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EVALUATION OF THE
BATTELLE AGGLOMERATING ASH BURNER PROCESS
FOR THE PRODUCTION OF MEDIUM BTU GAS

by

T. Y. CHANG
ALLAN S. WEST

Date Issued: June, 1979

Report Prepared by
SCIENTIFIC DESIGN COMPANY, INC.
Two Park Avenue
New York, New York 10016, U.S.A.

Relative to

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for the
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OAK RIDGE NATIONAL LABORATORY
Oak Ridge, Tennessee 37830
operated by
UNION CARBIDE CORPORATION
for the
U.S. DEPARTMENT OF ENERGY

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ORNL PERSPECTIVE

This report develops the cost of production of medium Btu gas via the Battelle Agglomerating Ash Burner Process (BAABP). For the purposes of this study a functional process was assumed. This assumed BAABP has to recommend it a high thermal efficiency and relatively low capital cost per unit of production. These factors stem primarily from multistage operation, which permits the use of air instead of relatively pure oxygen for combustion of carbon to generate heat and which isolates the products of combustion from the fuel gas. While process evaluations and comparisons are recognized to be inexact, the BAABP appears to have the potential for attainment of unexcelled thermal efficiency and unit production cost.

However, the development of the BAABP has been carried only through the laboratory and bench-scale testing phases; the process demonstration unit phase was terminated after experiencing a great deal of largely mechanical difficulty and before achieving integrated operation. The performance of the process demonstration unit was, at best, inconclusive. Hence, the assumption of a functional process has not been validated and is open to serious question. Both operability and reliability remain undemonstrated on the PDU scale and unattempted on a larger scale.

There is little doubt about the economic advantage that a satisfactorily developed BAABP would have over other processes for the production of synthesis gas or medium Btu gas. Likewise, there seems to be little doubt about the market demand for low cost synthesis/medium Btu gas. The central question is whether the problems that must be solved in order to achieve satisfactory operation are amenable to solution on the basis of what is currently known or might be found in the reasonably near future. Although a statement is made in the report to the effect that the process can be satisfactorily developed, this study is not an exhaustive examination of either the problems which remain to be solved or the developmental data that are available. Such an exhaustive examination appears to be the logical next step. We believe that the process, which has rather wide applicability and the apparent potential for both cost and fuel savings, merits this additional investigation.

ABSTRACT

The Battelle Agglomerating Ash Burner Process is in an early developmental stage. As a means for determining the magnitude of the incentive for expenditure of further development effort on this process, conceptual process design and associated cost estimates were prepared on the basis of anticipated but unproven, process and equipment performance. On the basis of the postulated performance, the estimated cost for the product gas, assuming private financing, is \$2.82/MM Btu. It is concluded that the process merits further development.

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I. INTRODUCTION AND SUMMARY

The Battelle Agglomerating Ash Burner Process has been under development since the early 1960's. In 1973 the Office of Coal Research (jointly with the American Gas Association for a short period of time) agreed to fund the further development of this technology for the production of high Btu gas. Various studies, including one by Scientific Design¹ (SD), have shown that the process is not economically attractive for the production of high Btu gas. Because of this and severe funding limitations the program was dropped from DOE's high Btu program fifteen months after mechanical completion of the Process Development Unit. A more suitable application of the Battelle Agglomerating Ash Burner Process might be the production of medium Btu gas. The prime purpose of this evaluation is to establish whether or not the process warrants further development efforts when the end-product is medium Btu gas.

Since the PDU never achieved integrated operation of the reaction system, only very limited process data are available. Because of this, a detailed review of the technical performance and operability of the process could not be made and SD had to assume what is thought to be a reasonable (although unproven) design for the gasification/combustion system. The assumed (or design basis) reaction system incorporates Battelle's latest thinking, plus suggestions and comments by SD. In conjunction with Battelle and DOE, SD prepared a Process Design Basis which, together with Battelle's postulated design, was used to prepare a conceptual design of a fully integrated facility to produce medium Btu gas from coal.

This report provides certain background information on the basic process and on the history of this particular study. Also provided is a description of the version of the process that was used as the basis for this study and cost estimate. Equipment descriptions and cost estimates are also provided. Certain questions arose during the course of the study concerning various facets of the process. The questions were transmitted to Battelle, whose response is included in the Appendix.

The estimated capital cost for a 250×10^9 Btu/day plant constituted and operated according to the projected design and yield slate is \$370,000,000 (mid-1978 dollars). Assuming private investor financing methods, the estimated cost of the gas is $\sim$ \$2.82/MM Btu.

SD's evaluation of the Agglomerating Ash Burner Process/Fluidized Bed Gasifier as conceived by Battelle is that the process can be developed to competitively produce medium Btu gas, but further study, including experiments in integrated operability, controlability, and performance, are still required. The scant qualitative experimental data available are not sufficient to prove that a production unit will perform in accordance with the assumptions used in this report; therefore, the economics developed must be viewed with some skepticism. Furthermore, certain potential process and mechanical problems exist that must be solved before a costly larger facility, such as a pilot or demonstration plant, can be justified. The functions requiring demonstration include:

- Feasibility of "raining" the hot ash agglomerates through the Gasifier.
- Adequate control of the Combustor reaction conditions (temperature, solid particle size, bed depth, and gas velocity) for steady self-agglomeration.

- Capability of the process to accept caking coals without the need for external pretreatment of the feed material.
- Demonstration of integrated operation of the Gasifier/Combustor system including the circulation of large quantities of hot ash.
- Discharge of most of the incoming sulfur as H_2S in the fuel gas from the Gasifier and a minor fraction as SO_2 in flue gas from the Combustor.
- High fractional conversion of steam and carbon in the Gasifier.
- Rate of production of minor components such as ammonia, phenols and other organics.
- Satisfactory mechanical performance of the Gasifier-Combustor complex, including the heat recovery system.

Despite the lack of data and potential process problems SD believes that economics of the Battelle process for medium Btu gas are extremely attractive and further study of the process is warranted.

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1. Scientific Design Company, Evaluation of the Battelle Agglomerating Ash Burner High Btu Coal Gasification Process, ERDA/010404, NTISDE, Oct., 1977
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II. BACKGROUND INFORMATION

The Battelle Agglomerating Ash Burner Process is a pressurized two-stage, fluidized-bed system involving the combustion of coal or char in one fluidized bed and the steam gasification of coal in a separate fluidized bed. The heat for the endothermic gasification reaction is provided for by the circulation of hot ash between the burner and the gasifier. It is claimed that the fluidized bed combustion step operates under temperature and velocity conditions such that the ash in the coal is agglomerated in a controlled way into free-flowing, inert solid pellets. These pellets or ash agglomerates serve as a direct contact heat-transfer medium in the gasifier.

The initial development of the Agglomerating Ash Burner Process was carried out by Battelle in the early 1960's under the sponsorship of the Union Carbide Corporation. That work included non-integrated bench scale studies of the agglomerating burner and gasifier. Other experiments were conducted to estimate turbine-blade erosion by the fly ash contained in the flue gas from the combustor. All of the work supported by Union Carbide was done at atmospheric pressure.

After several attempts, beginning in 1965 by Union Carbide and later by Battelle, to interest the Office of Coal Research (OCR) in the process, OCR and the American Gas Association (AGA) agreed in 1972 to jointly sponsor further development. Under this agreement, the development work was supervised by an operating

committee drawn from AGA and OCR and financed one-third by AGA and two-thirds by OCR. AGA withdrew its support in August 1974 and the work has since been supervised and financed entirely by OCR and more recently by ERDA and its successor DOE.

In January, 1973 OCR awarded Battelle a contract to design, construct and operate a Process Development Unit (PDU) which would ultimately achieve integrated operation of the burner and gasifier under pressure. The design of the PDU to be located in West Jefferson, Ohio began in January 1973 and was essentially completed in January 1975. Field construction began in June of 1974 and was successfully completed in mid-August, 1976.

During the last half of 1976 the plant was commissioned and solids circulation was established at low temperatures and pressures. Efforts during 1977 were directed at achieving integrated burner/gasifier operations but mechanical and solids transfer problems prevented this.

