%Fixed Carbon	50.01
	<u>100.00</u>
Btu	13203.
%Sulfur	3.70

FUSION TEMPERATURE OF ASHReducing

Initial Deformation	2020°F	
Softening (H = W)	2100°F	H is Cone Height
Softening (H = 1/2W)	2140°F	W is Cone Width
Fluid	2260°F	

Hardgrove Grindability Index 56

MINERAL ANALYSIS OF ASH

	<u>Ignited Basis % Weight</u>
Phos, pentoxida, P ₂ O ₅	0.10
Silica, SiO ₂	47.96
Ferric oxide, Fe ₂ O ₃	25.59
Alumina, Al ₂ O ₃	17.51
Titania, TiO ₂	0.89
Lime, CaO	1.00
Magnesia, MgO	0.95
Sulfur trioxide, SO ₃	0.71
Potassium oxide, K ₂ O	2.19
Sodium oxide, Na ₂ O	1.60
Undetermined	1.50
	<u>100.00</u>

SECTION 3

SITE INFORMATION

3.1 Introduction

Murphy Hill is a rural site in northeast Alabama that has been named by TVA as the preferred location for the coal gasification demonstration plant. While final decision on site selection must await environmental review proceedings, Murphy Hill shall be assumed as the plant site for the purpose of conceptual design.

Murphy Hill is located along the Tennessee River about 12 miles northeast of Guntersville, and about 30 miles southeast of Huntsville, Alabama. The site is approximately 600 feet above mean sea level (MSL) and has no existing foundations or obstruction. The site is accessible only by barge and road.

The public road within the site boundaries may be relocated. However, any new location must continue to permit public access through the site area between the northeast and southwest boundaries.

The geographical location of the shoreline may not be altered. However, dredging and the installation of pilings, caissons, or mooring cells (as required for barge handling) are acceptable shoreline modifications. Dredging material may be used as site fill.

Topographical and other types of maps of the site and its vicinity include:

1. Columbus City quadrangle, Alabama-Marshall Co., 7.5 minute series, (topographic) 90-NW, photo revised 1970.
2. Site topo map depicting site boundaries, reproducible (unnumbered)
3. Site topographic map depicting site boundaries, SK-GCW 22080. This drawing is a duplicate of the reproducible in 2 above, revised to show a public use area.
4. Navigation chart No. 502, Tennessee River, Guntersville Lake, mile 360.0 to mile 371.2.
5. Detailed topographical maps (sepias) 461N801-101, 102, 103, 104, 105, 106, 108, 109, 113, and 114.
6. Flowage topography maps, Guntersville Reservoir, Dwg. Nos. 97.0.1 and 97.0.0.5, Rev. 0.
7. Land Map - 421K700

Information related to meteorological, river water, and other site-specific conditions is provided in the following sections.

3.2 Annual Extreme and Average Site Conditions

Maximum design wet bulb temperature	80°F
Yearly average wet bulb temperature	55°F
Max/Min design dry bulb temperature	97°/14°F
Yearly average dry bulb temperature	60°F
Design point for freeze protection (winterizing)	9°F
Design frost line, feet below grade	2'-6"
Maximum rainfall in a 12-hour period ¹	5 in.
Maximum rainfall in a 24-hour period ¹	6 in.
Average yearly rainfall	54 in.
Maximum wind speed design at 100 year recurrence	75 mph
Prevailing wind direction/speed in summer, mph	S/9
Prevailing wind direction/speed in winter, mph	N/9
Water evaporative rate, inches/year	37
Maximum/minimum river temperature °F	86/40
Maximum/minimum normal river elevation, feet above MSL	595/593
500-year flood elevation ²	597.5
Maximum/minimum river flow rate (winter/summer respectively), ft ³ /sec	44,200/27,100
Yearly average river flow rate, ft ³ /sec	36,000
Channel velocity: (a) 1-year flood	2 ft/sec
(b) 5-year flood	2.5 ft/sec
(c) 40-year flood	3 ft/sec
Maximum winter/summer river velocity (50 ft from shore) ft/sec	2/1
Normal atmospheric pressure	14.4 psi @ 600' above MSL

¹With a probable recurrence interval of once in ten years.

²For design purposes, the river water intake facility, deck elevation, and all coal handling equipment shall be above elevation 598; all main plant structures shall be above elevation 606.

3.3 Monthly Average Site Conditions

	JAN	FEB	MAR	APRIL	MAY	JUNE	JULY	AUG	SEPT	OCT	NOV	DEC
Dry Bulb Temperature °F	40	42	50	64	69	76	78	77	72	62	50	45
Wet Bulb Temperature °F	37	37	44	56	62	69	72	71	66	56	45	41
Dewpoint °F	31	30	37	49	57	64	74	67	63	51	40	37
Relative Humidity %	73	68	64	63	68	69	68	75	75	71	72	75
Rainfall, inches	N/A	4.5	8.1	3.5	N/A	4.8	5.0	3.7	4.0	4.2	3.6	5.4
Water Temperature* °F	41	44	55	60	68	77	83	83	81	74	62	51
Wind Speed, mph	9	9	9	9	8	7	6	5	6	7	8	9

33

*10 ft below surface

3.4 Percent Occurrence of Various Wind Speed Classes (Yearly Average)

Wind Speed Class (mph)	0-3	4-7	8-12	13-18	19-24	>24
	34.5	28.1	24.1	11.8	1.2	0.2

3.5 Percent Occurrence of Wind Directions (Yearly Average)

Wind Direction	N	NE	E	SE	S	SW	W	NW
	10.3	7.5	2.3	3.4	16.7	4.8	4.2	5.1

3.6 River Water Analysis (in Milligrams/per Liter)

	<u>Median</u>	<u>Maximum</u>	<u>Minimum</u>
Silica (SiO ₂)	5	6	3
Calcium (Ca)	19	23	15
Magnesium (Mg)	3.8	4.8	2.0
Sodium (Na)	5.3	24	1
Bicarbonate (HCO ₃)	50	62	38
Sulfate (SO ₄)	9.9	16	6.3
Chloride (Cl)	8	31	3
Nitrate (NO ₃)	1.3	2.8	0
Dissolved Solids (180°C)	84	160	56
Hardness, as CaCO ₃	62	76	49
pH (SU)	7.4	7.9	6.9
Color (PCU)	5	20	0
Iron	negligible		
Fluorides	negligible		

SECTION 4
ENVIRONMENTAL CONTROL GUIDELINES

4.1 Air Emissions

In the absence of existing environmental regulations governing specifically air emissions from coal gasification plants, Contractor shall recommend to TVA the use of any applicable Federally proposed standards or emission levels consistent with the use of the best practical control technology.

If an auxiliary boiler is to be used, it must meet New Source Performance Standards for electric utility steam generating units promulgated by EPA on June 11, 1979 (44 FR 33580, 40 CFR Part 60). If a Claus sulfur recovery plant is to be used, it must meet standards promulgated on March 8, 1974 (39 FR 9308, 40 CFR Part 60) under subpart J: "Petroleum Refineries."

4.2 Liquid Effluents

It shall be assumed that the gasification plant will be subject to effluent guidelines and standards similar to those promulgated for the steam-electric power generating and the petroleum refinery point source categories:

1. Steam-Electric Power Generating Point Source Category - 40 CFR 423.15, Standards of Performance for New Sources
2. Petroleum Refinery Point Source Category - 40 CFR 419.15, Standards of Performance for New Sources

In addition, the following effluent guidelines shall apply. The pH shall be no less than 6.0 and no greater than 9.0 standard units. There shall be no discharge of floating solids or visible foam in other than trace amounts. Instantaneous total suspended solids in effluent discharge shall not exceed 50 mg/l. This suspended solids limit is applicable to any flow up to the flow resulting from a 24-hour rainfall with a probable recurrence interval of once in ten years. If an impoundment is utilized, it shall be capable of containing a 10-year, 24-hour rainfall event.

4.3 Solids and Liquids Impoundment

For the base-case design, all ash, sludge, and water containment ponds shall be unlined. For the off design investigations, the containment ponds shall be lined with an impermeable blanket (e.g., compacted clay liners).

SECTION 5

BASIS AND ASSUMPTIONS FOR ECONOMIC ASSESSMENTS

- 5.1 Module life: 20 years after startup (as defined by present schedule) with no salvage value. No cost assumed for disassembly and disposal.
- 5.2 Module service of stream factor: 90 percent at 100 percent design rating.
- 5.3 Capital and operating cost estimates: Stated in January 1980 dollars.
- 5.4 Financing: Government appropriations (disregard income and property tax).
- 5.5 Economic evaluation rate: 12 percent.
- 5.6 General and administrative expense: 5 percent of total operating and maintenance cost.
- 5.7 Schedule of cash flow by year: Breakdown of plant subsystems by both capital and operating expenses.
- 5.8 Product gas cost: To be presented as a levelized unit cost (\$/MMBtu) over economic life of plant.
- 5.9 Cost of delivered coal: 1980 Value, \$/MMBtu \$1.25.
Escalate each year to the end of 1985 by 9 percent.
Escalate each year from 1986 to end of 1995 by 8 percent.
Escalate each year from 1996 onward by 7 percent.
- 5.10 Limestone* cost: \$13/ton as received (escalate at same rate as coal).
- 5.11 Coke (for gasifier starter) cost*: \$60/ton, sized (escalate at same rate as coal).
- 5.12 Byproduct credit: No credit for byproduct except for excess electricity and coal fines (see values below). See Table 5.8 for byproduct credit for sensitivity analysis case.
- 5.13 Excess electricity sales: Same as cost to plant.
- 5.14 Coal fine sales: 80 percent of ROM coal cost.
- 5.15 Land cost*: \$3,000/acre.
- 5.16 Clearing and grubbing*: \$2,000/acre.
- 5.17 Excavation*: (a) earth \$1.50/cubic yard, (b) rock \$10/cubic yard.
- 5.18 Fill (compacted)*: \$3/cubic yard.
- 5.19 Electric power cost: See Table 5.1.

*1980 Cost, unless stated otherwise, escalate according to table 5.6.

5.20 Construction labor: See Table 5.2.

5.21 Operating labor: See Table 5.3.

5.22 Maintenance labor: See Table 5.4.

5.23 Escalation rates: See Table 5.6.

5.24 Sensitivity analysis: See Table 5.7.

5.25 Byproduct values for sensitivity analysis: See Table 5.8.

Table 5.1. ELECTRIC POWER COST

<u>Year</u>	<u>Energy Charge Mills per kWh</u>	<u>Demand Charge \$/kW per month¹</u>
1980	17.47	4.80
1981	20.59	5.85
1982	21.82	6.31
1983	22.92	6.96
1984	24.25	7.53
1985	26.05	7.94
1986	26.14	8.81
1987	26.32	9.84
1988	28.39	9.96
1989	27.40 ²	11.73
1990 ³	29.17 ²	12.92

¹Peak for each respective month.

²Projected lower values for 1989 and 1990 reflect the manner in which the TVA power system expansion is planned.

³The cost of power and demand charges for years beyond 1990 shall be assumed to escalate at a rate of 7.76 percent per year.

Table 5.2. CONSTRUCTION LABOR COST¹

<u>Craft</u>	<u>Rate/Hour</u>
Boilermakers	\$16.00
Bricklayers	12.00
Cement Masons	11.00
Millwrights	12.50
Carpenters	11.50
Painters	12.50
Electricians	14.00
Ironworkers	13.50
Machinists	11.00
Outside Machinist	12.00
Sheetmetal	14.50
Steamfitters	14.50
Teamsters	9.00
Laborers	8.00
Oper Engr	11.00

¹1980 construction trades and labor wages and fringe benefits. (Composite of foremen and journeymen). Θάρσο θοξίνφεφ το α δηώσε οθ ηρωκ φουωυθ

Table 5.3. OPERATING LABOR¹

	<u>Annual Salary Rate, \$</u>
<u>1. Plant Superintendent's Office</u>	
Plant Superintendent	40,800
Assistant Plant Superintendent	37,400
Administrative Services Superintendent	28,000
Administrative Officer	19,200
Stores Records Clerk	14,100
Payroll Clerk	15,200
Clerk	12,700
Clerk-Typist	10,400
Safety Engineer	15,800
<u>2. Plant Operations Section</u>	
Plant Operating Supervisor	34,500
Clerk-Stenographer	12,700
<u>2A. Plant Operations</u>	
Shift Engineer	27,600
Assistant Shift Engineer	22,600
Senior Switchboard Operator	20,300
Unit Operator	20,300
Assistant Unit Operator	17,000
Auxiliary Operator	15,300
Student Instructor	26,900
Student General Plant Operator	14,200
<u>2B. Yard Operations</u>	
Yard Operations Supervisor	24,000
Coal Hauling Foreman	14.00*
HEO Pr	13.00*
HEO Pr Apprentice	11.00*
Coal Tower Foreman	14.00*
Coal Car Dump Operator	14.00*
Track Foreman	14.00*
Laborer	8.50*
Student Plant Laborer	7.50*

¹Rates and structure based on representative TVA coal-fired power plant (see Table 5.5). Values listed are 1980 rates. Annual rates are convertible to hourly rates by dividing the annual rates by 2,080 (52 weeks times 40 hours). For total annual labor cost add 42 percent for fringe benefits.

*1980 hourly wages and fringe benefit rates for trades and labor.

Table 5.3. (continued)

<u>3. Plant Results Section</u>	<u>Annual Salary Rate, \$</u>
Plant Results Supervisor	34,500
Assistant Plant Results Supervisor	28,000
Instrument Unit Foreman	21,600
Instrument Mechanic	14,50*
Instrument Mechanic Apprentice	11.00*
Mechanical Unit Foreman	21,600
Engineering Aide	16,200
Chemical Unit Foreman	21,600
Chemical Lab Analyst	16,200
Materials Tester	16,200
<u>4. Miscellaneous</u>	
Boilermaker Foreman	22,700
Boilermaker	19,200
Boilermaker Apprentice	15,873
Janitor (Senior)	14,665
Janitor	13,500

*1980 hourly wages and fringe benefit rates for trades and labor.