In February 1977 SD was awarded a contract to provide technical and economic evaluations of competing processes, concepts and systems for the production of high Btu gas from coal. As part of this work SD was asked to provide an economic evaluation of the Battelle Process to establish its relative attractiveness compared to the alternative high Btu processes studied in the CFBraun Report². The results of this evaluation showed that, for high Btu gas, the Battelle process was economically unattractive when compared to the other processes. Despite our earlier findings, the Battelle process appeared to have advantages when medium Btu gas is the desired product. This, combined with the continued industrial interest in the process, induced ERDA in September 1977, to request SD to perform a similar evaluation of the Battelle process for

the production of medium Btu gas. This new study was to include an analysis of PDU data which had been generated to date and that which would be produced during the course of the study. However, upon review with Battelle it became clear that the PDU program had generated very little useful data to that time; very shortly thereafter ERDA decided to mothball the unit effective March 31, 1978.

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2. C.F.Braun, Factored Estimates for Western Coal Concepts, (FE-2240-5), October, 1976.

III. DESIGN BASIS

BASIC ASSUMPTIONS

The basic assumptions used in this study generally follow the "Coal Gasification Commercial Concepts Gas Cost Guidelines" prepared by CFBraun in January 1976. The major "Guidelines" used in this evaluation are summarized below:

PLANT CAPACITY: The plant is designed to produce a medium BTU gas from Eastern bituminous coal by the Battelle Agglomerating Ash Burner Process. The fuel gas shall have a high heating value (HHV) of 250 billion Btu/Stream day and be compressed to 500 psig for distribution by pipeline to industrial users.

SERVICE FACTOR: Service factor of the plant is to be 90 percent (7884 Hrs/Yr).

PLANT SITE: The plant will be located in southern Illinois.

SITE CONDITIONS: The plant will be built on relatively level and dry land with soil load bearing capacity of 4,000 pounds per square foot.

SEISMIC CONDITIONS: The plant will be located in Seismic Zone 1.

CLIMATIC CONDITIONS: The plant design is to be based on the following climatic conditions.

Temperature, °F	
Summer dry bulb	95
Summer wet bulb	75
Winter dry bulb	5
Design for freeze protection	
°F with 30 MPH wind	-15
Design frost line, feet below surface	4
Wind speed (miles per hour)	
Average	4
Peak gust	113
Plant elevation, feet above sea level	650
Normal atmospheric pressure, psia	14.4

COAL ANALYSIS: Illinois No. 6 coal will be used as the feed material. Battelle have supplied the following analyses for this coal.

Proximate Analysis, as received

	<u>Wt. %</u>
Moisture	10.00
Volatile matter	32.31
Fixed carbon	46.84
Ash	10.85
	<u>100.00</u>

Ultimate Analysis (dry)

	<u>Wt. %</u>
Carbon	68.64
Hydrogen	4.59
Nitrogen	1.66
Oxygen	8.49
Sulfur	4.23
Ash	12.39
	<u>100.00</u>

Heating value of coal as received Btu/lb (HHV) = 11,277

Heating value of dry coal Btu/lb (HHV) = 12,530

COAL DELIVERY AND ASH DISPOSAL: The coal will be delivered to the plant in railroad cars. The ash will be cooled by a dry process and will be shipped back to the coal mines in the returning rail cars.

COAL STORAGE: Storage and handling facilities will be provided for the supply of coal to the plant. The coal storage piles shall have a capacity of 14 day supply based on normal consumption of the plant.

WATER: Raw water of suitable quality will be delivered to the plant battery limits.

WATER STORAGE: A water storage pond containing two days water needs is included.

PRODUCT GAS SPECIFICATIONS: The product fuel gas consists mainly of hydrogen and carbon monoxide with a small amount of carbon dioxide. The product is a medium Btu fuel gas with a high heating value (HHV) of over 300 Btu per standard cubic foot. The H_2S content will be low enough to meet the current EPA regulations.

BY-PRODUCTS: By products of the plant - ammonia, sulfur and anhydrous sodium sulfate will be sold for their commercial value.

EXPORT STEAM: Any excess steam produced in the plant will be exported as 1200 psig, 900°F superheated steam.

ENVIRONMENTAL REQUIREMENTS: The following environmental standards must be met:

Water Quality Standards - All water discharged to the sewer shall meet EPA regulations.

Solids Disposal - Ash from the coal gasification will be returned to the mines.

Flue Gas Discharge to the Atmosphere - The flue gas produced in the Combustor and vented to the atmosphere from the top of the citrate absorbers of the sulfur recovery plant shall have a sufficiently low SO₂ content to meet current EPA regulations. The solid particulate content of the vent gas shall not exceed 0.03 pound per million Btu (HHV) of the coal fed to the plant.

EQUIPMENT SPARING: The equipment sparing philosophy will be as follows:

- Compressors, expanders and blowers will not be spared.
- Pumps will be spared. The spare may be either a complete spare or a 50 percent spare in cases where two pumps are normally operating.
- Conveying equipment will not be spared
- A spare coal mill will be provided with a capacity equal to 50% of the total plant capacity.

UNIT COSTS AND PRICES: Unit costs and prices as of May 1, 1978 for a southern Illinois location are tabulated below:

Raw Water, Dollars/Thousand Gallons	\$ 0.40
Design Coal, Dollars/Million Btu as Received	1.00
Export Steam, Dollars/Million Btu equiv.	1.00
Sulfur, Dollars/Long Ton	25.00
Ammonia (Anhydrous), Dollars/Short Ton	170.00
Sodium Sulfate (anhydrous), Dollars/Short Ton	52.00
Operating Labor, Dollars/Hour	7.00
Construction Labor, Dollars/Hour	10.00
Ash Disposal, Dollars/Short Ton of Ash	1.00

GAS COST BY UTILITY FINANCING METHOD: The first year of gas cost and average gas cost over 20 years by the utility financing method are based on the following parameters:

- 20-year project life.
- 20-year straight line depreciation on plant investment, allowance for funds used during construction, and capitalized portion of startup costs.
- 75/25 equity ratio.
- Nine percent interest on debt.
- 15 percent return on equity.
- 48 percent federal income tax rate.

GAS COST BY PRIVATE INVESTOR FINANCING: The calculation of gas cost by the private investor financing method is based on the following parameters:

- 20-year project life.
- 16-year sum of the years digits depreciation on total investment.
- 100 percent equity capital.
- 12 percent DCF return rate.
- 48 percent Federal income tax rate

PROCESS DESIGN BASIS FOR THE GASIFIER/COMBUSTOR SYSTEM

Since the PDU was shutdown by the DOE before integrated operation was achieved very little data are available with which to reliably design a commercial facility. It was therefore agreed with the sponsors that an assumed design of the gasification/combustion system be used as the basis for the study. The following section discusses this assumed design which incorporates a number of very significant changes from the design used by SD in our original high Btu gas study.

A. Gasifier - Combustor Design

This evaluation utilizes Battelle's Gasifier design in which hot and fairly uniform sized particles of agglomerated ash are gravity fed from the Combustor to the top of the Gasifier. In the Gasifier the hot ash particles fall through a fluidized bed of char particles to supply the heat required for gasification. The segregated mode of operation is similar to that used in the CO₂ Acceptor pilot plant and has been successfully demonstrated in practice.

A more detailed discussion of the latest Battelle design of Gasifier and Combustor will be presented later in Section V - Process Description.

B. Coal Feeding and Split of Coal Sulfur Between Fuel and Flue Gases

In the present design all of the coal is fed to the Gasifier. This is an important improvement over the previous design in which approximately 18% of the coal was fed to the Combustor. This new design has two important advantages. The first advantage involves the sulfur in the coal which is either converted to H₂S in the Gasifier or burned to SO₂ in the Combustor. Because H₂S is much easier to remove than SO₂, it is advantageous to maximize the H₂S formation. Battelle thermobalance experiments show that 80%

of the sulfur in the coal can be converted to H_2S in the Gasifier. The partially converted coal (char) from the Gasifier is fed to the Combustor where the remaining 20% of the original sulfur is burned to SO_2 . This new split of coal sulfur will effect great savings in sulfur recovery as will be discussed later in Section IV - Processing Scheme. The second advantage of feeding all of the coal to the Gasifier is that the unconverted carbon in the Gasifier char can be almost completely consumed for a productive purpose in the Combustor. The utilization of the last fraction of coal carbon can be accomplished much more economically in the Combustor than in the Gasifier, because of the difference in reaction rates.

C. Design Conditions for the Gasifier and the Combustor

The following information was provided by Battelle:

Gasifier	Gas Outlet Pressure = 100 psig Gas Outlet Temperature = 1750°F
Combustor	Gas Outlet Pressure = 100 psig Gas Outlet Temperature = 2050°F

Inlet steam/carbon ratio in Gasifier = 1.0 lb/lb
 Gasifier carbon conversion = 62%
 Gasifier steam decomposition = 80%

Raw Fuel Gas Composition (approximate)

	<u>Mol. %</u>
H_2	51.0
CO	32.2
CO	3.5
CH_4	3.1
H_2	1.2
N_2	0.6
NH_3	(600 ppm)
H_2O	8.4
	<hr/> 100.00

Flue Gas Composition:

	<u>Mol %</u>
N ₂	78.05
CO ₂	19.52
O ₂	0.99
SO ₂	0.24
H ₂ O	1.20
	<u>100.00</u>

Notes on Above Raw Fuel and Fuel Gas Composition

- (1) Based on performance data from commercial gasifiers operating under similar temperatures, bed configurations, and mode of coal feeding, about 10% of the nitrogen in the coal appears in the fuel gas as ammonia with the balance evolved as nitrogen gas.
- (2) From the operation of similar gasification plants it is conservatively estimated that about 500 ppm by weight of coal appears in the fuel gas as phenol. No tar is present in the fuel gas because the coal is fed at the bottom of the Gasifier and any tar that may be formed will decompose in the hot fluidized bed during the upward passage of the gas.
- (3) Based on Battelle PDU data, 5% excess air is used in the burning of char in the Combustor.
- (4) The amount of CO in the flue gas from the Combustor is negligible, according to Battelle PDU performance.

IV. PROCESSING SCHEME

Illinois No. 6 Coal is being used to make a medium Btu fuel gas by the Battelle Agglomerating Ash Burner Process. The capacity of the plant is 250×10^9 Btu (HHV) per stream day and the fuel gas is delivered to the pipeline at 500 psig. The coal enters the plant in railroad cars and after grinding and screening only coal particles smaller than 6 mesh size are used as feed to the gasifiers. To insure that the coal is free-flowing it is dried in a flash drying system before being fed to the bottom of the gasifier by a proprietary injection system.

The Battelle designed Gasifier-Combustor system is used for the coal gasification. Circulating hot coal ash from the Combustor is used as the heat source for the endothermic reactions in the Gasifier. The hot ash particles fall through a fluidized char bed and the cooled ash is recycled from the bottom of the Gasifier to the Combustor where it is reheated by burning in air unconverted char; the heated ash moves by gravity back to the Gasifier.

Unlike most other coal gasification processes, the Battelle process produces two, instead of one, hot gas streams - a 1750°F raw fuel gas from the top of the Gasifier and a 2050°F flue gas from the top of the Combustor. As will be discussed later in this section, proper heat recovery from these two hot gas streams is extremely important to the commercial viability of the process.

Allied Chemical's Selexol process has been chosen to remove the H_2S from the compressed fuel gas. Selexol solvent is very selective in absorbing H_2S in preference to CO_2 and thus it is possible to recover the absorbed H_2S from the Selexol regenerator in a concentration high enough for use in Claus and Citrate sulfur recovery plants.

COAL PREPARATION AND FEEDING

Coal Milling

For grinding the feed coal from $1\frac{1}{4}$ " x 0" to minus 6 mesh size, cage mills are suitable. To minimize the formation of fines, closed circuit rather than once through milling has been adopted. In this arrangement acceptable product from the mills passes 6 mesh screens and oversize coal particles not passing the screens are recycled to the mills. With the once-through mode of operation, the milling would have to be more severe because the product must be under 6 mesh size. More severe milling would of course result in more fines.

Coal Drying

Coal dryers which are separate from the coal mills are used. While integrated milling and drying are widely practiced for fine pulverization of coal for boiler firing and other uses, it is believed that separate milling and drying is a more flexible and reliable operation to produce minus 6 mesh coal particles. Also, for coal of this size range, relatively inexpensive flash dryers can be used. The flash dryers chosen for this plant will be designed for once-through gas flow and will use hot air heated by recovery of waste heat from the process, aided by a trim heater which burns a small fraction of the desulfurized product fuel gas.

Coal Feeding

For bottom feeding of the Battelle Gasifiers, a coal injection system such as offered by Petrocarb, Inc appears to be best suited. Dried coal brought up to system pressure in alternately pressured lockhoppers and stored in feed tanks, is injected with recycle product gas in eight equal streams to each Gasifier. These eight streams enter the Gasifier through eight equally spaced and widely separated nozzles around the bottom periphery of the gasification zone. To control the distribution of the coal feeds the Petrocarb system employs a gas with a high dew point to inject the coal via the feed nozzles located below the feed tanks. The coal flows from the feed tanks in a dense phase flow fluidized transport mode. To introduce the coal particles to the Gasifier in dilute phase flow, highly superheated steam is injected into the eight coal feed streams just prior to their entry into the Gasifier.

Since coal is injected into the Gasifier as multiple streams of highly dispersed solids and is heated rapidly through the caking temperature range in a fluidized-bed region where the coal particles will be separated by moving ash and char particles, Battelle believes that the coal particles will not agglomerate in the bed and that, therefore, pretreatment is not necessary.

GASIFIER - COMBUSTOR AND HEAT RECOVERY SYSTEMS

From the coal and steam fed to the Gasifier and the char and air fed to the Combustor, the following three hot streams are produced:

- Raw fuel gas at 1750° from the top of the Gasifier
- Flue gas at 2050°F from the top of the Combustor
- Ash at about 1700° from the bottom of the Gasifier

The scheme to recover heat and energy from these three streams is as follows:

1. The heat recovery scheme maximizes the production of high-pressure, superheated steam from the hot fuel gas and flue gas streams. All steam is generated at 1200 psig, 950°F. The remaining heat in the fuel and flue gases leaving the high-pressure waste heat boilers is used to preheat boiler feedwater. The moderate amount of low-pressure steam required for the plant is supplied by extraction from the condensing turbines driving the air compressors for the Combustors. The two Gasifiers require a total of 640,000 lb/hr of process steam at 150 psig. This steam is supplied by expanding the 1200 psig, 900°F steam through non-condensing turbines which economically produce about 40,000 horsepower for the plant. This furnishes part of the power required for product gas compression. The turbine exhaust steam is superheated to about 1300°F by exchange with hot, raw fuel gas before being fed to the Gasifiers.
2. Cooled and clean raw fuel gas at 100°F is compressed to 520 psig for H₂S extraction by the Selexol process and delivery to the pipeline. The compression is done in two stages, with heat recovery by preheating boiler feedwater after the first and second stages of compression and also by preheating air for coal drying after the second stage of compression.

Cooled and clean flue gas at about 500°F and 105 psia is passed through an expander discharging the gas at 250°F and 17 psia. The power recovered by the expander is used to supply part of the power required by the air compressor, with the balance furnished by a condensing steam turbine. The heat remaining in the flue gas at 250°F is used to adiabatically saturate the gas with water vapor to make

it suitable for feed to a venturi scrubber and Citrate absorber.

3. The hot ash, which leaves the bottom of the Gasifier at about 1700°F, is cooled to 250°F by direct contact with air and is subsequently sent to disposal as a dry granular solid. The air used in cooling the ash is thus heated to 720°F and is routed for use in the coal dryers.

SULFUR RECOVERY

Because of the 80/20 split of the coal sulfur to form H_2S in the Gasifier and SO_2 in the Combustor, as mentioned previously in Section III - Design Basis, it is possible to use part of the H_2S extracted by the Selexol process to react with the SO_2 absorbed from the flue gas by the citrate solution. This liquid-phase Claus reaction converts absorbed SO_2 to elemental sulfur, provided sufficient H_2S is available for this reaction. The Selexol-extracted H_2S , which is present in excess of that required for the SO_2 recovered by the Citrate unit, is processed for sulfur recovery in a Claus unit. The Claus tail gas is treated for clean-up of residual sulfur compound in a Beavon-Stretford unit before venting to the atmosphere.

An additional advantage of using the Selexol process to extract the H_2S is that the product gas leaving the absorber is dried sufficiently by the Selexol solvent to meet pipeline specifications.

LIQUID EFFLUENT TREATMENT

Due to the bottom feeding of the coal to the Gasifiers, the raw fuel gas produced is free of tar and contains very little ammonia or organics. Based on the performance of similar commercial gasifiers, it is estimated that only about 10% of the nitrogen in the coal appears as ammonia in the fuel gas. The amount of phenols formed is conservatively estimated to be about 500 lb/hr for the whole plant.