Table 5.3. (continued)

<u>Classification</u>	<u>Annual Salary¹</u>	<u>Classification</u>	<u>Annual Salary¹</u>
<u>Class A Operator</u>	\$17,900	<u>Laborer (Unclassified)</u>	
Air Separation Control Operator		<u>Plant Laborer</u>	\$13,100
Ammonia Plant Control Operator		Ammonia Laborer	
Ammonia Plant Process Operator		Bagging and Loading Laborer	
Ammonia Storage and Distribution Operator		Pilot Plant Laborer	
Boiler House Operator		Pilot Plant Operator Trainee	
Gasification-Purification Control Operator		Relief Operating Laborer	
Granulator System Operator		Water Plant Laborer	
Liquid Fertilizer Unit Operator		<u>Class C Operator</u>	\$15,200
Neutralizer and Concentrator Operator		Bagger and Weigher	
Nitric Acid Operator		Conveyor Operator	
Pilot Plant Operator, A		Fertilizer Loader	
Relief Operator, A		Pilot Plant Operator, C	
Urea Unit Control Operator		Relief Operator, C	
Urea Unit Process Operator		<u>Class B Operator</u>	\$16,400
Water Plant Operator		Acid Pumper and Adjuster	
<u>Foreman</u>	\$21,200	Air Separation Auxiliary Operator	
Acid Unit Foreman		Ammonia Plant Auxiliary Operator	
Ammonia From Coal Foreman		Boiler House Auxiliary Operator	
Bagging and Loading Foreman		Coal Handling and Utility Operator	
Pilot Plant Shift Foreman		Gas Purification and Sulphur Recovery Auxiliary Operator	
Urea Unit Foreman		Granulator System Auxiliary Operator	
Utilities Foreman		Liquid Fertilizer Unit Auxiliary Operator	
Chemical Plant Foreman	\$25,500	Loading Checker	
<u>Operator Trainee</u>		Nitric Acid Auxiliary Operator	
Chemical Plant Operator		Overhead Crane Operator	
Trainee I--A	\$13,500	Pilot Plant Operator, B	
Chemical Plant Operator		Relief Operator, B	
Trainee I--B	\$13,900	Slurry-Preparation-Gasification Auxiliary Operator	
Chemical Plant Operator		Storage and Loading Operator	
Trainee II	\$15,200	Waste Water Treatment Auxiliary Operator	
Chemical Plant Operator			
Trainee III	\$16,400		

¹TVA schedule of trades and labor classes and rates of pay schedule C. Regular operating work--Division of Chemical Operations and Division of Chemical Development. Values listed are 1980 rates. Annual rates are convertible to hourly rates by dividing the annual rates by 2,080 (52 weeks times 40 hours). For total annual labor cost add 42 percent for fringe benefits.

Table 5.4 MAINTENANCE LABOR

<u>Classification</u>	<u>Plant Maintenance</u>	<u>Turnaround Maintenance</u>
<u>Supervision¹</u>		
Electrical and Mechanical Maintenance Supervisors	34,500	
Assistant Electrical or Mechanical Maintenance Supervisors	28,000	
Mechanical Engineers	15,800	
Engineering Aide	13,800	
<u>Foremen²</u>		
Abestos	16.50	20.50
Boilermaker	17.50	21.50
Electrician	15.50	19.00
Ironworker	15.00	18.00
Machinist	13.50	17.00
Sheetmetal Worker	16.00	19.50
Steamfitter	16.00	19.50
Carpenter	13.00	16.00
Painter	13.00	16.00
Truck Driver	10.50	13.00
Laborer	10.00	12.00
<u>Journeyman</u>		
Asbestos Worker	15.50	18.50
Boilermaker	16.00	19.50
Electrician	14.50	17.50
Ironworker	14.00	16.50
Machinist	12.50	15.50
Sheetmetal Worker	15.00	18.00
Steamfitter	15.00	18.00
Carpenter	12.00	15.00
Painter	12.00	15.00
Truck Driver	9.50	11.50
Laborer	8.50	10.00

¹Rates and structure based on representative TVA coal-fired power plant (see Table 5.6). Values listed are 1980 rates. Annual rates are convertible to hourly rates by dividing the annual rates by 2,080 (52 weeks times 40 hours). For total annual labor cost add 42 percent for fringe benefits.

²Schedule of hourly trades and labor rates; 1980 rates. Rates rounded to a whole or half-dollar. Values include fringe benefits.

TABLE 5.5 TVA'S JOHN SEVIER COAL-FIRED POWER PLANT
OPERATIONS SECTIONS

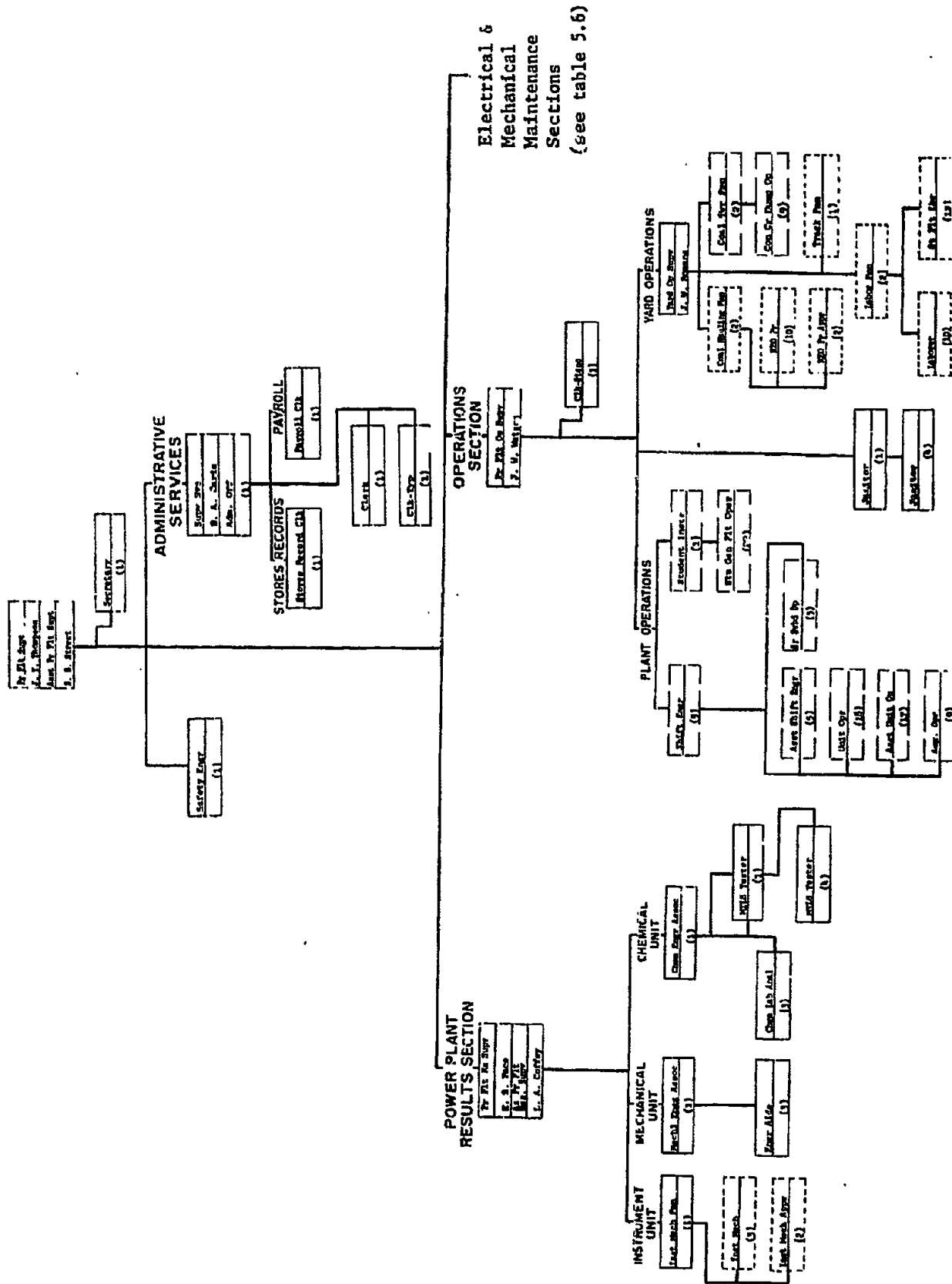


TABLE 5.6

TVA'S JOHN SEVIER COAL-FIRED POWER PLANT
MAINTENANCE SECTIONS

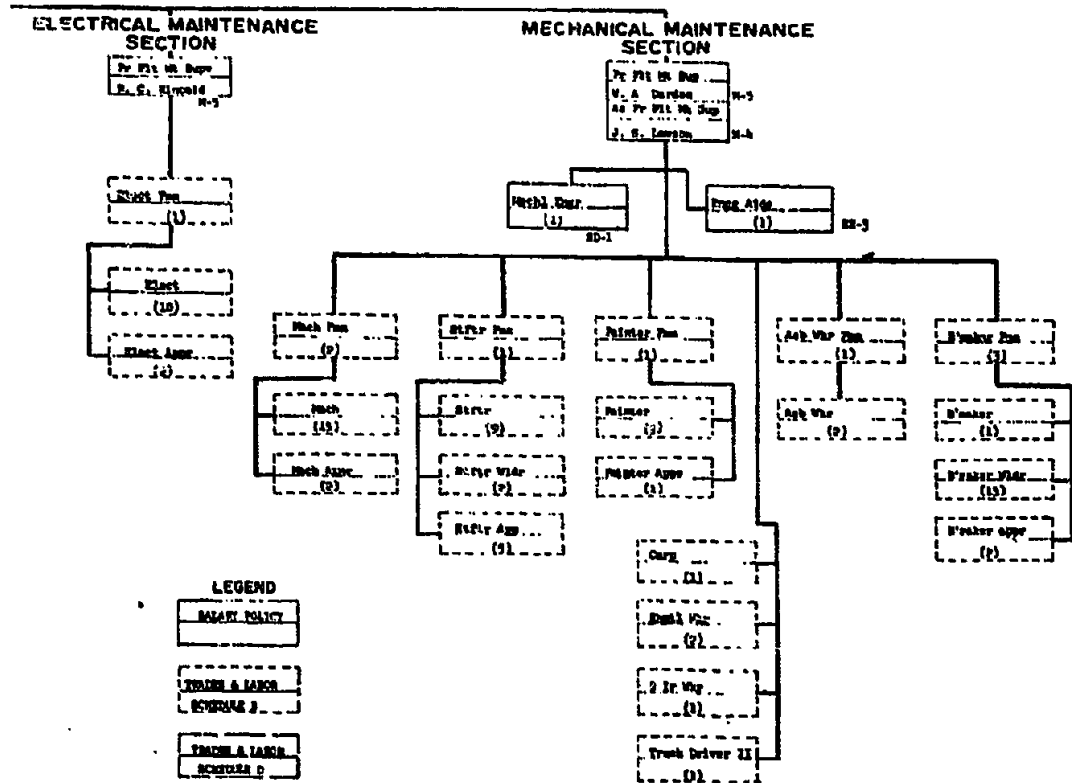


Table 5.7. ESCALATION RATES

CONSTRUCTION: T&L Wages and Fringe Benefits

<u>Rates On</u>		<u>Higher Than</u>
1/1/81	are estimated to be 8.5%	1/1/80
1/1/82 thru 1/1/86	are estimated to be 9.0%	Each previous January 1
January 1 each year	are estimated to be 8.0%	Previous January 1

MAINTENANCE: Annual T&L Salaries and Fringe Benefits

<u>Salaries On</u>		<u>Higher Than</u>
1/1/81	are estimated to be 9.0%	1/1/80
1/1/82 thru 1/1/86	are estimated to be 8.5%	Each previous January 1
January 1 each year	are estimated to be 7.5%	Previous January 1

PLANT OPERATORS: Annual Salary Policy Salaries and Fringe Benefits

<u>Salaries On</u>		<u>Higher Than</u>
7/1/80 thru 7/1/86	are estimated to be 8.0%	Each previous July 1
July 1 each year	are estimated to be 7.5%	Previous July 1

CONSTRUCTION: Materials and Equipment*

<u>Prices On</u>		<u>Higher Than</u>
1/1/81	are estimated to be 10%	1/1/80
1/1/82 thru 1/1/86	are estimated to be 9.5%	Each previous January 1
January 1 each year	are estimated to be 8.0%	Previous January 1

*Contractor should recommend to TVA the use of higher escalation rates for materials that have historically been subject to abnormally high price increases

Table 5.8. SENSITIVITY ANALYSIS APPLIED TO COST OF GAS

	<u>Increment</u>
1. Coal Cost	+ 50%
2. Capital Cost Variation ¹	+ 25%
3. Operating Costs	+ 50%
4. Service Factors (Base Case = 90%)	80%, 70%, 60%
5. Byproduct Value	See table 5.9
6. Design/Construction Period per Module	± 1 year
7. Operating Life, Years	+ 5, + 10
8. Sulfur in Product Gas ²	To 1.0 ppm
9. Product Gas Pressure	Max = 800 psi Min = 200 psi ³
10. Economic Evaluation Factor	T.B.D.

¹Contractor may recommend alternate increment and suggest a list of equipment for which contingencies are to apply.

²Contractor is to use factored estimates for determining gas cost at sulfur level that fall below the capability of the design sulfur removal system.

³Or lowest practical value above 200 psi permitted by design constraints (contractor to recommend value).

Table 5.9. BYPRODUCT VALUES FOR SENSITIVITY ANALYSIS¹

Sulfur, \$/Ton	70.00
Sulfuric Acid, \$/Ton	60.00
Ammonia (Anhydrous), \$/Ton	130.00
Naphtha (120-320°F), \$/Gal	0.80
Light Oil (300-700°F), \$/Gal	0.80
Tar (700°F+), \$/Gal	0.60
Phenols, \$/Gal	0.75
Coal Fines, \$/Ton	80% of ROM coal cost
Export Power, ¢/kWh	Same as cost to plant
Methanol, ¢/Gal	35

¹Except for coal fines and electric power, escalate byproduct values at same rate as coal prices.



2.3 FOSTER WHEELER SUPPLEMENT



SECTION 2.3

TVA COAL GASIFICATION STUDY

FOSTER WHEELER SUPPLEMENT TO TVA DESIGN CRITERIA¹

1. General

- o Conceptual designs for coal gasification plants using the following gasifiers are to be prepared by Foster Wheeler:
 - Lurgi dry ash gasifier
 - B & W entrained flow gasifier
 - K-T gasifier
 - Texaco gasifier

- o Each of the above plants will be designed for gasification of 20,000 tons per day of Kentucky #9 bituminous coal. The plant will contain four modules with 5000 tons per day of coal gasified in each module.

- o Limited evaluation information will be prepared on a plant based on the BGC/Lurgi slagging gasifier.

2. Process Constants

Standard Conditions

Standard Pressure	=	14.7 psia
Standard Temperature	=	60°F = 520.°R

¹TVA Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant dated March, 1980.

Report No. EW-17



Atomic Weights

H	1.008	N	14.01
C	12.01	S	32.07
O	16.00	A	39.94
		Cl	35.46

Dry Air Composition

O ₂	20.93 vol. %	
N ₂	78.11 vol. %	(includes Ne + He)
A	0.93 vol. %	
CO ₂	0.03 vol. %	

100.00 MW = 28.82

0.02 moles H₂O vapor/mole dry air

Gas Law Constant

R = 10.734 $\frac{(\text{psia} \text{ (ft}^3\text{)})}{\text{°R}}$

Heat Of Combustion of Gases

	<u>Btu/SCF</u>	
	<u>GROSS</u>	<u>NET</u>
H ₂	325	275
CO	321	321
CH ₄	1012	911
C ₂ H ₄	1609	1508
C ₂ H ₆	1784	1631
C ₃ H ₈	2563	2358
C ₃ H ₆	2370	2217
C ₄ H ₁₀	3373	3314
C ₅ H ₁₂	3914	3624
NH ₃	441	364
H ₂ S	646	595

Other hydrocarbons - use API Technical Data Book

Heat of Combustion of Solids (on a moisture free basis)

$$(2) \text{ GHV Btu/lb} = 14,658C + 56,878H + 2,940S - 658\text{Ash} - 5153$$

(O + H)

$$\text{LHV Btu/lb} = \text{GHV} - 9,472 \left(H - \frac{O}{8} \right)$$

C, H, O, S, N, Ash, in weight fraction of elements
and Ash

3. Coal Properties

- o Average properties of coal will be used for conceptual design.
- o Allowance will be made for variations in coal properties by selected over-design factors for individual plant sections.