Ammonia in the raw fuel gas would go into the water condensate during the cooling of the raw gas and would be accompanied by about an equi-molar quantity of H_2S . To minimize NH_3 and H_2S in the sour water condensate, U.S. Steel's Phosam Absorption System is used to remove NH_3 from the raw gas at about $190^{\circ}F$, before water vapor starts to condense. Absorbing ammonia at this relatively elevated temperature has another advantage in that most of the small amount of COS (About 10% of the quantity of H_2S) in the raw gas is hydrolyzed by water on the absorption trays to CO_2 and H_2S . If present in appreciable quantities, COS would cause some difficulty in the Selexol absorbers.

The sour water from the raw gas coolers is steam stripped to remove the small amounts of NH_3 and H_2S . The overhead vapor is piped to the Stretford Absorber in the sulfur recovery plant. The stripped water may contain up to 500 lb/hr of phenols which is removed by Amberlite resins in a packaged unit offered by the Rohm and Haas Company. The phenols in the absorption beds are eluted by methanol and then separated in a distillation column. The phenols are recovered as 72 wt.% water solution, which is used as fuel in the plant. The treated water, containing about 10 wt. ppm phenols, will be used as make-up for the cooling tower.

V. PROCESS DESCRIPTION

This section presents a description of the process including a description of flow, process flowsheets, summary material and energy balances, equipment lists and specifications of major equipment.

The overall process facility is divided into the following major units or sections.

<u>Flowsheet Drawing No.</u>	<u>Unit No.</u>	<u>Description</u>
1950-X-101	-	Block Flow Diagram
1950-X-102	11	Coal Storage & Reclaiming
1950-X-103	12 & 14	Coal Preparation & Coal Feeding
1950-X-104A	15	Gasifier, Combustor & Gasifier Heat Recovery
1950-X-104B	16A	Raw Gas Scrubbing & Ammonia Removal
1950-X-105	15B	Fuel Gas Energy Recovery, Air Compression, & Reject Ash Facilities
1950-X-106	23	Product Gas Compression, Drying and H ₂ S Removal
1950-X-107	16B	Ammonia Recovery
1950-X-108A	21A	Fuel Gas Cleaning & Sulfur Recovery (Citrate Plant)
1950-X-108B	21B	Sulfur Recovery (Claus & Beavon Plant)
1950-X-109	22	Solids Disposal
1950-X-110	20	Liquid Effluent Treating
1950-X-111	30	Steam, Condensate & Power Generation
1950-X-112	31	Plant Water System

The overall process scheme is illustrated in the block flow diagram, Drawing 1950-X-101, which shows the relationship between the various units. The equipment designations used on the flowsheets and the equipment lists are illustrated by the following:

MC-1510 A/F (2)
Gasifier Cyclones

MC = Cyclone Symbol

1510 = Cyclone in Unit 15

A/F = 6 Duplicate Cyclones in one train

(2) = 2 trains in parallel

DESCRIPTION OF FLOW

Drawing 1950-X-102 (Unit 11) - Coal Storage and Reclaiming

Coal laden railroad cars from the mines are dumped into the Coal Receiving Bin (FB-1110). Tramp iron is removed by a magnetic separator (PX-1110) and the coal is sampled and crushed for lab analysis in the Automatic Sampler Package (SX-1110). Coal is distributed to storage piles, with a total capacity of 14 days, by the Travelling Boomstacker (PX-1112).

A bucket wheel reclaimer (PX-1113) removes coal from the storage piles and after a secondary tramp iron removal (PX-1115), the coal is conveyed by the Collecting Conveyor (PC-1118) to Coal Preparation, Unit 12.

Drawing 1950-X-103 (Units 12 and 14) - Coal Preparation and Coal Feeding

The 1- $\frac{1}{4}$ " x 0" coal received by the conveyor (PC-1201) is stored in the Storage Bin (FB-1201) for metered distribution to two of the three Coal Cage Mills (SS-1201 A/C).

Cage mills were selected because they can reliably grind coal to the required size distribution. To reduce production of fines, the mills are designed to operate at relatively low conversion to minus 6 mesh, using a 6 mesh screen to separate the recycle (oversize) stream.

Coal from the mills is conveyed to the Screen Feed Distribution Bin (FB-1202) and is distributed to ten screens (MS-1201 A/J) for one cut separation. The oversize ($\sim$ 32% over 6 mesh) is recycled back to the mill feed. Screen under-size (minus 6 mesh) is conveyed to a Dryer Feed Distribution Bin (FB-1203) for metered distribution to four flash dryers in Coal Dryer Package (ME-1201).

Sized coal is dried from 10% to 5% moisture because the coal feeding system requires a dry, free flowing material for successful operation. After drying, the coal is transferred to the proprietary Petrocarb feeding system.

Dried coal is sampled during transfer, by an Automatic Sampler Package (SX-1201) and conveyed to Coal Feeding, Unit 14.

Dry coal from Unit 12 is conveyed to the prepared Coal Bin (FB-1401) which has a one hour surge capacity. The coal is then gravity fed to Lock Hoppers in the Petrocarb feed system (FF-1401) and pressurized with inert gas for discharging to the Primary Injectors below. After emptying, the lock hoppers are depressurized and another batch of coal is charged from the bin above. Duration of the cycle is about 20 minutes and is automatically controlled. From the Primary Injectors, which are continuously maintained at the required pressure by recycle product gas, the coal is pneumatically conveyed by recycle product gas in 8 equal streams to feed nozzles located on the Gasifier. Just prior to their entry into the Gasifier, the streams are further diluted by superheated steam to lessen the chance of coal caking.

Drawing 1950-X-104A (Unit 15A)-Gasifier, Combustor and Gasifier Heat Recovery

Coal and steam react in the Gasifier to produce a gas rich in hydrogen and carbon monoxide. This gas is subsequently cooled and processed to produce a medium Btu fuel gas.

Coal is injected into the fluidized bed of the Gasifier (DR-1510) by a special coal feed system as described in the previous section (Unit 14). The Gasifier is a pear-shaped refractory-lined vessel approximately 41 feet diameter at the top and 30 feet at the lower end. A fluidized bed of char is maintained in the reaction zone; heat for the reaction is supplied by a falling stream of hot ash particles which enter the reaction zone at 2050°F and leave at about 1700°F. A downward-moving bed of cooled ash is maintained in the lower cylindrical section of the Gasifier under conditions somewhat below incipient fluidization.

The hot ash is introduced to the Gasifier at eight (8) points symmetrically located at the top. Due to their high density and large particle size the ash particles fall through the fluidized char bed. The process conditions at the top of the Gasifier are 1750°F and 100 psig. The superficial gas velocity at the top is 2 feet per second. The design process conditions will result in a solids residence time of about 2-½ hours. Based on Battelle thermobalance experiments this residence time is adequate to achieve a carbon conversion of approximately 62%. The unconverted coal is removed from the side near the top of the fluidized char bed and transported by recycle flue gas to the bottom inlet pipe of the Combustor.

The ash stream from the reaction zone flows downward as a moving bed and is discharged from the bottom of the Gasifier. Steam is introduced at about 1300°F through a pipe grid located at the bottom of the moving ash bed for

even distribution throughout the cross section of the Gasifier in order to maintain a fluid, but unagitated condition in the ash bed. The ash discharged from the bottom of the Gasifier is divided into two streams. A small stream, the net reject ash from the Gasifier -Combustor loop, is discharged through the Reject Ash Tanks (FF-1513 A/B) to the dry ash disposal system described below under Drawing 1950-X-105. The much larger circulating ash stream, amounting to about 12 million pounds per hour, is transported by hot air to the bottom of the elevated Combustor. An air classifier built into the lowest point of the circulating ash lift pipe allows ash particles larger than 1/8" to drop out of the conveying air; these particles then collect in a small pot below the bottom of the vertical lift line and are subsequently conveyed to the ash disposal system with the reject ash.

The Gasifier effluent gas is passed through the Gasifier Cyclones (MC-1510A/F) to remove entrained char particles. The char particles are transferred to the Combustor by recycle flue gas and the gas leaving the cyclones flows to the heat recovery system to be described later.