4. Water Temperatures

- o Maximum temperature for heat exchange of untreated water is 105°F.
- o Design temperature for water from cooling towers is 88°F.
- o Maximum temperature for cooling water from heat exchangers is 120°F.
- o Design river water temperature is 88°F.

5. Plant parameters

- o Module Capacity
Each module will gasify 5000 tons per day of coal as fed to the gasifiers.
- o Plant Capacity
Each plant will consist of four identical and independent modules gasifying a total of 20,000 tons per day of coal

(2) IGT Coal Technical Data Book



as fed to the gasifiers.

o Plant Battery Limits Definition

The plant battery limits includes all roads, electrical substation, barge, loading and unloading facilities, plant processing units and support systems including storage, solid waste disposal area, cooling towers, settling ponds, water and sewage treatment facilities, fire protection systems, and buildings.

o Exclusions

The plant does not include product gas odorization and product gas distribution system.

o Plant and Module Operation

The plant will be designed to operate at constant output of fuel gas product at design rate, 24 hours per day, 7 days per week. Each module will be designed for a service factor of 90% on a yearly basis. No provision will be made in the conceptual plant designs to maintain design product gas output during module or plant outages.

o Plant Sections (as required for each gasification process)

<u>Section No.</u>	<u>Title</u>
100	Coal Preparation
200	Air Separation
300	Coal Gasification (coal feed, gasification, gas cooling and scrubbing, gas liquor preparation, ash removal from gasifiers)
400	Acid Gas Removal
500	Gas Compression
600	Sulfur Recovery (Claus plant, tail gas cleanup, prilling)



<u>Section No.</u>	<u>Title</u>
700	Sour Water Stripping
800	Ash/Slag Handling
900	Phenol Recovery
1000	Ammonia Recovery
1100	Coal Handling (receiving, storage, transfer) Support Systems
1200	Utility Area Raw Water Storage and Treatment Potable Water Treatment BFW and Condensate Treatment Steam Generation Plant and Instrument Air and Inert Gas
1300	Cooling Water System
1400	Flare and Incinerator
1500	Wastewater Treatment
2000	General Facilities Long Term Solid Waste Storage By-products and Chemicals Storage Power, Lighting, Communications Roads and Fences Firewater System Sewage Plant Interconnecting Piping
2100	Buildings
2200	Dock Facilities

O Equipment Numbering System

To the extent required for conceptual designs, the following system of numbering equipment will be used:

- equipment letter prefix - Table
- gasification process

Dry ash	No. 21
B&W	No. 27
K-T	No. 32
Texaco	No. 35



- following the prefix designating the coal gasification process, a two-digit number will designate the module number and number of trains in a module.
- The above two-digit number will be followed by the equipment letter designation and the equipment number.
- letters A, B, . . . following the equipment number designates the operating and spare equipment.

Example: Consider an air compressor which is the second of two part-capacity machines in the second of two air separation units in the first module of a plant based on Lurgi dry ash gasifiers:

Equipment No. 21-12-C102 B



TABLE S-I
CATEGORY OF EQUIPMENT
LETTER PREFIXES

The following tabulation presents the standard letter prefixes to be used for the various categories of equipment noted when assigning item numbers to major pieces of equipment.

<u>Letter Prefix</u>	<u>Category of Equipment</u>
AG	Agglomeration Equipment, Briquetting Pelletizers, Pelletizers
B	Blowers, Exhausters, Fans
BL	Blenders, Coaters, Homogenizers, Kneaders
BN	Bins, Hoppers
C	Compressors (all), Vacuum Pumps
CL	Clarifiers, Thickeners
CR	Conveyors, Bucket Elevators, Feeders
CT	Cooling Towers
D	Drums, Accumulators, Kettles (Vats), Pots, Receivers
SH	Deaerating Heaters
DR	Dryers
DS	Desuperheaters
E	Exchangers, Aftercoolers, Air Coolers, Condensers, Kettles (Heat Exchange), Preheaters, Reboilers, Waste Heat Exchangers
EG	Electric Generators
EJ	Expansion Joints
EL	Elevators (Personnel, Freight)
F	Filters (all types)
FL	Flares
H	Heaters, Furnaces, Ovens, Reboilers, Retorts
J	Ejectors, Eductors, Injectors, Spargers
K	Kilns (all)



TABLE S-I
CATEGORY OF EQUIPMENT
LETTER PREFIXES

<u>Letter Prefix</u>	<u>Category of Equipment</u>
M	Mixers, Agitators
MCC	Motor Control Centers
P	Pumps (all)
PK	Packaging Equipment, Bag Fillers, Fillers, Flatteners, Packagers, Pelletizers
PG	Package Units, Air Separation Units, Chemical Additive Units, Desalters, Hydrogen Generating Units, Inert Gas Generators, Refrigeration Units, Water Treating Units
R	Reactors, Regenerators
RD	Rotating Disc Contactor Drives, Special Drives
S	Separators (all)
SC	Sample Coolers
SG	Sample Generators, Boilers
SL	Silencers
ST	Stacks, Chimneys
SW	Switchgear (Electric)
T	Towers, Absorbers, Distillation Columns
TK	Tanks, Gas Holders, Storage Tanks, Vats
TR	Transformers



6. Storage Parameters

Coal Storage

- as received, live 14 days
- prepared, enclosed live 56 hours
- as received, dead 90 days
- fines, enclosed 5 days

- Ground Limestone 14 days
- Sulfur 30 days
- Ammonia 30 days
- Phenol 10 days
- Naphtha, Oil, Tar 10 days

7. Capacity Definitions for Module Sections

7.1 General

- o The number of trains for any module section shall be selected to achieve 90% on-stream factor for each module.
- o The design capacity of any module section (expressed as a percentage of the normal material balance) shall be selected to provide for variation in coal properties. For this purpose, 1.3 standard deviations from average Coal properties will be assumed to cover 90% of coal property variations.



7.2 Coal Preparation

Design Capacity	300% of normal material balance for the 4 module plant
Number of Trains	2 - 50% trains 2 crushers per train 2 pulverizers per train (if specified by FW)

7.3 Air Separation

Design Capacity	105% of normal material balance
Number of Trains	2 - 50% trains

Provide liquid oxygen storage and oxygen vaporizer system for 24 hours normal oxygen consumption for one module.

Provide liquid nitrogen storage (250 tons) and nitrogen vaporizer (100 tons/hr.).

7.4 Coal Gasification

Design capacity and number of trains specified by coal gasification process owner to provide about 95% on-stream factor for the coal gasification section.

7.5 Acid Gas Removal

Specified by Lurgi for dry ash and slagging gasifiers
For module using B&W, K-T, and Texaco gasifiers:

Design Capacity	105% of normal material balance (gas flow)
-----------------	---

	115% of normal material balance (sulfur compound content)
--	--

Number of Trains	2 - 50% trains
------------------	----------------



7.6 Gas Compression

Design Capacity 105% of normal material
balance

Number of Trains 1 train

7.7 Sulfur Recovery

Design Capacity 115% of normal material
balance

Number of Trains 1 per module + 1 100%
spare for 4 modules

7.8 Sour Water Stripping

Design Capacity 105% of normal material
balance

Number of Trains 1 train

7.9 Ash/Slag Handling

Design Capacity 125% of normal material
balance

Number of Trains 1 train

7.10 Phenol Recovery

Design Capacity 105% of normal material
balance

Number of Trains 1 train

7.11 Ammonia Recovery

Design Capacity 105% of normal material
balance

Number of Trains 1 train



7.12 Coal Handling

Design Capacity	500% of normal material balance for 4 module plant
Number of Trains	2 50% trains except 1 train for barge unloading

7.13 Utility Area

Raw Water Treatment and Storage

Design Capacity	115% of normal material balance for 4 module plant 24 hour storage basis normal material balance for 4 module plant
Number of Trains	1 train

Potable Water Treatment

Design Capacity	115% of normal requirement for 4 module plant
Number of Trains	1 train

BFW and Condensate Treatment

Design Capacity	115% of normal material balance
Number of Trains	4 trains

Steam Generation

Design Capacity	115% of normal material balance
Number of Trains	1 train for each module, plus 1 spare boiler



Plant and Instrument Air and Inert Gas

Design Capacity	Nitrogen provided from
Number of Trains	air separation plant.

7.14 Cooling Water System

Design Capacity	120% of normal material
	balance
Number of Trains	4 trains

7.15 Flare and Incinerator

Design Capacity	Raw gas make from 1 module
Number of trains	4 trains

7.16 Waste Water Treatment

Design Capacity	120% of normal material
	balance
Number of Trains	1 train



SECTION 3.1

GAS CLEANUP

North Carolina



Introduction

The gasification of coal produces a raw gas containing impurities which must be removed before distribution of the product fuel gas. Impurities in the raw gas may be solid particles, sulfur compounds, ammonia, HCl, HCN, tar, oil or phenols.

The reasons for producing a clean fuel gas product are to:

- o Eliminate pollution from users equipment.
- o Ability to use gas in chemical processes where impurities would harm a catalyst, process or product.
- o Eliminate corrosion or blockage from distribution system.
- o Use of product gas without harm to users equipment.

Impurities and Pretreatment

The gasifiers evaluated by FWEC for TVA, namely; Lurgi Dry Ash, BGC/Lurgi Slagger, Babcock and Wilcox, Koppers-Totzek and Texaco all produce raw gas with different concentrations of impurities as well as different ratios of main components. The Lurgi Dry Ash and the BGC/Lurgi Slagger both produce large quantities of impurities such as tar, oils, fatty acids, and phenols. All gasifiers produce H_2S , HCN, COS, and NH_3 . All gasifier raw gas contains some coal dust entrained from the gasifier. The countercurrent action in Lurgi and BGC/Lurgi causes a low temperature at the top of the gasifier. The coal pyrolyzes before gasification producing large quantities of coal tars, phenols, etc. The entrained bed gasifiers operate cocurrently at high temperatures producing no coal tars, phenols, oil, or fatty acids and also producing a very low methane content gas.



The Lurgi Dry Ash gasifier operates at a low temperature controlled by a high ratio of steam to oxygen to prevent clinkering of the ash. The high steam rate in the gasifier produces a high CO_2 and H_2 content gas by means of the water gas shift reaction. In order to prevent plugging of the gasifier downstream equipment, the Lurgi raw gas is washed with a recycled oily water emulsion. This washing step removes most of the tar and coal dust. It removes some of the oils, HCN, NH_3 and phenols. It removes almost no light hydrocarbons, such as naphtha and very little H_2S , COS, and CO_2 .

The BGC/Lurgi slagging gasifier produces similar impurities as the Lurgi Dry Ash. However, tars, oils, and phenols are produced in slightly smaller quantities due to the higher temperature of the slagger.

Traditionally every Lurgi designed gasifier has used the Rectisol process to remove acid gases. The Rectisol process is ideally suited to Lurgi gasifier because it uses low temperature methanol as a physical solvent. No pretreatment of the gas is required to protect the solvent because light hydrocarbons, such as naphtha, condense at low temperature and are absorbed by the methanol solvent. Fortunately, a simple method of separation of methanol and naphtha is available by the use of azeotropic distillation. The methanol solvent is also inert to such impurities as HCN, formic acid, NO_x (present in "K-T" raw gas) or any other acid impurities because methanol is a physical solvent, not a chemical reacting solution like Benfield or Sulfinol. Selexol is also a physical solvent "polyethylene glycol - dimethylether." The solvent



has a high boiling point and it would be difficult to separate from hydrocarbons of similar boiling point carried over in the gas. To prevent contamination of Selexol (or Benfield or Sulfinol) by hydrocarbons such as oil or naphtha, it is necessary to pretreat the gas by means of a low temperature oil wash followed by an activated carbon adsorption to remove the last traces of oil.

The pretreatment step explained above is not necessary in gasification with B&W and with Texaco due to low or no hydrocarbon production. A slightly different pretreatment step may be necessary with "K-T" as this process may produce very small quantities of NO_x which react to form solid deposits. The NO_x is converted by reduction in a "pretreatment" step before acid gas removal.

Hydrolysis of COS

Removal of COS from the raw gasifier gas is required to meet product gas specifications. Some acid gas removal systems which depend upon physical absorption such as Selexol generally do not absorb as much COS as H_2S , requiring large increases in circulation rate to meet low total sulfur requirements. COS is also not a strong acid and so reacts only partially with chemical type of acid gas removal solvents such as alcohol amines. COS tends to degrade alcohol amines such as DEA or DIPA requiring replacement of expensive solutions.

COS may be hydrolyzed to CO_2 and H_2S . The equilibrium is favored by low temperature. However, the rate of reaction drops at low temperature such that even with active catalysts



the reaction temperature is never allowed to fall below about 150°C. Since hot carbonate systems such as Benfield operate in the range of 110-120°C, some hydrolysis does occur. However, the reaction of COS in a Benfield solution is not fast due to the weak catalysis of the solution. By raising temperature some improvement does occur in hydrolysis over a Benfield solution. However, the temperature can be raised a very minor amount before interfering with the solubility of acid gases requiring a higher circulation rate of the activated hot carbonate solution. Raising the temperature also has a deleterious effect on the formation of potassium formate from the carbon monoxide in the gas. The Babcock & Wilcox gas has a Benfield limit of 80°C to minimize formates. As a result, the Benfield system may also require a Catalytic COS hydrolysis reactor ahead of the absorber or between stages to insure removal of total sulfur for moderate or low concentrations in the product MBG.

Systems of acid gas removal such as Selexol may be able to physically absorb (and remove) COS. However, such absorption may incur a high penalty for even minor amounts of COS removal in high circulation rates and high utilities. The Selexol system is often less expensive and meets total sulfur specifications much easier if the COS is first catalytically hydrolyzed and then the H₂S is removed by the Selexol process.

The Rectisol removal process operating at a much lower temperature than Selexol has no difficulty removing COS, mercaptans, or HCN and so requires no catalytic hydrolysis before Rectisol.