The purpose of the Combustor (BF-1510) is to burn the unconverted char produced in the Gasifier to supply the heat required to raise the temperature of the circulating ash stream from approximately 1700°F to 2050°F. The Combustor is a vertical tapered cylindrical vessel, 26 feet in diameter at the top section and 12 feet in diameter at the bottom. The process conditions at the top are 2050°F and 100 psig. Oxidation of char occurs in the fluidized bed of ash. The enlarged top section serves as a classifier in which small ash agglomerates grow and settle to the bottom. The heated circulating ash stream leaves the Combustor at the bottom side of the classifying section and flows to the Ash Distributor Tank (FF-1510) located above the Gasifier. The hot flue gas from the top of the Combustor is passed through the Combustor

Cyclones (MC-1512 A/F) to remove entrained ash particles which are returned to the Combustor using flue gas as a transport medium.

The hot Gasifier effluent gas flows from the Gasifier Cyclones to the Gasifier Heat Recovery System (BE-1510). This is comprised of a package unit to recover heat while cooling the gas to 725°F and is designed to:

1. Heat the 150 psig exhaust steam from the non-condensing turbines from 415°F to about 1300°F before feeding to the Gasifiers, and
2. Generate 1250 psig steam from boiler feed water and superheat to 905°F.

This heat transfer service is severe from the standpoint of fouling, erosion and gas temperature. An even more severe service is the cooling of the combustor flue gases. Some experience is available with commercially available units used in a similar service. A manufacturer of such equipment sized and costed the heat exchangers specified herein for gasifier product and combustor flue gas service.

The effluent gases leaving the package heat recovery units flow to the Gasifier Multicyclones (MC-1511 A/J) to collect char particles which are transported by recycle flue gas to the Combustor.

The heat in the gas leaving the Multicyclones is recovered by preheating boiler feed water in the No. 1 Gas Cooler (ES-1510) and the No. 2 Gas Cooler (ES-1511), which cool the gas from 725°F to 188°F. This latter temperature is very close to the dew point of the raw fuel gas and is chosen in order to maximize the heat recovery from the fuel gas stream. The gases leaving the No.2 Gas cooler flow to the Venturi Scrubbing System will be described in the next section.

Drawing 1950-X-104B (Unit 16A) - Raw Gas Scrubbing and Ammonia Removal

The purpose of this section is to process raw product gas from the Gasifier to:

- Remove residual traces of dust
- Remove Ammonia
- Cool the gas to 100°F and condense out the unreacted steam.

The dust in the raw gas stream is removed by scrubbing with a circulating stream of water in Venturi Scrubber (DT-1610). The mixture of gas and water is separated in the Knock-out Tank. A small sidestream of the circulating slurry is bled off to a hydroclone (MC-1610) where the solids are removed. The effluent water stream is fed to an Oil Water Separator (FA-1610) to remove the small amount of oil that might be present and is pumped and combined with the solids stream from the Hydroclone. The combined effluent stream is fed to the Purge Stripper (DT-1612) where the small amount of dissolved NH_3 and H_2S is stripped off by live steam. The stripper bottom stream is passed through Slurry Coolers (ES-1611) on its way to the Solids Disposal System-Unit-22, while the stripper overhead vapor is returned to the Venturi Scrubber Knockout Tank.

The gas stream leaving the Venturi Scrubber at 188°F is then passed through the Phosam Absorber (DT-1611) to remove ammonia. Phosam is a proprietary process of US Steel and uses an ammonia - phosphoric acid solution to scrub the gas stream. The advantage of using the Phosam process is that ammonia can be removed without absorbing the acid gases - CO_2 and H_2S .

The gas stream leaving the Phosam Absorber is cooled by boiler feed water from 188°F to 162°F in No. 3 Gas Cooler (ES-1612) and is then further cooled from 162°F to 100°F Gas Cooler (ES-1613), using cooling water. Water vapor is condensed in the cooler and separated from the gas stream in the K.O. Tanks (FF-1610). The condensed sour water is discharged to the Liquid Effluent Treating Unit 20. The gas stream from the K.O. Tank flows to Product Gas Compression shown in Drawing 1950-X-106.

Drawing 1950-X-105 (Unit 15B) - Flue Gas Energy Recovery, Air Compression

And Reject Ash Facilities

The hot flue gas from the Combustor, which leaves the Combustor Cyclones (MC-1512A/F) at 2050°F, is processed through the Heat Recovery System (BE-1551). The flue gas is cooled to 760°F by giving up heat to preheated boiler feed-water to 1250 psig, 950°F superheated steam. The flue gas then heats the hot air from the Air Compressor (CC-1551) from 533°F to 700°F, which is the design temperature for the air used for burning the char in the Combustor. The flue gas leaves the Heat Recovery System at 610°F and is used to preheat the high-pressure boiler feedwater, and in so doing, is itself cooled to 498°F. The flue gas then passes through Multi-cyclones (MC-1551 A/F) to remove the fine ash particles in the gas. The temperature of 498°F is chosen such that the flue gas will not be lower than about 250°F after expansion from 105 PSIA to 17 PSIA in the Flue Gas Expander (CE-1551). At 250°F the flue gas is hot enough to prevent condensation of sulfuric acid containing moisture in the expanded flue gas. The Flue Gas Expander supplies part of the power required by the Air Compressor (CC-1551).

The flue gas leaving the expander is piped to the Citrate Sulfur Recovery Plant (Unit 21A) to be described under the section on Drawing 1950-X-108A. A small fraction of the flue gas is recycled by the Recycle Compressor (CC-1552) to the Gasifier-Combustor area for use in conveying hot char.

The air required for the Combustor is compressed from 14.4 psia to 140 psia in a two stage Air Compressor (CC-1551). The first stage is an axial compressor which compresses the air to 80 psia. This interstage air is cooled from 477°F to 342°F by preheating boiler feedwater before being compressed in the second-stage centrifugal compressor to 140 psia and a temperature of 533°F. This hot air is further heated by hot flue gas to 700°F as previously described.

The Air Compressor (CC-1551) is driven partly by the Flue Gas Expander (CE-1551) and partly by the Turbine Drive (CT-1551) which runs on 1200 psig and 900°F steam and exhausts to a condensing system at 2-½" Hg absolute.

The reject ash leaves the Gasifier at about 1700°F and is collected in lock hoppers and pneumatically conveyed by cold air to the Rotary Ash Cooler (ME-1561). The pneumatic conveyor is actually the first stage of a two stage ash cooling process. In this first stage the ash is cooled from about 1700°F to 750°F by co-current flow of cold air and hot ash. The 750°F ash is then cooled counter-currently by cold air to 250°F to facilitate ash disposal. The hot air from the pneumatic conveyor and from the Rotary Ash Cooler, after dust removal, are combined to give about 150,000 lb/hr per train of hot air at 720°F which is available for use in the Coal Dryer (ME-1201).

The cooled ash at 250°F is conveyed to the Cold Ash Storage Bin (FB-1561) located in an elevated structure straddling the railroad. The dry ash will be dumped by gravity to railroad cars on their return trip to the coal mines.

Drawing 1950-X-106 (Unit 23) - Product Gas Compression, Drying and H₂S Removal

The raw fuel gas from Unit 16A together with a small amount of recycled gas from the Selexol Flash Drum (FF-2304) are compressed to 535 psia in a two-stage compressor. The First Stage Product Gas Compressor (CC-2301) is driven by a noncondensing steam turbine (CT-2301), which expands 1200 psig steam to 150 psig for use as feed to the Gasifier. A condensing steam turbine (CT-2302) provides the power for the Second Stage Compressor (CC-2302). Part of the heat from the first and second-stage compression is used to preheat boiler feedwater, while part of the heat from the second-stage compression is used to preheat the air used in the drying of the feed coal. The raw fuel gas then passes through the Product Gas Aftercooler (ES-2306), where it

is cooled by cooling water to 100°F and its moisture content is greatly reduced.

In the Selexol Absorber (DT-2301) the H_2S is selectively absorbed by the circulating Selexol solvent and the moisture content of the raw gas is reduced to meet the pipeline specification of 11 pounds of water vapor per million standard cubic foot of fuel gas.

The rich solvent leaving the Selexol Absorber passes through the Selexol Expansion Turbine (GT-2301), which supplies about 60% of the power required for pumping the Selexol solvent from the bottom of the Selexol Regenerator to the top of the Absorber. The liquid and gases leaving the Selexol Expansion Turbine enter the Selexol Flash Drum (FF-2304). In order to minimize the loss of CO , H_2 and CH_4 , the vapor leaving the Flash Drum is recycled back to the suction of the Product Gas Compressor (CC-2301). The solvent leaving the Flash Drum is heated up in the Selexol Feed Bottom Exchanger (ES-2302) on its way to the Selexol Regenerator (DT-2302). The regeneration of the solvent is carried out by inert gas stripping, using vent gas from the Stretford sulfur recovery unit. The Regenerator is designed to produce an overhead gas containing about 35% H_2S , a concentration high enough for use in the sulfur recovery plant.