COS Hydrolysis Risks

COS is fairly easy to hydrolyze over various catalysts at 300 to 400°F with excess water vapor. Hydrolysis of COS has been practiced on occasion commercially to remove this impurity from natural gas, using alumina, bauxite, or similar catalysts. The problem is somewhat more difficult in a coal conversion raw synthesis gas which contains many compounds and impurities. The possible reactions and products are many and some such as methanation, shift, forming organic sulfur compounds, organic acids, and alcohols may be deleterious. Also to be considered is the possibility that the COS hydrolysis may not be successful in hydrolysis of 95% of the COS, converting something less than that amount of COS. If this should occur, the following options exist:

- Increase catalyst volume
- Increase water vapor in feed gas to drive the reaction further hydrolysis.
- Change catalyst. If an alumina type was used consider bauxite, zeolites or others.
- Consider catalysts recommended by other vendors.
- If hydrolysis is followed by Selexol, consider increasing Selexol circulation rate and decreasing lean solution temperature to absorb more COS.
- Benfield has suggested the use of two COS hydrolysis units each operating at 80% for a total of 96% hydrolysis. One unit was placed before the acid gas removal unit and the second was placed between two absorption stages in the acid gas removal unit.



All of the above options involve some expenditure of money. Solutions are available if the originally installed COS hydrolysis system does not convert COS to the extent promised.

Selectivity of H₂S Removal

In order to economically convert sulfur compounds in the raw gas to elemental sulfur, it is required that the acid gas removal system produce an acid gas with at least 20% H₂S going into the Claus Unit. The penalty for lower concentrations of H₂S is the use of oxygen enriched air to the Claus unit, the use of excess fuel gas in the Claus furnace to maintain a high temperature or the use of a separate sulfur - oxygen combustion chamber to prepare highly concentrated sulfur dioxide to react with the acid gas to form elemental sulfur. These remedies increase expenses and are also untried developments. A better solution would be to feed the low H₂S concentration acid gas into a Stretford unit which converts H₂S to elemental sulfur. This is expensive and the Stretford unit only converts H₂S. The Claus unit, however, when operating at high temperature can convert ammonia to elemental nitrogen and water, thus eliminating a source of NO_x. Raw gas with low CO₂ content such as produced from B&W and from BGC/Lurgi Slagging gasifiers need not be treated selectively since the criteria of 20% concentration of H₂S will be met even if all the CO₂ is removed with the H₂S.



3.1A The Lurgi Dry Ash and BGC/Lurgi Slagging Gasifiers

Conclusions

The Lurgi Dry Ash Gasifier produces a raw gas partially cleansed of particulates, oil and tar by means of a Lurgi quench-scrubber and the method chosen for sulfur removal is the Lurgi Rectisol process. The basis of this choice is experience, ease of handling heavy hydrocarbons and ability to meet low sulfur specifications.

Raw Gas Quality

The raw gases from the Lurgi Dry Ash and from the BGC/Lurgi Slagger gasifiers are similar in that a high concentration of tar, oil, naphtha, and phenols are produced in moving bed gasifiers. They differ, however, in that the BGC/Lurgi Slagger requires very little steam to the gasifier and as a result the raw gas is high in carbon monoxide and low in carbon dioxide and in steam as compared with high CO₂ and steam in the raw gas from the Lurgi Dry Ash. As a result the Sulfur removal from Lurgi Dry Ash raw gas may have to be selective or if nonselective then a Stretford sulfur plant is used instead of a Claus plant.



Particulate Removal

Lurgi Dry Ash (and BGC/Lurgi Slagger) use a proprietary scrubbing column directly after the gasifier. The reason for locating the scrubber directly after the gasifier is that the Lurgi process operating at low temperature produces a large amount of tar which together with the entrained particles tend to plug the piping unless the gasifier exit is washed with "gas-liquor" and scrubbed in the waste heat recovery and in gas cooling sections. This is necessary in order to maintain good heat transfer rates for the exchanger equipment.

Other types of fines removal systems such as electrostatic precipitators, and cyclones are not sufficiently efficient for removal of fines in the size range of one micron and filters such as bag, packed bed, sand or moving bed are not applicable, not highly efficient or not sufficiently developed.

Sulfur Removal

For the present Phase I study, sulfur compounds in the product gas are limited to 200 ppm (v) by specification. Raw gas may contain 17,000 ppm requiring a removal of 98.8% of all the sulfur in the raw gas. In the coal gasification plant conceptual designs, gas cleanup will operate at a pressure slightly lower than the gasifier pressure. Compression will be applied to the clean gas sufficient to produce product gas at a pressure of



600 psig at the plant limits. Sensitivity studies will involve delivery pressures of 800 psig and of gasifier operating pressure less equipment pressure drops. At the higher gas pressure of 800 psig, the same pressure for acid gas removal will be used with the exception that the clean gas from the treating section will be further compressed to 900 psig. For production of gas at gasifier pressure less the equipment pressure drops, the same pressure for acid gas removal will again be used, but compression will be eliminated.

Selection of acid gas removal processes in general change with operating conditions such as temperature, pressure, and capacity, with sulfur specifications, with CO₂ concentration, and with minor constituents which may affect the solvent. The following factors are involved in the choice of process:

- o Ability of purification process to meet sulfur specifications.
- o Selectivity of process to keep H₂S concentration in the acid gas high enough for feed to a Claus plant (important only in the Lurgi Dry Ash and Texaco gasifiers.)
- o Tendency of fuel gas components to contaminate the solvent. Examples are naphtha and oil absorbed by solvents and not easily stripped and HCN reacting with MEA or with Sulfinol.
- o Energy consumption of each process.
- o Removal of nitrogen compounds which may cause NO_x pollution.
- o Tendency of some solvents to contaminate the fuel gas with arsenic or other poisons.



After scrubbing, the gas is low in temperature and it is best to treat at a temperature below 300°F, in order to maintain high thermal efficiency.

Table 3.1A-1 lists a number of possible processes. Some processes lack selectivity and others may entrain compounds harmful to man and equipment. Some processes must be eliminated due to lack of experience or acceptance of these processes, except in very limited fields. Also eliminated are those batch-cycle processes which cannot be applied to large scale and large amounts of H₂S.



Form No. 130-171

TABLE 3.1A-1
 PROCESSES* FOR REMOVING H₂S AND CO₂ FROM LOW - AND INTERMEDIATE - BTU GASES

PROCESS	SCRBENT	REMOVES	TEMPERATURE
Amine	Monethanolamine (MEA), 15% in water	CO ₂ , H ₂ S	70-100 °F
Economine	Diglycolamine (DAG), 50-70% in water	CO ₂ , H ₂ S	70-100 °F
Alkaziid	Solution M or DIK (potassium salt of dimethylamine acetic acid), 25% in water	H ₂ S small amount of CO ₂	ambient
Benfield Catacarb	Hot potassium carbonate, 20-30% in water, (also contains catalyst)	CO ₂ , H ₂ S Selective to H ₂ S	100-250 °F
Purisol (Lurgi)	N-methyl-2-pyrrolidone	H ₂ S, CO ₂	ambient or lower
Fluor	Propylene carbonate	H ₂ S, CO ₂	ambient or lower
Selexol (Allied)	Dimethyl ether polyethylene glycol	H ₂ S, CO ₂	ambient or lower
Rectisol (Lurgi)	Methanol	H ₂ S, CO ₂ Selective to H ₂ S	cold
Sulfinol (Shell)	Tetrahydro Thiophene 1-1, dioxide (sulfolene) plus diisopropanol amine (DIPA)	H ₂ S, CO ₂ Selective to H ₂ S	ambient
Giammarco-Vetrocoke	K ₃ AsO ₃ activated with arsenic	H ₂ S	50-212 °F
Stretford	Water solution of Na ₂ CO ₃ and anthraquinone disulfonic acid (ADA with activator of sodiummeta vanadate	H ₂ S	approx. 50-100 °F

*Flow diagrams for these processes can be found in Hydrocarbon Processing, Vol. 4, April 1973): 87-109



These criteria reduce the available gas treating processes for the Lurgi type Gasifiers to the following list:*

- o Hot Carbonate (Benfield)
- o Selexol (Allied Chemical)
- o Rectisol (Lurgi)
- o Sulfinol (Shell Development)

Rectisol is a good application for both Lurgi Dry Ash and BGC/Lurgi Slagger because gas pretreatment to remove hydrocarbons is not necessary with the Rectisol solvent. Rectisol also completely dries the gas and easily removes all COS and HCN. Other possible processes are Selexol and Benfield or Sulfinol with advantage of some processes depending upon the partial pressure of acid gases as shown in Figure 3.1 A-1. Partial pressure of acid gas is not a sufficient criteria to choose a process as sulfur purity specifications, sulfur selectivity and energy consumption are also important.

From approximate raw gas compositions and quantities (Table Nos. 3.1A-2 and 3.1A-3) for the moving bed gasifiers, it can be seen that CO concentrations are very high. The Benfield process is limited by the possible formation of formic acid produced by the reaction of CO at high concentrations with a strongly basic solution

* A description of the above processes and a reference to literature are found after section 1.1C1.



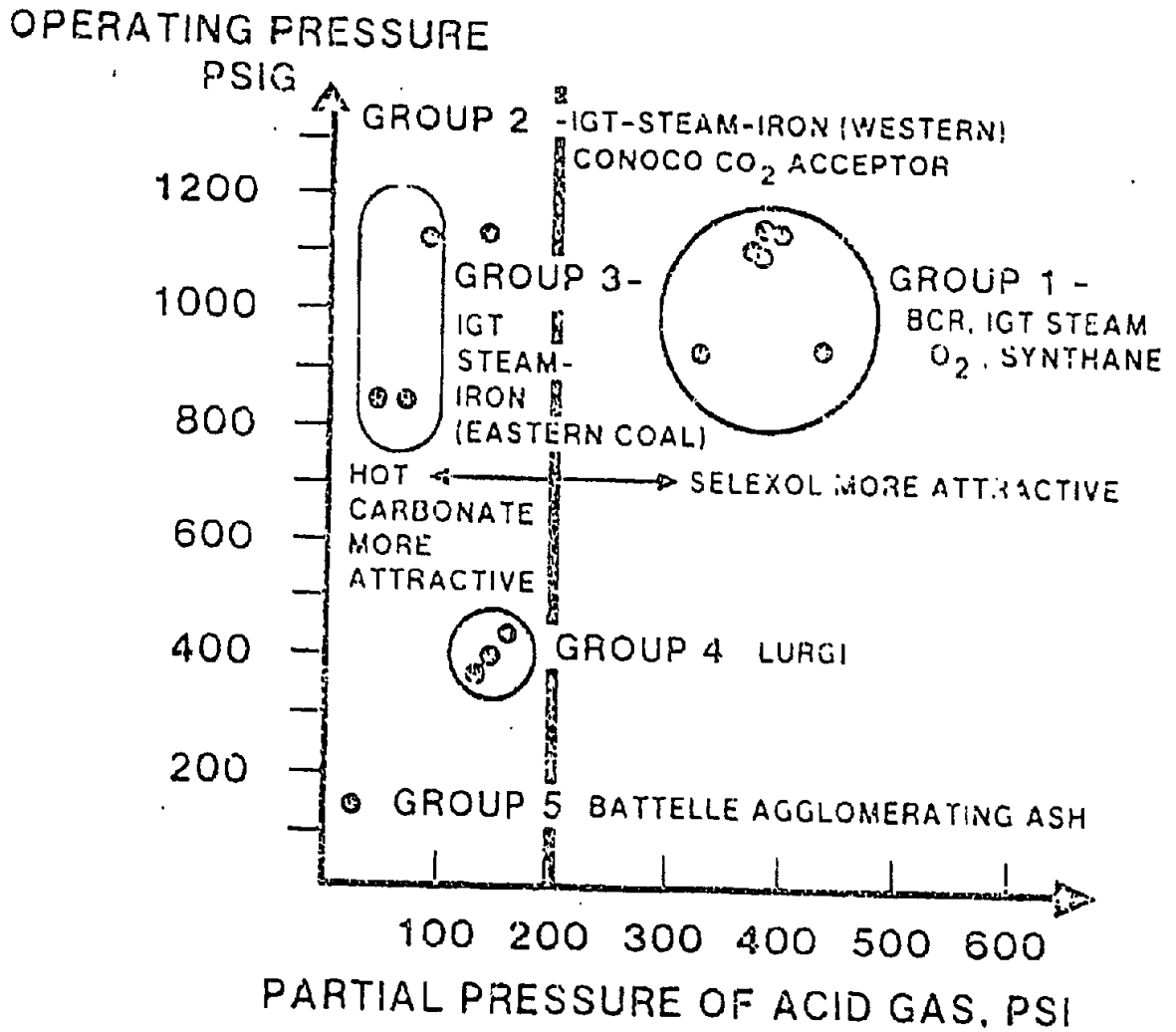
at temperatures above 100^o-120^oC. The Benfield solution at low temperatures, below 110-120^oC would not hydrolyze COS, which is necessary if the product gas specification of 200 ppm is to be met, thus both a COS catalytic hydrolysis unit and a pretreatment step to remove hydrocarbons would be required for Benfield.

Both Benfield and Sulfinol solutions would be adversely affected by strong acids such as formic acid and hydrogen cyanide.

COMPARISON OF ACID GAS
REMOVAL PROCESSES

FIGURE 3.1A-1

PLOT OF FEED STREAMS



Ref. No. 5



TABLE 3.1A-2

LURGI DRY ASH GASIFIER

ACID GAS REMOVAL

A. Approximate Composition of Feed Gas

<u>COMP</u>	<u>MOL%</u>	<u>M/HR</u>	<u>LBS/HR</u>
H ₂	39.06		
CO	18.57		
CO ₂	29.95		
C ₁	9.56		
C ₂ ⁺	0.65		
N ₂ ,Av	0.23		
H ₂ S	1.51		
COS	<u>0.04</u>		
	99.57	32,411	679,423
H ₂ O	<u>0.43</u>	<u>140</u>	<u>2,520</u>
	100.00	32,551	681,943

Pressure Psig 370

Temperature of 120

MW 20.95

Plant Specification

Product gas total sulfur ppmv (max) 200

H₂S in acid gas min vol% 20

Form No. 130-171



TABLE 3.1A-3

BGC/LURGI SLAGGER GASIFIER

ACID GAS REMOVAL UNIT

A. Feed Gas Composition (Approx)

<u>Comp</u>	<u>Vol %</u>	<u>M/Hr</u>	<u>lbs/Hr</u>
H ₂	25.81		
CO	58.58		
CO ₂	6.43		
C ₁	5.96		
C ₂ ⁺	0.16		
H ₂ , Ar	0.71		
H ₂ S	1.81		
COS	<u>0.11</u>		
	99.57	28,701	610,424
H ₂ O	<u>0.43</u>	<u>709</u>	<u>12,758</u>
	100.00	29,410	623,000

Pressure Psig 370

Temp °F 120

MW = 21.18

Gas purity ppmv sulfur 200

Minimum H₂S in acid gas Mol% 20

FORM NO. 130-1/1



Selection of Acid Gas Removal Processes

The acid gas removal method for the Lurgi Dry Ash and BGC/Lurgi gasifiers is characterized by a base case operating pressure of 370 psig and a purification of the product gas to a sulfur content of 200 ppm (v). The acid gas removal method selected for the Lurgi Dry Ash and BGC/Lurgi Slagger gasifiers is the Rectisol process.

The reasons for the above choice are the ability of the Rectisol process to reduce the sulfur level to 0.1 ppm, selectivity of the removal process towards sulfur removal with minimum carbon dioxide removal, minimum poisoning of solution and no pre-treatment costs for hydrocarbon removal, no requirement for COS hydrolysis, and long experience with the use of Rectisol applied to many Lurgi plants.

Benfield for this application would require some additions to meet specifications, such as:

- o Pretreatment of gas to reduce hydrocarbon contamination of Benfield solution.
- o Catalytic COS hydrolysis before Benfield or between absorption stages.
- o Acceptance of large losses of solution, especially for the ease of BGC/Lurgi due to the formation of potassium formate.



Sulfinol and Selexol are applicable in the 200 ppm sulfur purity range of product gas. Selexol requires a pretreatment step or the oils and phenols will contaminate the Selexol solvent. Sulfinol also requires pretreatment and COS hydrolysis. Sulfinol cannot reduce the sulfur content to lower than 5 ppm.

*Economics of Acid Gas Removal for Lurgi Dry Ash and BGC/Lurgi

Tables 3.1A-4 and 3.1A-5 summarize the economics for the removal of acid gases from the Lurgi type of processes. The utility costs are based upon the prices of \$1.25/mm Btu for coal plus the cost of 2.7¢/Kwhr for power, with steam at \$1.50/1000 lbs and cooling water at 2¢/1000 gals. It may be noted that Sulfinol is more expensive than other processes for Lurgi Dry Ash due to the high CO₂ content of the raw gas, most of which must be removed by Sulfinol. The BGC/Lurgi does not include Benfield due to the high partial pressure of CO forming formates. Of the other acid gas removal methods no one process is clearly cheaper than the other. Criteria other than economics, involving experience, flexibility to reduce sulfur content to values below 1 ppm, and ability to recover heavy hydrocarbons make Rectisol the process of choice.

*.Utilities and Plant Investments are taken from References 4, 8, 15, and 16.

.Other operating costs include labor, benefits, supervision, G and A, maintainance and supplies.

.Payout includes depreciating interest, land, and working capital.



TABLE 3.1A-4

ECONOMICS OF ACID GAS REMOVAL (PER MODULE)

LURGI DRY ASH CASE

<u>A. Process</u>	<u>Rectisol</u>	<u>Selexol</u>	<u>Benfield</u>	<u>Sulfinol</u>
Operating Pressure Psig	370	370	370	370
Gas Purity PPM Sulfur	200	200	200	200
<u>B. Operating Cost</u>	<u>MM \$/Yr.</u>	<u>MM \$/Yr.</u>	<u>MM \$/Yr.</u>	<u>MM \$/Yr.</u>
Steam @ 1.50 \$/1000 lbs	1.98	1.01	1.97	3.88
C.W. @ 2 ¢/1000 Gals (BFW x 10)	0.28	1.33	0.28	0.55
Power @ 2.7 ¢/Kw Hr	0.56	1.29	0.70	0.76
Solvent	0.09	0.08	0.05	0.30
Other Operating Costs	<u>1.00</u>	<u>1.00</u>	<u>1.00</u>	<u>1.00</u>
	3.91	3.71	4.00	6.49
Approx. Plant Inv. MM \$	27	24	20	30
Addition for Pretreatment MM \$		4	4	4
Addition for COS Hydrolysis MM \$		<u>2</u>	<u>2</u>	<u>-</u>
Total MM \$	(27)	(30)	(26)	(34)
Assume 6 Yr. Payout	4.50	5.00	4.33	5.67
Operating Costs & Utilities	<u>3.91</u>	<u>3.71</u>	<u>4.00</u>	<u>6.49</u>
Approx. Cost MM \$/Yr.	8.41	8.71	8.33	12.16



TABLE 3.1A-5

ACID GAS REMOVAL (PER MODULE)

BGC/LURGI SLAGGER GASIFIER

<u>A. Process</u>	<u>Rectisol</u>	<u>Selexol</u>	<u>Sulfinol</u>
Pressure Psig	370	370	370
Gas Purity PPMV(s)	200	200	200
<u>B. Operating Cost MM \$/Yr</u>			
Steam at 1.50 \$/1000 lbs	0.97	0.96	1.36
C.W. at 2 ¢/1000 Gal	0.27	0.31	0.39
Power at 1.7 ¢/Kw Hr	1.44	0.87	0.46
Solvent	0.08	0.10	0.15
Other Operating Costs	<u>1.00</u>	<u>1.00</u>	<u>1.00</u>
	3.76	3.24	3.36
Plant Investment (Approx.) MM \$	30	24	26
Additional for Pretreat		4	4
Additional for COS Hydrolysis		2	
Total	(30)	(30)	(30)
6 year payout MM \$/Yr	5.00	5.00	5.00
Operating Cost	<u>3.76</u>	<u>3.24</u>	<u>3.36</u>
	8.76	8.24	8.36



Alternative Acid Gas Removal Applications

A. Pressure Variations

The product gas delivery pressures of 600 and 800 psig have little effect on the acid gas removal system since the acid gas removal operates at 370 psig and the clean product gas is simply boosted in pressure to 600 or 800 psig.

B. Variation in Sulfur Specification in Product Gas

The drop of the total sulfur specification from 200 ppm to 1 ppm imposes additional work on the absorption solvent, increasing both the operating cost and the capital investment for both Selxol and Rectisol. Sulfinol can not reduce sulfur below 5 ppm. Benfield can not meet 1 ppm sulfur due to both COS which is partially hydrolysed by Benfield solution and also to organic sulfur compounds such as mercaptans which are not removed by Benfield solution. The economics of acid gas removal at both 1 ppm and 200 ppm total sulfur specification are not clear-cut. This can be seen in Table 3.1A-6. In the absence of a definite economic advantage it would be best to choose Rectisol based upon many years of experience of Lurgi Dry Ash gasifiers operating with Rectisol units.



TABLE 3.1A-6

ACID GAS REMOVAL TO 1 PPM SULFUR

LURGI DRY ASH PROCESS

<u>A. Process</u>	<u>Rectisol</u>	<u>Colexol</u>
Pressure Psig	370	370
Gas Purity PPM Sulfur	1	1
<u>B. Operating Costs</u>	<u>MM \$/Yr.</u>	<u>MM \$/Yr.</u>
Steam @ 1.50 \$/1000 lbs	2.28	1.21
C.W. @ 2 ¢/1000 Gal	0.36	0.40
Power @ 0.027 \$/Kw Hr	1.38	2.28
Solvent	0.11	0.10
Other Operating Costs	<u>1.00</u>	<u>1.00</u>
	5.13	4.99
Approx. Plant Inv. MM \$	32	27
Addition for Pretreatment MM \$	-	4
Addition for COS Hydrolysis MM \$		<u>2.0</u>
Total MM \$	(32)	(33)
6 Year Payout MM \$/Yr.	5.33	5.50
Operating Costs Plus Utilities MM \$/Yr.	<u>5.13</u>	<u>4.99</u>
Total Cost MM \$/Yr.	10.46	10.49

Form No. 130-171



Removal of Carbon Dioxide

Some carbon dioxide is removed in all acid gas removal systems. The amount of removal is a function of the selectivity of the process. In addition to the amount of CO_2 which is removed together with the H_2S , an additional amount of CO_2 in some gasification process such as Texaco and Lurgi Dry Ash, must be removed from the product gas in a second stage acid gas absorption in order to meet the product gas heating value requirements. These gasifiers operate with high steam feed which reacts with carbon monoxide to form hydrogen and carbon dioxide by the water-gas shift reaction.

The amount of CO_2 remaining in Lurgi type of gasifier product gas may be quite high and still meet the heating value requirements due to the high methane content of the Lurgi gas. Economically it does not pay to remove any more CO_2 than that necessary to just meet the product gas heating requirements. Documentation for the above statement is presented in Table 3.1A-7. The cost of removing one mole of CO_2 is 19.2¢. The savings in compression and transportation is 5.9¢ per mole of CO_2 for a net cost of 13.3¢/mol.



TABLE 3.1A-7

Economics of Removing Excess CO₂ for Gas Heating Value

o Acid Gas Removal		
1976 Cost	117 MM \$	(250 MM SCFD SNG)
Eng, Fee Contingency	<u>26</u>	
	143 MM \$	
o Escalate to 1980 = 143 (1.09) ⁴ = 143 x 1.41 = 202 MM \$		
o Mols CO ₂ Per Hour = 37,000		
o Utilities		
		<u>\$/Hr.</u>
Steam 900,000 \$/hr @ 1.50/1000		1350
Power 6300 Kwh/h x 2.7 ¢/Kwh		170
C.W. 32,000 Gpm x 2 ¢/1000 Gal x 60		<u>38</u>
		1558
Other		1348
6 yr payout = \$202 MM/6 x 8000		<u>4208</u>
		\$7114

Removal

$$\text{Cost Per Mol CO}_2 = \frac{\$7,114 \times 100}{37,000} = 19.23 \text{ ¢/Mol CO}_2$$

Cost of Compression and Transportation of Residual CO₂

o Power to compress one mol of CO ₂ from 350 to 650 psia	
o BHP = 40/MM CF/24 Hr x $\frac{380 \times 24}{10^6}$ = 0.365 BHP/Mol CO ₂	
o Actual Drive HP = 0.365 BHP x 1.8 = 0.657 HP = 0.5 Kw/Mol CO ₂	
o 6 Yr payout $\frac{3000 \text{ \$/HP} \times .657}{6 \text{ yr} \times 8000 \text{ h/yr}} \times 100 \text{ ¢/\$} = 4.10 \text{ ¢}$	
Power = 0.5 x 2.7 ¢/Kw	1.35 ¢
Transport at 300 miles	<u>0.43 ¢</u>
	5.88 ¢/mol

Net Cost of Removing One Mol of CO₂

$$\text{Cost} = 19.23 - 5.88 = 13.35 \text{ ¢/Mol}$$

Form No. 130-171

3.1B-1 BABCOCK AND WILCOX GASIFIERIntroduction

The gasification of coal by Babcock and Wilcox (B & W) gasifier produces a raw gas containing impurities such as solid particles, sulfur compounds, ammonia, HCl, HCN. The B&W gasifier produces no hydrocarbons, phenols or tars and produces little methane. The Texaco and Koppers-Totzek gasifiers are similar to B&W in that all three are high temperature entrained flow gasifiers producing no tars or oils.

Conclusions

The Babcock and Wilcox gasifier raw gas has dust and particulates removed by means of venturi type scrubbers and the sulfur in the gas is removed by means of the Selexol process. Selexol is not the clear economic choice.



However, it is selected on the basis of:

- o Good operating experience
- o Ability to reduce sulfur to very low levels (0.1 ppmv)
- o Selectivity to H₂S removal

Particulate Removal

The Babcock and Wilcox process operates at a much higher temperature than the BGC or Lurgi processes and produces no tars, oils or phenols and produces very much less methane and ammonia than Lurgi. After heat exchange the gas is cooled and scrubbed with a venturi type scrubber to remove solids. However, since the gas must be treated and compressed, solid particles above one micron in diameter should be removed. To meet this stringent specification, two stages of venturi scrubbing are provided. Other types of fines removal such as electrostatic precipitators, and cyclones are not sufficiently efficient for removal of fines in the size range of one micron and filters such as bag, packed bed, sand or moving bed are not applicable, not highly efficient or not sufficiently developed.



Sulfur Removal

An alternative to treating for sulfur removal at gasifier pressure and then compressing clean gas would be the compression of sour gas with all acid gases to a few atmosphere pressure above pipeline pressure and then treat the sour gas at a higher pressure than in the base case. The alternative case of treating at high pressure was not chosen even with the acid gas removal system operating at a lower circulation and at slightly lower capital and operating cost than in the lower pressure gas treating case. The reason is that compression of sour gas is more expensive due to both higher capacity than compressing sweet gas and due to higher equipment cost because of metallurgical and mechanical modifications of the compressor as well as increased maintenance costs resulting from the handling of sour, dirty gases.

The case of modification of gas specification to 1 ppm total sulfur instead of 200 ppm at 600 psig may limit the choice of process as some processes may not be capable of meeting the specification of 1 ppm(v) total sulfur.

Process Selection for Gas Purification

The following factors are involved in the choice of process:

- o Ability of purification process to meet sulfur specifications of 200 ppmv.
- o Tendency of fuel gas components to contaminate the solvent. Examples are naphtha and oil absorbed by solvents



and not easily stripped and HCN reacting with MEA or with Sulfinol.

- o Ability of process to remove total sulfur to very low concentrations of 1 ppm.
- o Energy consumption of each process.
- o Removal of nitrogen compounds which may cause NO_x pollution.
- o Tendency of some solvents to contaminate the fuel gas with arsenic or other poisons.

Acid gases (H₂S and CO₂) may be removed or reduced by many solvents and many processes. Before treating, the gas is scrubbed in the venturi with water. The scrubbing removes most of the particulates as well as most of the HCN, HCl, and NH₃. After scrubbing, the gas is low in temperature and it is best to treat at a temperature below 300°F, in order to maintain high thermal efficiency

These criteria reduce the available processes to the following list:

- o Hot carbonate (Benfield)
- o Seloxol (Allied Chemical)
- o Rectisol (Lurgi)
- o Sulfinol (Shell Development)

Rectisol may not be competitive in the gas specification range of 200 ppmv total sulfur. Rectisol has the ability to reduce the sulfur content to 0.1 ppm. None of the processes



require pretreatment for oil removal as the B & W process does not produce hydrocarbons.

From approximate raw gas compositions and quantities (Table No. 3.1B-1) it can be seen that the amount of H_2S is approximately the same for each of the five types of gasifiers. The CO_2 is lowest in the B & W gasifier, and CO is highest in concentration. At the concentrations of components given by B & W the Benfield process is borderline in applicability. The CO concentration is 60 volume % and CO partial pressure is 144 psia. The Benfield process may be in trouble due to the formation of formates with carbon monoxide requiring a high makeup rate of solution, and due to non-regenerability with high H_2S and low CO_2 concentrations.

The acid gas removal from Babcock and Wilcox raw gas is characterized by a base case operating pressure of 200 psig and a purification of the product gas to a sulfur content of 200 ppm (v). The acid gas removal method selected for B & W is the Selexol process.

The reasons for the above choice, as applied to the base case conditions, are technical considerations, sulfur removal with minimum carbon dioxide removal, minimum poisoning of solution, and no pretreatment costs as well as the ability to remove sulfur to a level of less than 1 ppm.