The Regenerator bottoms is heated in the Selexol Reboiler (ES-2304) by 50 psig steam. It is then pumped through the Selexol Feed Bottom Exchanger (ES-2302) and the Selexol Cooler (ES-2303) to the top of the Selexol Absorber.

The fuel gas leaving the top of the Selexol Absorber is the pipeline Product Gas. Three small streams of fuel gas are diverted from the product stream for the following uses:

1. Injection of coal from the Petrocarb feed system to the Gasifier.
2. Reduction of the SO_2 in the Claus tail gas to H_2S for the Beavon Plant.
3. Use in a trim heater to boost the temperature of the air that has been preheated by the hot ash and the Product Gas Compressor discharge to the level required for coal drying.

Drawing 1950-X-107 (Unit 16B) - Ammonia Recovery

The Phosam process for ammonia recovery is a proprietary process of US Steel. However, there is adequate published information available to permit the conceptual design of the process.

The ammonia in the fuel gas stream is removed by the absorption into a lean solution of ammonia-phosphoric acid in the Phosam Absorbers (DT-1611) shown on Drawing 1950-X-104B. The rich phosam solution leaving the bottom of the absorbers is pumped through the Phosam Solution Exchanger (ES-1650) to raise the temperature of the rich solution and, at the same time, to reduce that of the lean solution going to the absorbers. The rich Phosam solution then flows through a Contactor (FF-1655) where the light components H_2S and CO_2 , are flashed off and recycled to the Phosam Absorbers. After the Contactor the rich solution flows through the Phosam Stripper Primary Condenser (ES-1652) to the top of the Phosam Stripper (DT-1651). In the Phosam Stripper, live steam is added and the column bottoms are also reboiled by indirect steam heating to drive off a vapor mixture of water and ammonia. The water-ammonia vapors from the top of the Phosam Stripper are partly condensed in the Primary Condenser (ES-1652) by heat exchange with the incoming rich Phosam solution. The water-ammonia vapor mixture is completely condensed in the Secondary Condenser (ES-1653) by heat exchange with the incoming rich Phosam solution. The water-ammonia vapor mixture is completely condensed in the Secondary Condenser (ES-1653) using cooling water prior to collection in the Aqua

Ammonia Tank (FF-1652). The lean Phosam solution leaves the bottom of the stripper and flows to the Phosam Absorbers. It is cooled in Phosam Solution Exchanger (ES-1650) and Phosam Solution Cooler (ES-1651) before entering the top of the Phosam Absorbers which are located downstream of the raw gas Venturi Scrubbers (DT-1610) shown in Drawing 1950-X-104B.

The aqua-ammonia stream from the top of the Phosam Stripper is fractionated in an Ammonia Fractionating Column (DT-1653). Ammonia is taken off overhead, condensed in the Ammonia Condenser (ES-1655) and collected in Ammonia Reflux Tank (FF-1656). The ammonia product is stored in the Product Ammonia Tanks (FF-1653 and 1654). Water in the aqua-ammonia feed plus that condensed from the live steam injected at the bottom of the stripper is accumulated at the bottom of the column and discharged to the Sour Water Stripper (DT-2010).

Phosphoric acid is periodically added to the circulating Phosam solution to maintain the required concentration. The process is carefully controlled to maintain the water balance in the system.

Drawing 1650-X-108A (Unit 21A) - Flue Gas Cleaning & Sulfur Recovery (Citrate Plant)

The flue gas leaving the expander (CE-1551) at 17 psia and 250°F is cleaned in a venturi scrubber to remove the last traces of solid particulates before being treated in the Citrate Absorber. For both the scrubbing and the Citrate Absorber, it is necessary to saturate the flue gas. This is done in the gas saturator before the venturi scrubber by spraying a part of the plant blowdown stream from the Cooling Water Pumps (GP-3105). After leaving the scrubber, the flue gas enters the absorber in which 90% of the inlet SO₂ is absorbed in the re-circulating citrate solution before being vented from the top of the absorber. The SO₂ absorbed in the citrate solution is reacted with the H₂S

in the off gas from the Selexol Regenerator (DT-2302) in the agitated Citrate Reactor. The unreacted gas vents from the reactor and is sent to the Claus Plant, Drawing 1950-X-108B, for further sulfur recovery. The sulfur produced in the agitated reactor is melted and sent to a molten sulfur storage unit. The molten sulfur from the Citrate Plant, together with that from the Claus-Beavon Plant (Drawing 1950-X0108B), are stored in the same sulfur storage pit. It can be sold either as molten sulfur or as pellets made by a package pellitizer.

Some of the sulfur entering the Citrate Plant is oxidized to the sulfate form. A purge stream is processed in a crystallizer to produce Glaubers salt which is then dehydrated in a fluid-bed to anhydrous sodium sulfate, a by-product of the plant.

The steam condensates produced in the Citrate and Claus-Beavon Plants are collected in the Condensate Tank (FF-2101) and pumped by the L. P. Condensate Transfer Pump (GP-2101) to the Deaerator (ED-3001) in Drawing 1950-X-111.

Drawing 1950-X-108B (Unit 21B) - Sulfur Recovery (Claus and Beavon Plant)

The vent gas from the Citrate Reactor, consisting of mainly CO_2 and unreacted excess H_2S , is sent to the Claus furnace. Here about two thirds of the gas is by-passed while the balance is burned in the furnace with air to convert the H_2S to SO_2 . After heat recovery, the furnace outlet gas is mixed with the by-passed gas and the mixture is processed in a two-stage Claus converter, with the sulfur recovered in the condensers as molten sulfur. The condensers are designed to produce 50 psig steam. The molten sulfur is collected and transferred to the storage pit shown in Drawing 1950-X-108A.

The tail gas from the Claus plant, containing inerts and the un-reacted H_2S and SO_2 , is treated in the Beavon plant. A small stream of the product fuel gas is used to preheat the Claus tail gas and reduce the SO_2 in the gas to H_2S in the Beavon reactor. The reactor effluent gas is cooled in a waste heat boiler to produce 50 psig steam and then further cooled by cooling water before being fed to the Stretford. Here the H_2S in the gas is reduced to less than 10 ppm by volume by absorption in Stretford solution. In this process the cooled, hydrogenated tail gas is contacted by aqueous sodium carbonate solution where the hydrogen sulfide is absorbed and sodium hydrosulfide is formed. The solution also contains soluble pentavalent vanadium complexes which oxidize the hydro-sulfide ion to elemental sulfur while being itself reduced to a quadrivalent state. The absorber liquid effluent flows to an oxidizer where air is blown into the stream, oxidizing the vanadium to its original pentavalent stage, aided by disodium anthraquinone disulfonate. The oxidized solution is cooled in a cooling tower and then recycled to the absorber.

Elemental sulfur, finely divided, is floated to the solution surface and removed as heavy froth from which the sulfur is recovered by melting. The molten sulfur flows to the storage pit shown in Drawing 1950-X-108A. Part of the vent gas which leaves the top of the Stretford absorber is used as the source of inert gas for the whole plant. This gas consists of mainly carbon dioxide and nitrogen. It is used as stripping gas for the Selexol regenerator and also used for the lock hoppers in the coal feeding system after being compressed to 300 PSIG by the Inert Gas Compressor (CC-2161).

Drawing 1950-X-109 (Unit 22) - Solids Disposal

The purpose of the Solids Disposal Unit is to treat all the solids-bearing slurry streams and produce a solid cake for recycle to the coal gasification

section. The bulk of the ash from coal gasification is, however, disposed of separately and in the dry state. See description of flow for Drawing 1950-X-105.

Slurry streams from the Coal Dust Collector, Gasifier Venturi Scrubber, and Fuel Gas Venturi Scrubber are processed in a packaged Filter (MF-2211) for recovery of the fine particles of coal, char and ash to be recycled to the Coal Feed Injection System (FF-1401) shown in Drawing 1950-X-103. The water from the filters is discharged as waste water.