TABLE NO. 3.1B-1

BASE CASE ACID GAS REMOVAL

BABCOCK AND WILCOX GASIFIER

A. Raw Gas To Acid Gas Removal

<u>Comp</u>	<u>Vol%</u>	<u>M/Hr</u>	<u>lbs/Hr</u>	<u>Mw</u>
H ₂	29.324			
CO	60.491			
CO ₂	4.554			
NH ₃	0.003			
HCl	0.044			
N ₂ Ar	3.274			
H ₂ S	1.425			
COS	<u>0.097</u>			
	99.212	33,958	721,864	
H ₂ O	<u>0.788</u>	<u>270</u>	<u>4,860</u>	
	100.000	34,228	726,724	21.23

Pressure Psia 240

Temperature °F 120

Specification

1. Total sulfur in treated gas ppmv max 200
2. Minimum H₂S in acid gas vol % 20
3. Alternative sulfur specification ppm 1

Form No. 130-171



A study of the raw gas compositions presented in Table 3.1B-1 indicates that no selectivity toward H_2S removal is required for the Babcock and Wilcox gasifier. The reason is that even if all the CO_2 were removed with the H_2S the concentration of H_2S in the acid gas would be sufficiently high (above 20% H_2S) to feed to a Claus unit. Also the raw gas minus the sulfur components is high enough in heating value so that it is not necessary to remove a pure CO_2 stream to increase the heating value of the gas.

The Benfield application to either B & W or BGC/Slagging Lurgi raw gas is limited by incomplete COS hydrolysis, formate formation and a high H_2S to CO_2 ratio. The Sulfinol application to B&W is economic as may be seen in Table No. 3.1B-2, but it is limited by a maximum reduction of sulfur to 5 ppmv. Sulfinol may only be applied to the B & W (Case I) due to the very high solubility of any hydrocarbon in the Lurgi Dry Ash or BGC/Slagging Lurgi, requiring very large solution circulation rates and heavy hydrocarbons in acid gas to the Claus unit produces dark sulfur in both Lurgi cases forcing the use of pretreatment to remove naphtha and light hydrocarbons.



TABLE 3.1B-2

ACID GAS REMOVAL (PER MODULE)BABCOCK AND WILCOX GASIFIER

<u>Process</u>	<u>Rectisol</u>	<u>Selexol</u>	<u>Benfield</u>	<u>Sulfinol</u>
Operating Pressure Psig	225	225	225	225
Feed Gas Temperature °F	120	120	220	120
Gas Purity PPMV of Sulfur	200	200	200	200
<u>Operating Costs in MM \$/Yr</u>				
Steam at 1.50 \$/1000 lbs	0.72	0.91	2.19	1.96
C.W. at 2 ¢/1000 Gal	0.31	0.27	0.59	0.38
Power at 2.7 ¢/Kw Hr	1.30	1.58	0.35	0.63
Solvent	0.08	0.10	0.28	0.12
Operating Operating Costs	<u>1.00</u>	<u>1.00</u>	<u>1.00</u>	<u>1.00</u>
Total MM \$/Yr	3.41	3.86	4.41	4.09
<u>Plant Investment (Approx MM \$)</u>				
Added COS Hydrolysis		<u>2</u>	<u>2</u>	<u>2</u>
Total	(28)	(24)	(20)	(22)
6 year payout MM \$/Yr	4.67	4.00	3.33	3.67
Operating Costs	<u>3.41</u>	<u>3.86</u>	<u>4.41</u>	<u>4.09</u>
Total MM \$/Yr	8.08	7.86	7.74	7.76



Alternative Acid Gas Removal Applications

A. Pressure Variations

The product gas delivery pressure of 600 and 800 psig has no effect on the acid gas removal system since the acid gas removal operates at 200 psig and the product gas is simply boosted in pressure to 600 and 800 psig.

The case of product gas delivery pressure at gasifier pressure (less system pressure drop) means that for the B & W case the acid gas system will be operating at about 200 psig.

B. The Alternative Sulfur Specifications

Lowering the sulfur specification from 200 ppm in the direction of 1 ppm total sulfur changes the availability of treating processes, but does not modify the choice of the Selexol process for the application to 1 ppm of total sulfur provided a COS hydrolysis unit is included. The Sulfinol system should not be used below 5 ppm. Benfield may not have difficulty if all the sulfur compound were H_2S . However, with COS, even with a hydrolysis step, the unit may have difficulty attaining 1 ppm. Both Rectisol and Selexol plus COS hydrolysis can attain the low sulfur specification. In the absence of a clear economic advantage of one over the other, it would be best to follow the same choice at 1 ppm as was chosen for 200 ppm total sulfur, namely the Selexol process.

Variation in CO_2 & COS Content

There is no economic incentive to remove the few percent of CO_2 from B & W raw gas. The economics is similar to that shown in Table 3.1A-7. The COS must be reduced for Selexol and Benfield to meet a total sulfur specification of 1 ppm.



The COS does not have to be hydrolyzed for Rectisol or Sulfinol. For the 200 ppmv sulfur specifications, Benfield must use catalytic hydrolysis of COS to meet specifications and the Selexol may use catalytic hydrolysis of COS because of less expensive design for Sulfur Removal.



3.1B-2 TEXACO GASIFIER

Introduction

The gasification of coal by Texaco gasification process produces a raw gas containing impurities which must be removed before distribution of the product fuel gas. Impurities in the raw gas may be solid particles, sulfur compounds, ammonia, HCl, HCN. Texaco produces no oils, tars or phenols and very little methane. Both Texaco and Kopper-Totzek raw gas cleanup is similar to Babcock and Wilcox. The differences between entrained flow gasifiers will be discussed.

Conclusions

The Texaco gasifier raw gas has dust and particulates removed by means of venturi type scrubbers and the sulfur in the gas is removed by means of the Selexol process. Selexol is not the clear economic choice.

However, it is selected on the basis of:

- o Good operating experience
- o Ability to reduce sulfur to very low levels (0.1 ppmv)
- o Selectivity to H₂S removal

Particulate Removal

The gasifier effluent contains solid particles of soot and ash. Raw gas is first scrubbed with water in the convection cooler, where most of the ash is removed and leaves the conical shroud of the cooler as a slurry. The fine particles of ash and carbon leaving with the gas are removed in a three stage scrubbing system



after further cooling of the gas. This system assures a very efficient removal of solid particles as well as removal of HCN, HCl and NH₃.

Sulfur Removal

For the present Phase I study, sulfur compounds in the product gas are limited to 200 ppm (v) by specification. Raw gas may contain 15,000 ppm requiring a removal of 98.7% of all the sulfur in the raw gas. In the coal gasification plant conceptual designs, gas cleanup will operate at a pressure slightly lower than the gasifier pressure. Compression is not necessary to produce product gas at a pressure of 600 psig at the plant limits. Sensitivity studies will involve delivery pressures of 800 psig. At a higher gas pressure of 800 psig, a pressure of 600 psig for acid gas removal will be used with the exception that the clean gas from the treating section will be further compressed to 800 psig, thus eliminating the need for compression of wet sour gas with a larger compressor having more expensive metallurgy.

Process Selection for Gas Purification

From approximate Texaco raw gas composition and quantities (Table No. 3.1-B-3), it can be seen that the amount of H₂S is approximately the same as for B&W. The considerations that led to the selection of Selexol process for gas cleanup for B&W gasifier are also valid for Texaco process. The higher concentration of CO in Texaco gas makes Benfield process less appropriate due to the possibility of

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formate formation, while the higher operating pressure than in B & W case makes Selexol more efficient and less expensive as far as operating cost is concerned. Rectisol is slightly more expensive for the 200 ppm case and Sulfinol, while it is competitive for the 200 ppm case, cannot be applied for the alternate case (1 ppm) because of its inability to reduce the sulfur below 3 to 5 ppm.

The Selexol process is the selected cleanup process for the Texaco gasifier.



TABLE NO. 3.1B-3

BASE CASE ACID GAS REMOVAL

TEXACO GASIFIER

Raw Gas To Acid Gas Removal

<u>Comp</u>	<u>Vol%</u>	<u>lbmol/Hr</u>	<u>lbs/Hr</u>	<u>Mw</u>
H ₂	34.381			
CO	45.902			
CO ₂	16.539			
CH ₄	0.300			
N ₂ Ar	1.266			
H ₂ S	1.469			
COS	<u>0.005</u>			
	99.862	35635.4	777,216	
H ₂ O	<u>0.138</u>	<u>49.2</u>	<u>886</u>	
	100.000	35684.6	778,102	21.80

Pressure Psia 687

Temperature °F 100

Specification

1. Total sulfur in treated gas ppmv max 200
2. Minimum H₂S in acid gas vol % 20
3. Alternative sulfur specification ppm 1



Alternative Acid Gas Removal Applications

A. Pressure Variations

The product gas delivery pressure of 600 and 800 psig has no effect on the acid gas removal system since the acid gas removal operates at 670 psig and the product gas is simply boosted to 800 psig. As an alternate for 800 psig delivery pressure, the gasification process can operate at a pressure above 800 psig (800 psig plus the pressure drop through the system).

The most economical manner of producing product gas at 800 psig is to design the gasifier to operate at slightly above 800 psig, eliminating raw or product gas compression and only requiring oxygen compression to 800 psig. This is possible with Texaco as the pilot plant has operated at 1200 psig. Other gasifiers have not demonstrated high pressure operation.

The best method of producing 800 psig product gas for those gasifiers which must operate at low to moderate pressures is to remove acid gas at the gasifier pressure in such units as B&W (200 psig) or Lurgi (350 psig) and then compress clean gas to 800 psig. The K-T gasifier operation at atmospheric pressure is best compressed to 300 psig for acid gas removal and then recompressed to 800 psig. While economics may be similar for alternative positions of compression and acid gas removal, technically the more conservative engineering approach is the configuration which allows compression of a clean, dry gas, rather than a wet sour gas.



B. The Alternative Sulfur Specifications

Lowering the sulfur specification from 200 ppm in the direction of 1 ppm total sulfur changes the availability of treating processes, but does not modify the choice of the Selexol process for the application to 1 ppm of total sulfur provided a COS hydrolysis unit is included. The Sulfinol system should not be used below 3 to 5 ppm. Benfield may not have difficulty if all the sulfur compound were H₂S. However, with COS, even with a hydrolysis step, the unit may have difficulty attaining 1 ppm. Both Rectisol and Selexol plus COS hydrolysis can attain the low sulfur specification. In the absence of a clear economic advantage of one over the other, it would be best to follow the same choice at 1 ppm as was chosen for 200 ppm total sulfur, namely the Selexol process.

C. Variation in CO₂ & COS Content

For Texaco process, it is necessary to remove part of CO₂ from the raw gas in order to meet the HHV specification. The amount of COS in Texaco raw gas is very low. This combined with the necessity of CO₂ reduction makes unnecessary the hydrolysis of COS prior to H₂S absorption for 200 ppm sulfur content in product gas. For 1 ppm specification, and a higher content of COS in raw gas, the addition of COS hydrolysis unit is required.



As indicated in Table 3.1A-7 "Economics of Removing CO₂", the removal of CO₂ in excess of heating value requirements is not justified. For the Texaco case, with the gasifier operating at a bit above product gas pressure of 600 psig, there is even less justification as there is no credit for any savings of compression of CO₂ from gasifier pressure to product gas pressure.



3.1B-3 KOPPERS-TOTZEK GASIFIER

Introduction

The gasification of coal by Koppers-Totzek (K-T) gasifier produces a raw gas containing impurities which must be removed before distribution of the product fuel gas. Impurities in the raw gas may be solid particles, sulfur compounds, ammonia, HCl, HCN plus traces of NO_x , O_2 and elemental sulfur.

Conclusions

The Koppers-Totzek gasifier raw gas has dust and particulates removed by means of scrubbers, disintegrator/separator and electrostatic precipitators, and the sulfur in the gas is removed by means of the Selexol process. Selexol is not the clear economic choice.



However, it is selected on the basis of:

- o Good operating experience
- o Ability to reduce sulfur to very low levels (0.1 ppmv)
- o Selectivity to H₂S removal

Particulate Removal

Koppers-Totzek process, due to the high operating temperature, produces no tars, oils or phenols and produces a gas with a very low content of methane. Approximately 25% of the ash contained in the coal leaves the bottom of gasifier as molten slag. The remainder of the ash is carried with the gas. The gas leaving the gasifier is tempered with water quench before entering a waste heat boiler positioned immediately above the gasifier. Part of the fly ash settles into the bottom of the waste heat boiler and is removed as a dilute water slurry. Most of it is carried over to the cooler/washer tower where it is washed from the gas and removed as dilute water slurry. The third step of the fly ash removal takes place in the disintegrator/separator that follows the cooling washer. The scrubbed gas then flows to a blower which provides sufficient head to transfer the gas where before any operation, the gas passes through electrostatic precipitators for the final removal of fly ash. After this treatment, the gas is free of any solid particles.



Sulfur Removal

The K-T gasifiers operate at essentially atmospheric pressure. As a result, it is necessary to compress the gas before or after acid gas removal in order to deliver fuel gas product at a pressure of 600 psig as specified. Foster Wheeler analysis of possible alternates for compression and acid gas removal indicated that two major considerations are involved. The first consideration is the necessity of removing trace amounts of nitrogen oxides which are present in the raw gas in order to avoid operating difficulties during compression.

The Texaco and B&W raw gases do not seem to contain traces of nitrogen oxides. This appears reasonable as both Texaco and B&W operate at much higher pressure than K-T. Higher pressure and longer residence time at high temperatures mean that less likelihood exists of oxygen or oxides of nitrogen will break through in the reducing atmosphere of the raw gas.

Nitrogen oxides are formed in small amounts in the K-T gasifier by reaction of nitrogen compounds (either molecular nitrogen present in the oxygen supplied to the gasifier or nitrogen compounds formed from the coal during gasification) and oxygen at the high temperatures existing in the gasifier.



Patridge /1/ described difficulties experienced in the AECI coal-based ammonia plant located near Johannesburg, South Africa arising from deposition of elemental sulfur in compressors and in a Rectisol unit.

Elemental sulfur was apparently formed by oxidation of hydrogen sulfide by traces of oxygen in the gas, the reaction being catalyzed by traces of nitric oxides.

The difficulty is eliminated by passing the raw gas through a catalytic reactor where nitrogen oxides are reduced by hydrogen contained in the gas to molecular nitrogen and water. Trace amounts of oxygen may also be reacted in this step. This catalytic treatment must be accomplished before any major compression of the gas is carried out.

The second consideration in the processing of raw gas is whether the raw gas should be compressed to a pressure slightly higher than 600 psig and then treated to remove sulfur or compressed to an intermediate pressure, desulfurized, and then compressed to the final delivery pressure. The latter method, since it minimizes compression of wet sour gas in relatively more expensive compressors, was selected for this design.

/1/ Patridge, F.J., "Production of Ammonia Synthesis Gas by Purification and Shift Conversion of Gas Produced from Coal", The Chemical Engineer, February 1980, page 88 - 91.

46



TABLE NO. 3.1B-4

BASE CASE ACID GAS REMOVAL

KOPPERS-TOTZEK GASIFIER

Raw Gas To Acid Gas Removal

<u>Comp</u>	<u>Vol%</u>	<u>M/Hr</u>	<u>lbs/Hr</u>	<u>Mw</u>
H ₂	27.56			
CO	61.08			
CO ₂	7.04			
CH ₄	0.09			
HCl	-			
N ₂ Ar	1.95			
H ₂ S	1.70			
COS	<u>0.12</u>			
	99.54	32,691	721,715	
H ₂ O	<u>0.46</u>	<u>151</u>	<u>2,720</u>	
	100.00	32,842	724,435	22.06

Pressure Psia 315

Temperature °F 120

Specification

1. Total sulfur in treated gas ppmv max 200
2. Minimum H₂S in acid gas vol% 20
3. Alternative sulfur specification ppm 1

Alternative Acid Gas Removal ApplicationsA. Pressure Variations

The product gas delivery pressure of 600 and 800 psig has no effect on the acid gas removal system since the acid gas removal operates at 300 psig and the product gas is simply boosted in pressure to 600 and 800 psig.

B. The Alternative Sulfur Specifications

Lowering the sulfur specification from 200 ppm in the direction of 1 ppm total sulfur changes the availability of treating processes, but does not modify the choice of the Selexol process for the application to 1 ppm of total sulfur provided a COS hydrolysis unit is included. The Sulfinol system should not be used below 3 to 5 ppm. Benfield may not have difficulty if all the sulfur compound were H_2S . However, with COS, even with a hydrolysis step, the unit may have difficulty attaining 1 ppm. Both Rectisol and Selexol plus COS hydrolysis can attain the low sulfur specification. In the absence of a clear economic advantage of one over the other, it would be best to follow the same choice at 1 ppm as was chosen for 200 ppm total sulfur, namely the Selexol process.

C. Variation in CO_2 & COS Content

There is no economic incentive to remove the few percent of CO_2 from K-T raw gas. The COS must be reduced by hydrolysis for Selexol and Benfield to meet a total sulfur specification of 1 ppm. The COS does not have to be hydrolyzed for Rectisol or Sulfinol.



3.1C DESCRIPTION OF ACID GAS REMOVAL PROCESSES

I. The Benfield Process

The Benfield process was developed by H.E. Benson and J.H. Field and is based upon hot potassium carbonate solution's ability to absorb and regenerate acid gases. The Benfield innovation utilizes additives which catalyzed the rate of absorption and decreased the foaming tendency of the solution during regeneration.

The advantages for Benfield are:

- o Use of a non-volatile, low cost solvent.
- o Low solubility of gas components in solvent.
- o High temperature of 220-250°F operation increases thermal efficiency of process and reduces steam requirements.
- o Hydrolysis of part of the COS
- o Wide application in both synthesis gas, refinery and natural gas.

Various systems of equipment arrangement can be used to reduce the H₂S sulfur in the product gas to below 1 ppm. Arrangements such as split flow, two stage and Hi-pure are some of the methods used to reduce H₂S to very low levels. The Hi-pure configuration can also be used to reduce CO₂ to very low levels. The hot potassium carbonate solution is slightly selective to H₂S even in the simple single stage configuration. However, a high degree of selectivity is possible with the Hi-pure configuration.



For both the Babcock and Wilcox and the BGC/Slagging Lurgi raw gas it is not necessary to apply any more sophisticated configuration than the split flow system. The reason for this application is the moderate gas purity of 200 ppm total sulfur and the lack of any selectivity requirement because the CO_2 concentration is comparatively low.

The disadvantages of the hot carbonate, Benfield process are those common to many "chemical" type absorption systems. The solution is not regenerated when strong acids such as HCl, HF, HCN enter with the raw gas. The solution is not sufficiently high in temperature to hydrolyze all COS and not sufficiently low in temperature to prevent formation of formates and oxylates. The solution also cannot handle tars and oils due to high foaming tendencies in the regeneration. Ammonia is partially absorbed and leaves with acid gas. In order to properly apply the Benfield acid gas removal to any coal conversion process, the raw gas must be water-washed to remove HCl, HCN and ammonia. This is normally accomplished in the venturi scrubber. In addition to water scrubbing, the raw gas from the Lurgi or in the BGC/Lurgi slagging gasifiers must be scrubbed to remove all tar, hydrocarbons and phenols. This would require oil-washing generally followed by activated carbon absorption to remove traces of oil vapors.

For the Benfield application to the Lurgi Dry Ash system, it is necessary to use two-stages of absorption plus COS hydrolysis because a large amount of COS must be removed from the system in addition to the H_2S removed.



It may also be necessary to use two stages of catalytic hydrolysis of COS, one before the Benfield and one between the first and second absorption towers.

Typical schematics for Benfield split flow and two-stage are shown in Figures 3.1C-1 and 3.1C-2.

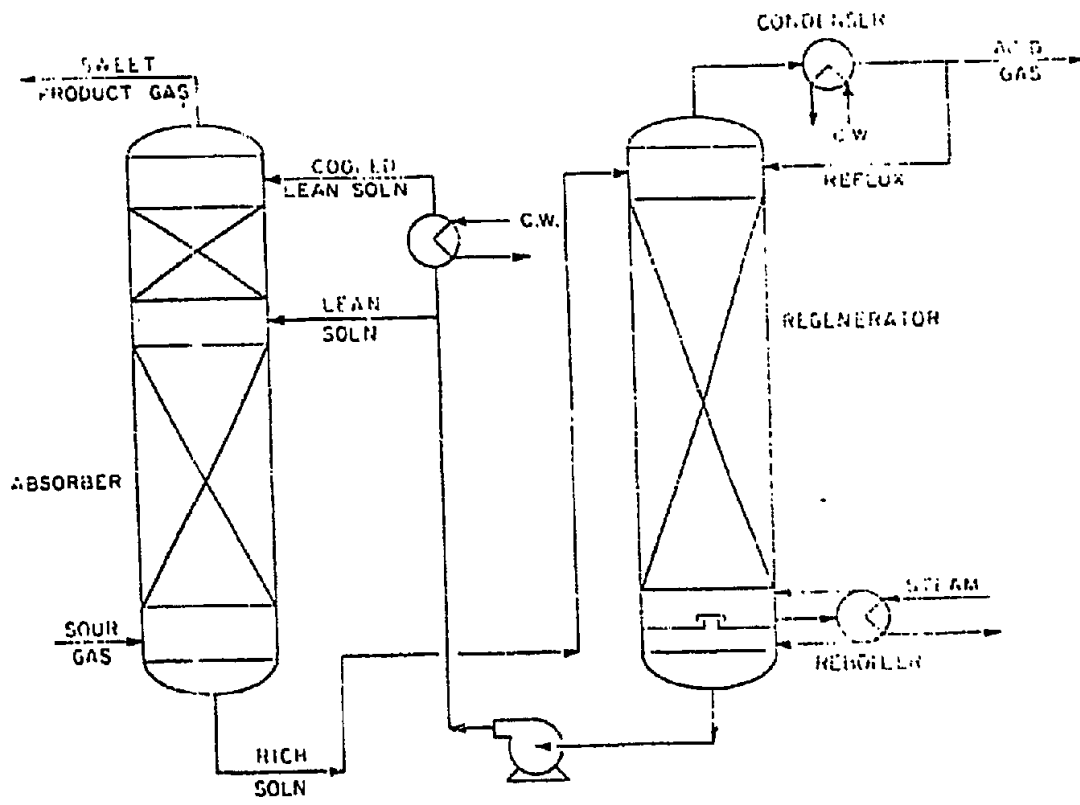


Fig. 3.1C-1 Illustration of Benfield split-stream design.

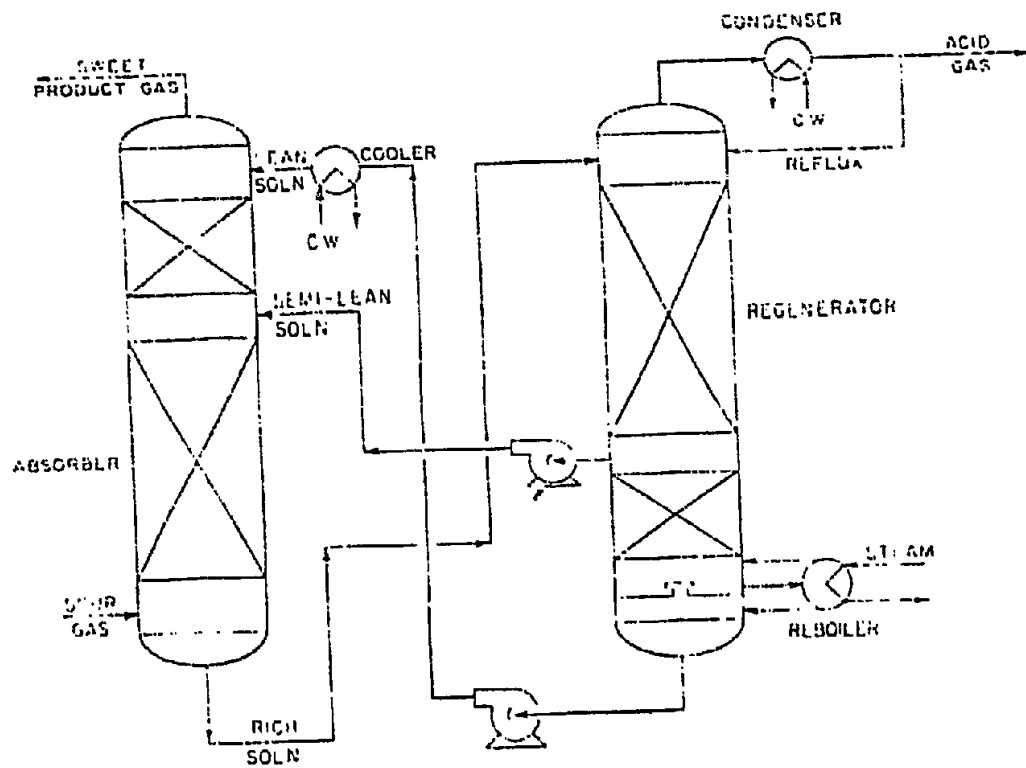


Fig. 3.1C-2 Illustration of Benfield two-stage design.



II. The Sulfinol Process

The Sulfinol process was developed by Shell Development and initially applied to natural gas treatment in 1963. Over a hundred applications exist covering both synthesis gas and natural gas applications.

The advantages cited for Sulfinol are:

- o Low solvent circulation rate
- o Low solution heat capacity
- o Low utility consumption
- o Low solution degradation, and low corrosion
- o High efficiency in COS, CS₂ and mercaptan removal
- o Low solvent vaporization losses
- o Freezing of solution without expanding

The principal disadvantages stated for the Sulfinol process are:

- o High absorption of hydrocarbons
- o High absorption of synthesis gas
- o High solvent costs
- o Expensive housekeeping
- o Impossible to meet low sulfur in the product gas because of the solvent having a sulfur compound, sulfolane.

The solution used in the Sulfinol process contains sulfolane (tetrahydro-thiophene dioxide), an alkanolamine usually DIPA and water. The concentrations are varied to some extent to suit the requirements. The components used indicate the solution has both



physical and chemical properties.

A Sulfinol flow sheet is shown schematically in Figure 3.1C-3. The operating pressure of 370 psig at the absorber forces an amount of hydrogen and carbon monoxide product gas into the solution. The absorbed gas is flashed at an intermediate pressure and returned to the absorber. The rich solution is sent to the regenerator tower and the acid is sent to the Claus unit.

The Sulfinol process may be applied to B&W gas without pretreatment because the B&W gas has no hydrocarbons except for a small percent of methane. The Lurgi type processes entrain too much tar, oil, phenols and aromatic naphtha, which would contaminate the Sulfinol solvent. Pretreatment by means of oil-wash, water-wash and activated carbon adsorption are required to reduce the oils to reasonable levels so as not to contaminate the Sulfinol solvent. With the protection of extensive cleanup for hydrocarbons, it is still necessary to use a complex reclaimer and to dispose of sludge or non-reclaimed mixtures of solution and impurities. This generally means a high makeup rate of an expensive solvent. The reason for degradation of the diisopropanolamine (DIPA) in the sulfolene solution is that it forms "oxazolidone" by reaction with CO_2 . Reclaiming with sodium carbonate partially recovers DIPA.

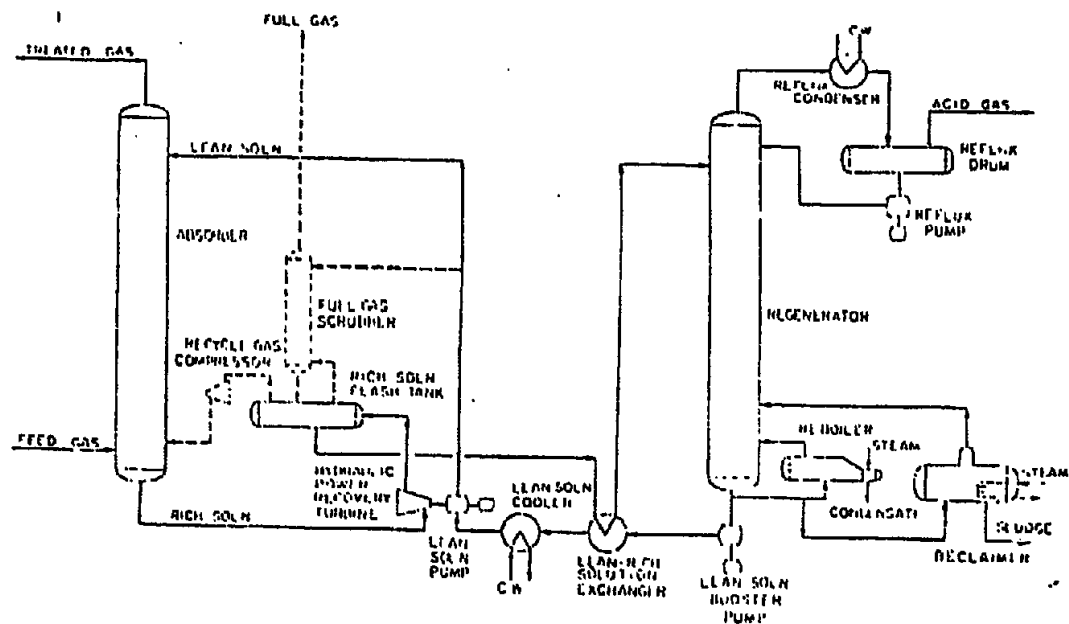


Fig. 3.1C-3 Typical Sulfinol flowsheet for high-pressure absorption.



III. The Selexol Process

The Selexol process, developed by Allied Chemical Corporation, involves selective physical absorption of such compounds as CO_2 , H_2S , COS, and mercaptans into the Selexol solvent (polyethylene glycol dimethyl ether) and subsequent regeneration of the solvent by low-pressure flashing and/or stripping with air, steam or flue gas. A flowsheet of a typical processing scheme is shown in Figure 3.1C-4.