Drawing 1950-X-110 (Unit 20) - Liquid Effluent Treating

The sour water effluent from the Gasifier section is treated in this section to remove ammonia and hydrogen sulfide in the Sour Water Stripper (DT-2010). The small amount of phenols in the stripped water is removed by absorption with Amberlite resin as manufactured by Rohm and Haas Co.

In the Sour Water Stripper (DT-2010) 50 psig steam is introduced to the base of the stripper to drive out the volatile matter. Steam in the overhead vapor is condensed in the Stripper Overhead Condenser (ES-2010) and refluxed to the top of the stripper. The sour water feed is heat exchanged with the bottoms stream in the Stripper Feed - Bottoms Exchanger (ES-2011) to conserve heat and introduced to the Stripper at the top tray. Hot water from the bottom of the Ammonia Fractionating Column (DT-1613) is also fed to the Sour Water Stripper to recover the heat in this stream and to strip out the small amount of contained ammonia. The Stripper bottoms stream leaving ES-2011 is further cooled by the Stripper Bottoms Cooler (ES-2012) and then flows to the Waste Water Surge Tank (FF-2011).

Stripped sour water from the Waste Water Surge Tank is treated in a packaged phenol removal unit, such as supplied by Rohm and Haas Co. In this unit the water passes through Amberlite resin absorption beds which reduce the phenol concentrations from about 3000 ppm down to about 10 ppm by weight. The treated water is used for cooling tower make-up. The absorbed phenol is elutriated from the resin bed by methanol and the phenol-methanol is then distilled in a column in which the methanol is recovered as overheads and recycled to the absorption beds. The phenol is recovered at the bottom at 72% concentration and is used as fuel in the dryer used to convert the Glaubers salt recovered in the Citrate and the Stretford Units to anhydrous sodium sulfate.

Drawing 1950-X-111 (Unit 30) - Steam, Condensate and Power Generation

This section describes the steam, condensate and power generation facilities.

Boiler feed water is pumped by the HP BFW Pumps (GP-3004) from the Deaerator (ED-3001) to the Gasifier and Flue Gas Heat Recovery systems where high pressure steam is generated at 1250 psig and 905°F. This high pressure superheated steam is distributed as follows:

1. Approximately 320,000 lb/hr per train is fed to a non-condensing turbine (CT-2301) exhausting at 165 psia and 400°F. The exhaust steam is then sent to the Superheater section of the Gasifier Heat Recovery system where it is superheated to about 1300°F and used to feed the Gasifier (DR-1510).
2. Approximately 172,000 lb/hr is supplied to the Condensing Turbine (NG-3001) drive of the electric generator which furnishes the power requirements of the plant (approximately 23,000 KW).
3. Approximately 182,000 lb/hr per train is fed to the Condensing Turbine Drive of the Second Stage Product Gas Compressor (CT-2302).

4. Approximately 363,000 lb/hr per train is fed to the Condensing Turbine Drive (CT-1551) of the Air Compressor. The moderate plant requirements for low-pressure steam are provided by extraction of 275 and 50 psig steam from this turbine.

The sulfur recovery plants have a net production of about 51,000 lb/hr of 50 psig steam. This is less than the requirements by about 27,000 lb/hr, and the deficiency is made up by extracting steam from the turbine drive for the Air Compressor.

About 2 million pounds per hour of boiler feedwater are required for the 1200 psig heat recovery boilers in the plant. Of this, about 60% is recycled steam condensate, the balance being made up by demineralized water. The bulk of the steam condensate comes from the three condensing steam turbine drives for drives for the Air Compressors, Product Gas Compressors and the electrical turbo-generator. These condensates are pumped from their respective condensers at 109°F and are preheated in the heat recovery train for the Gasifiers. The heating is done, in succession, in the No. 3 Gas Cooler (ES-1612) to 170°F and in the No. 2 Gas Cooler (ES-1511) to 274°F. The condensate then enters the Deaerator (ED-3001) to be further heated to 300°F.

Demineralized water at 75°F is used as make-up to the steam system. Part of the demineralized water is preheated in Product Gas Intercooler (ES-2301) to 208°F and the balance in Blowdown Cooler ES-3004) to 230°F. The heated demineralized waters are fed to the Deaerator (ED-3001).

The blowdowns from both the Gasifier and the Combustor Waste Heat Boilers are fed to the Boiler Blowdown Tank (FF-3001). The flashed steam flows to the Deaerator to heat the condensate and demineralized water to 300°F. The hot water from the blowdown tank is cooled to 100°F in the Blowdown Cooler (ES-3004) by part of the demineralized make-up water as described above. The cooled boiler blowdown water is used as make-up for the cooling tower.

A small fraction (74,000 lb/hr per train) of the 300°F deaerator feedwater is pumped to the sulfur recovery plants for use in the LP Waste Heat Boilers (molten sulfur condensers). The steam condensates from the sulfur recovery plants, together with those from the Phenol Removal Unit (PR-2010) and the Selexol Reboilers (ES-2304) are pumped back to the Deaerator.

Drawing 1950-X-112 (Unit 31) - Plant Water System

The plant Water System comprises the following:

- Water Reservoir
- Fire Water Pumps
- Fire Water Jockey Pump
- Filter Feed Pumps
- Anthracite Pressure Filters
- Demineralizer
- Demineralized Water Tank
- Demineralized Water Pumps
- Cooling Tower Chemical Feed System
- Waste Neutralizing Sump and Pump
- Potable Water System

The Raw Water make-up to the plant is added partly to the Cooling Tower reservoir (UC-3130) and partly to the Water Reservoir (FA-3101). From the Water Reservoir the water is pumped through the Anthracite Pressure Filters (MF-3101) and the Demineralizer Beds (UD-3103) to the Demineralized Water Tank (FI-3101). The flow requirement of Demineralized Water is 1620 gpm. Periodically individual Anthracite Pressure Filters are backwashed to

remove the collected foreign matter. The Demineralizer beds are periodically treated with dilute caustic and sulfuric acid solutions to regenerate the cation and anion resins.

The cooling tower is designed to cool 130,000 gpm cooling water from the turbine condensers from 100°F to 85°F as well as 40,000 gpm from process users from 108°F to 85°F. A portion of the circulating cooling water is purged from the system to avoid excessive buildup of dissolved solids in the system. Fresh raw water is added to make up the losses. The cooling water is treated with sodium metaphosphate to avoid corrosion and scale formation in the cooling water system.

BATTELLE AGGLOMERATING ASH BURNER PROCESS

SUMMARY MATERIAL BALANCE FOR MAJOR STREAMS

TABLE V-1

<u>INLET STREAMS</u>	<u>LBS/HR</u>	<u>PERCENT OF TOTAL</u>
Coal, Dry	1,011,300	15.1
Water in Coal	53,200	0.8
Combustion Air, Nitrogen	2,467,200	36.7
Combustion Air, Oxygen	749,500	11.2
Combustion Air, Water	24,300	0.4
Raw Water	<u>2,405,000</u>	<u>35.8</u>
Total	6,710,500	100.0
 <u>OUTLET STREAMS</u>		
Product Gas	1,102,700	16.4
CO ₂ Vent	20,000	0.3
Cooling Tower Losses	1,545,000	23.0
Flue Gas Vent	3,594,000	53.6
Ammonia	1,000	Nil
Sulfur	41,400	0.6
Reject Ash	125,300	1.9
Excess Steam	15,700	0.2
Water Losses	<u>265,400</u>	<u>4.0</u>
Total	6,710,500	100.0

BATTELLE AGGLOMERATING ASH BURNER PROCESS

SUMMARY ENERGY BALANCE

TABLE V-2

<u>ENERGY INPUT</u>	<u>MM BTU/HR</u>	<u>PERCENT OF TOTAL</u>
Coal to Process, HHV	12,670	100.0
Total Input	<u>12,670</u>	<u>100.0</u>
<u>ENERGY DISTRIBUTION</u>		
Product Gas, HHV	10,420	82.2
By-Products, HHV		
Ammonia	10	Nil
Sulfur	160	1.3
Excess Steam, HHV	20	<u>0.2</u>
Subtotal, Product, By-Products & Excess Steam	10,610	83.7
Consumption and Losses	<u>2,060</u>	<u>16.3</u>
Total Distribution	12,670	100.0
<u>COLD GAS EFFICIENCY, PERCENT</u>		<u>82.2</u>
<u>PLANT THERMAL EFFICIENCY, PERCENT</u>		<u>83.7</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