Flowsheet Description

Absorber

Removal of the sour gas components in the feed gas occurs in an absorption column designed to accommodate recycle gas and lean and semi-lean solvent feeds. The solubility of the acid gas components in the solvent is approximately directly proportional to the partial pressure of the gas in the feedstream. Hydrogen sulfide is about eight times more soluble in Selexol than carbon dioxide, permitting preferential absorption of hydrogen sulfide. This can be compared to the relative solubility in water of 4 to 1 for hydrogen sulfide over carbon dioxide. The heat of solution of the acid gases in the Selexol causes a small temperature rise across the absorber; the rise is slight compared to the much greater rise experienced when chemical solvents, with their accompanying large heats of reactions, are used as absorption media. As an example, the

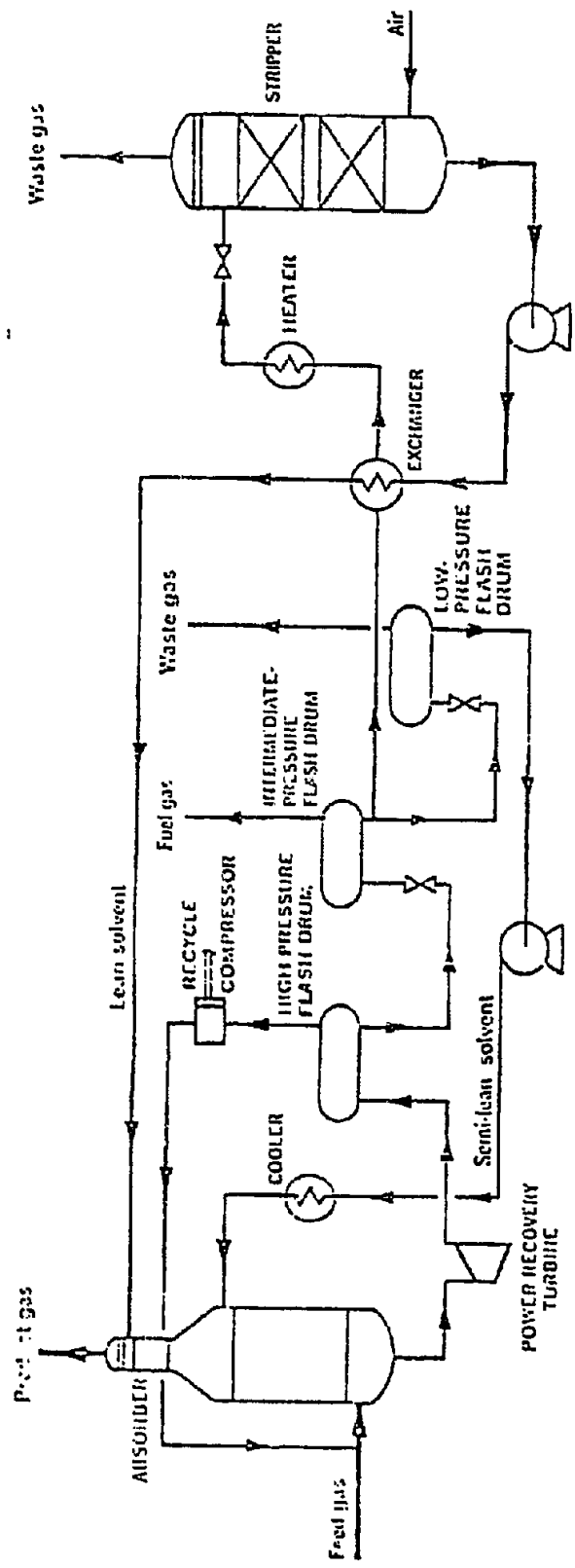


FIGURE 3.1C-4
 SCHEMATIC FLOW DIAGRAM FOR TYPICAL
 SELEXOL CLEAN-UP PROCESS



heat of solution of acid gases in Selexol due to physical absorption is approximately 25% of the heat reaction plus solution observed with a chemical absorbent like MEA. Selexol is non-corrosive so the material of construction of the absorber is carbon steel throughout.

Selexol Regeneration

Many variations of Selexol regeneration can be developed; an exploration of alternate schemes constitutes an important aspect at arriving at the best design.

The most usual scheme for processing the "loaded" Selexol exiting the absorber is to flash it at a number of successively lower pressure levels and strip the liquid product exiting the lowest pressure stage. The high pressure flash produces carbon dioxide plus a small quantity of hydrocarbons, if present in the feed. The flashed gas is compressed and recycled with the feed gas to the absorber to control methane solubility losses. Dissolved methane would normally be stripped out of the liquid solvent in the regenerator and would be lost in the Claus unit with the combustion of the acid gas. Gas recycle will also increase selectivity to H₂S removal and increase the carbon dioxide content of the product gas. The vapors from the intermediate stage flash vessel usually have sufficient fuel value to drive turbines or reciprocating engines. In passing through a turboexpander, the flashed gas from the intermediate pressure stage can be used to provide refrigeration to the plant to counteract the system heat inputs arising from pumping inefficiencies.



Part of the liquid stream from the intermediate-pressure vessel is flashed in the low pressure vessel. Waste gas is removed overhead, and is either vented or sent to a sulfur-recovery system. The liquid product (semi-lean solvent) of the low pressure stage is usually recycled to an intermediate tray of the absorber.

Another portion of the liquid product of the intermediate-pressure vessel is passed to a stripper. In the stripper, steam, or other stripping gas is used to drive-off waste gas and leave lean solvent. This lean solvent is recycled to the upper section of the absorber; the off-gas is either vented or processed further to sulfur.

Design Considerations

- o Relative Solubilities of $\text{CH}_3\text{SH}/\text{H}_2\text{S}/\text{COS}/\text{CO}_2/\text{CH}_4$

The setting of design operating conditions and selection of unit operation routes to effect an optimum design is largely dependent upon suitably manipulating the relative volatilities of the feed gas constituents in the presence of Selexol. In order of decreasing solubility within Selexol, the relative solubilities are: CH_3SH , H_2S , COS , CO_2 and CH_4 .

- o Temperature and Pressure

A decrease in temperature and an increase in pressure increases the solubilities of all components in Selexol. Temperature is a particularly significant design parameter. For example,



a decrease of only 10°F in temperature of the solvent at the bottom of the absorber produces a 15% drop in the amount of Selexol solvent required to effect a given acid gas absorption.

o Solvent Properties

The Selexol solvent, in addition to promoting a difference in relative acid gas solubilities by including solution non-idealities, also has favorable physical and chemical properties under absorption and regeneration operation conditions. It has a very low vapor pressure, minimizing solvent losses and also is chemically and thermally stable. Other features are that it is chemically non-reactive, non-foaming and non-corrosive and has a very low viscosity, minimizing pumping costs and also aiding mass and heat transfer. Its toxicity is low and its flash point, 304°F, is high enough to be safe for plant operating conditions. Its low specific heat keeps regeneration heat loads down and its capacity to absorb water helps dry the product gas, avoiding the need for additional drying equipment. The fact that Selexol does not contain any inorganic compounds is desirable since metal salts present in some clean-up solvents potentially can damage turbine blades .



o Energy Requirements

Selexol designs can be made essentially self-sufficient in pumping horsepower. Required energy can be supplied by flash gas via a turboexpander and hydraulic power via power-recovery turbines. Since the Selexol process improves in efficiency as the temperature is lowered some energy recoveries may be utilized to provide cooling effects.

o Stripping Gases

Several types of gases can be used as the stripping medium to desorb the acid gases and regenerate the Selexol solvent. These include air, steam, inert gas or fuel gas free of hydrogen sulfide. For maximum sulfur recovery, steam is used as the stripping medium.



o Effect of Carbonyl Sulfide

The common occurrence of carbonyl sulfide in gasifier effluents presents some special design considerations for the Selexol process. This is because carbonyl sulfide is intermediate in volatility between hydrogen sulfide and carbon dioxide in the presence of Selexol and therefore can become the limiting component in sulfur removal and in the production of a good Claus gas. This is particularly pertinent if more than 40% of the carbonyl sulfide has to be removed. Nevertheless it has been demonstrated ⁽⁹⁾ that under the proper design conditions Selexol can remove hydrogen sulfide and carbonyl sulfide to the ppm levels required to protect catalysts or eliminate sulfur pollution, and at the same time concentrate the hydrogen sulfide for use as a good Claus unit feed.

Economic Considerations

An economic evaluation of the Selexol process ⁽¹¹⁾ demonstrated that costs go through a minimum with regard to increases in absorber operating pressure and decrease as the temperature level decreases. Operating temperatures below 0°F are not practical because of increased solvent viscosity and the freezing point of Selexol, namely -20°F.

A cost comparison of Selexol with MEA and hot potassium carbonate clean-up processes on the same sour feed gas ^(10, 14) showed that the use of Selexol was particularly advantageous when comparing treating costs;



Selexol was one-third the cost of design routes using MEA, and one-half that of hot potassium carbonate processes.

Application to Coal Conversion Processing

Selexol is appropriate to handling the typical feed gas exiting from a coal gasifier, for example one containing 35% CO₂, 1% H₂S, and 50 ppm COS, with product gas specification of sulfur less than 1 ppm, vent gas to atmosphere, specified at less than 10 ppm H₂S, and with maximum H₂S (minimum CO₂) in the off-gas, so that the off-gas will be a suitable feed to a sulfur conversion process like a Claus plant or Stretford process. A less expensive installation would first hydrolyze the COS catalytically and then remove the more soluble H₂S.

IV. The Rectisol Process

The Rectisol process is a physical absorption process developed by Linde in Germany and licensed by Lotepro or by Lurgi. The process solvent is methanol which operates at a low temperature. The low temperature prevents the loss of valuable solvent by keeping the vapor pressure low and low temperature helps increase the solubility of H₂S and CO₂ in solution.

Various configurations of the Rectisol process may be operated to selectively remove H₂S, to remove H₂S with most of the CO₂ and to remove H₂S and CO₂ separately in relatively pure streams. Rectisol is capable of reducing the sulfur compounds to below 0.2 ppm.



The advantages of Rectisol are:

- o Capable of removing sulfur to very low levels
- o Highly selective to sulfur removal
- o Dehydrates gas at the same time as the sulfur is removed
- o Can produce relatively pure H_2S and separate CO_2 streams at the same time
- o Is not poisoned by salts, acids or cyanides
- o Can handle HCN, naphtha and oils removed from the gas
- o Has large background of experience in cleanup of raw gas from coal conversion, town gas and synthesis gas

The disadvantages of Rectisol process are:

- o Requires deep refrigerations with special refrigerants like propylene or the use of cascade refrigeration
- o Is expensive to operate
- o Is costly to construct

For the case of purifying gas to 200 ppm of total sulfur, the Rectisol system is slightly more expensive than others especially since selectivity is not required except for the Lurgi Dry Ash case. We have been informed by Lurgi that Rectisol is not a good application and that we should be looking at the use of Benfield, not Rectisol for all cases with 200 ppm total sulfur on the product gas. For the case of 1 ppm sulfur, the Rectisol process would be competitive and detailed information will be obtained from Lurgi.

RECTISOL PROCESS

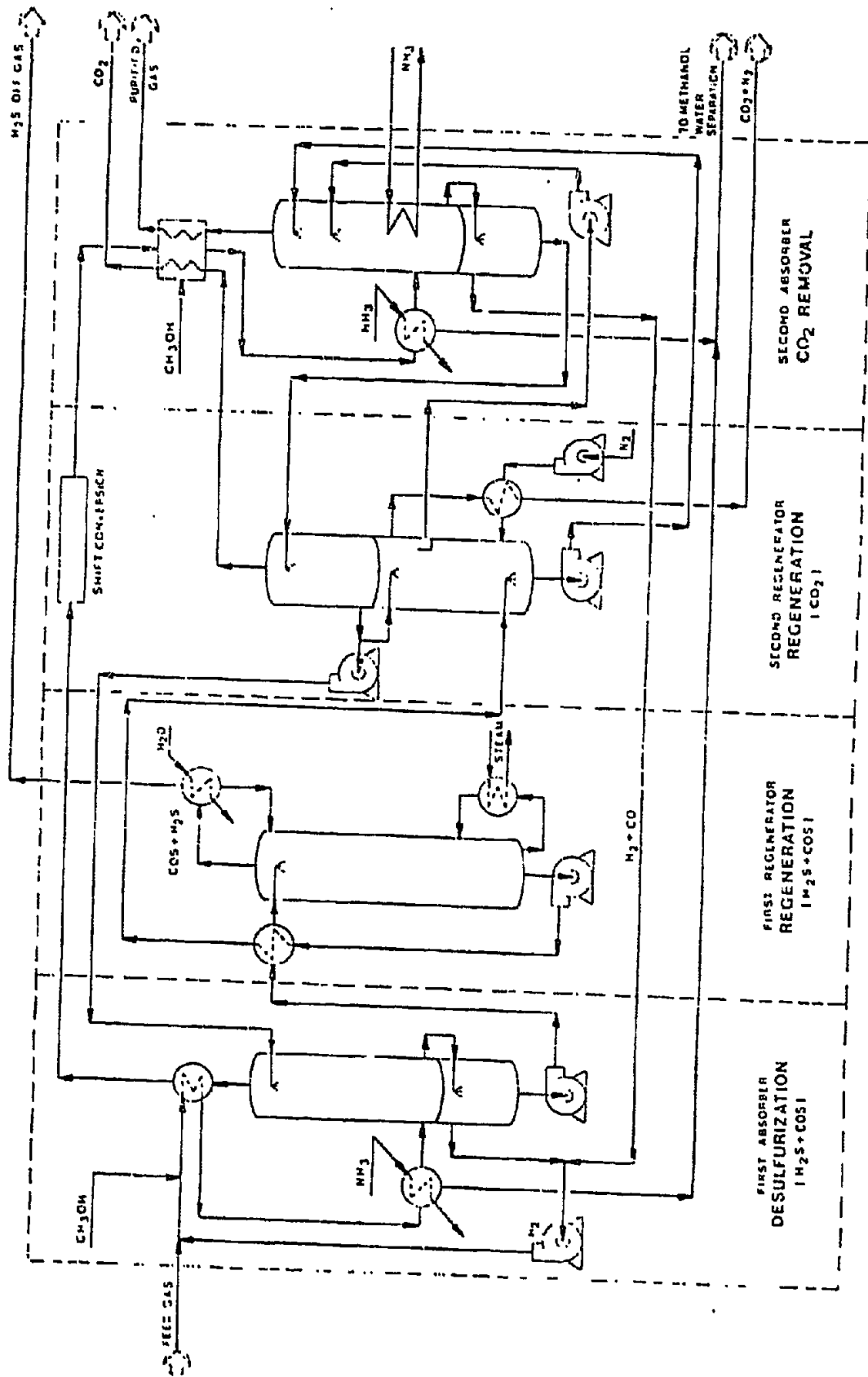


FIGURE 3.1C-5

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