MAJOR EQUIPMENT LIST

Dwg. 1950-X-102 (Unit 11) - Coal Storage & Reclaiming

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
FB-1110	1	Coal Receiving Bin
GP-1110A/B	2	Storage Area Sump Pumps
MS-1110	1	Sampler Guard Electromagnetic Separator
MS-1111	1	Mill Guard " "
PC-1110	1	Coal Receiving Conveyor
PC-1111	1	Received Coal Transfer Conveyor
PC-1112	1	Calibration Chain and Reel
PC-1113	1	Belt Scale
PC-1114	1	Cross Conveyor #1
PC-1115	1	Cross Conveyor #2
PC-1116	1	Cross Conveyor #3
PC-1117A/B	2	Bedding & Reclaim Conveyors
PC-1118	1	Collecting & Transfer Conveyor
PS-1110	1	Diverting Chute
PX-1110	1	Sampler Guard Electric Hoist and Monorail
PX-1111A/B	2	Belt Trailers
PX-1112	1	Traveling Boomstacker
PX-1113	1	Bucket Wheel Reclaimer
PX-1114	1	Reclaimer & Stacker Transfer Car
PX-1115	1	Mill Guard Electric Hoist and Monorail
SX-1110	1	Automatic Sampler Package
UT-1110	1	Rotary Car Dumper

MAJOR EQUIPMENT LIST

Dwg. 1950-X-103 (Unit 12 & 14)-Coal Preparation & Coal Feeding

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
FA-1201	1	Cleanup Sump
FB-1201	1	Mill Feed Storage Bin
FB-1202	1	Screen Feed Distribution Bin
FB-1203	1	Dryer Feed Distribution Bin
FB-1401	1	Prepared Coal Bin
FF-1401A/B(2)	4	Coal Feed Package Injection System
GP-1201A/B	2	Cleanup Sump Pumps
ME-1201	4	Coal Dryer Package
MC-1201	1	Coal Dust Scrubber Package
MS-1201A/J	10	Screens (Rotex - Vibrating)
PC-1201	1	Mill Storage Bin Feed Conveyor
PC-1202A/C	3	Mill Bin Discharge Belt Scales
PC-1203A/C	3	Mill Feed Conveyors
PC-1204	1	Collecting & Transfer Conveyor to Screen Feed Bin
PC-1205A/B	2	Screen Oversize Collecting Conveyors
PC-1206A/D	4	Screen Oversize Transfer Conveyors
PC-1207A/B	2	Screen Undersize Collecting Conveyors
PC-1208	1	Dryer Bin Feed Conveyor
PC-1209A/D	4	Dryer Feed Belt Scales
PC-1210A/D	4	Dryer Feed Conveyors
PC-1211A/D	4	Dryer Product Conveyors
PC-1212	1	Dry Coal Collecting Conveyor
PC-1401	1	Prepared Coal Bin Feed Conveyor
PS-1201A/C	3	Mill Feed Chutes
PS-1202A/C	3	Mill Discharge Chutes
PS-1203A/J	10	Screen Feed Chutes
PS-1204A/J	10	Screen Oversize Chutes
PS-1205A/J	10	Screen Undersize Chutes
PV-1201A/C	3	Mill Bin Discharge Vibratory Feeders
PV-1202A/J	10	Screen Bin Discharge Vibratory Feeders
PV-1203A/D	4	Dryer Bin Discharge Vibratory Feeders
SS-1201A/C	3	Coal Cage Mills
SX-1201	1	Automatic Sampler Package

MAJOR EQUIPMENT LIST

Dwg. 1950-X-104A (Unit 15A)- Gasifier, Combustor & Gasifier Heat Recovery

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
BE-1510(2)	2	Gasifier Heat Recovery System
BF-1510(2)	2	Combustor
DR-1510(2)	2	Gasifier
ES-1510(2)	2	No. 1 Gas Cooler
ES-1511(2)	2	No. 2 Gas Cooler
FF-1510(2)	2	Ash Distribution Tank
FF-1511(2)	2	Combustor Cyclone Collection Tank
FF-1512(2)	2	Multicyclone Char Collection Tank
FF-1513A/B(2)	4	Reject Ash Tanks
FF-1514(2)	2	Gasifier Cyclone Collection Tank
FF-1515(2)	2	Gasifier Recovered Char Collection Tank
FF-1516(2)	2	Oversize Ash Tank
MC-1510A/F(2)	12	Gasifier Cyclones
MC-1511A/J(2)	20	Gasifier Multicyclones
MC-1512A/F(2)	12	Combustor Cyclones

MAJOR EQUIPMENT LIST

Dwg. 1950-X-104B (Unit 16A)-Raw Gas Scrubbing & Ammonia Removal

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
DT-1610(2)	2	Venturi Scrubber & Knockout Tank
DT-1611(2)	2	Phosam Absorber
DT-1612(2)	2	Venturi Scrubber Purge Stripper
ES-1611(2)	2	Slurry Cooler
ES-1612(2)	2	No. 3 Gas Cooler
ES-1613(2)	2	No. 4 Gas Cooler
FA-1610(2)	2	Oil-Water Separator
FF-1610(2)	2	K.O. Tank
GP-1610A/B(2)	4	Venturi Scrubber Circulation Pumps
GP-1611A/B(2)	4	Venturi Scrubber Sidestream Pumps
GP-1612A/B(2)	4	Rich Phosam Solution Pumps
MC-1610(2)	2	Hydroclone

MAJOR EQUIPMENT LIST

Dwg. 1950-X-105 (Unit 15B) - Flue Gas Energy Recovery, Air Compression
& Reject Ash Facilities

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
BE-1551(2)	2	Flue Gas Heat Recovery System
CB-1561(2)	2	Ash Heat Recovery Exhaust Blower
CC-1551(2)	2	Air Compressor
CC-1552(2)	2	Flue Gas Recycle Compressor
CE-1551(2)	2	Flue Gas Expander
CT-1551(2)	2	Air Compressor Turbine Drive
ES-1551(2)	2	Air Compressor Intercooler
ES-1552(2)	2	HP BFW Preheater
FB-1561	1	Coal Ash Storage Bin
FF-1551(2)	2	Ash Tank
FF-1552(2)	2	Ash Collection Tank
MC-1551A/F(2)	12	Flue Gas Multi- Cyclones
MC-1561(2)	2	Ash Conveyor Cyclone
ME-1561(2)	2	Rotary Ash Cooler w/Cyclone
ME-1562(2)	2	Fine Ash Cooler
PC-1561(2)		Cold Ash Conveyor
PV-1561(2)	2	Rotary Feeder
PV-1562(2)	2	Rotary Feeder
PV-1563(2)	2	Rotary Feeder
PV-1564/1567	4	Ash Loading Feeders

MAJOR EQUIPMENT LIST

Dwg. 1950-X-106 (Unit 23)-Product Gas Compression, Drying & H₂S Removal

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
CB-2301(2)	2	Coal Dryer Air Blower
CC-2301(2)	2	First Stage Product Gas Compressor
CC-2302(2)	2	Second Stage Product Gas Compressor
CT-2301(2)	2	Product Gas Non-Condensing Turbine Drive
CT-2302(2)	2	Product Gas Condensing Turbine Drive
DT-2301(2)	2	Selexol Absorber
DT-2302(2)	2	Selexol Regenerator
ES-2301(2)	2	Product Gas First Intercooler
ES-2302(2)	2	Selexol Feed Bottoms Exchanger
ES-2303(2)	2	Selexol Cooler
ES-2304(2)	2	Selexol Reboiler
ES-2305(2)	2	Dryer Air Heater
ES-2306(2)	2	Product Gas Aftercooler
ES-2307(2)	2	Product Gas Second Intercooler
ES-2308 (2)	2	BFW Preheater
FF-2301(2)	2	Intercooler K.O. Drum
FF-2302(2)	2	Aftercooler K.O. Drum
FF-2303(2)	2	Compressor Condensate Flash Drum
FF-2304(2)	2	Selexol Flash Drum
FF-2305(2)	2	Absorber K.O. Drum
FF-2306(2)	2	Regenerator K.O. Drum
FF-2307(2)	2	Selexol Reboiler Condensate Tank
GP-2301A/C(2)	6	Selexol Pumps (4 pumps motor driven; 2 pumps driven by expansion turbine)
GP-2302A/B(2)	4	Reboiler Condensate Pumps
GT-2301(2)	2	Selexol Expansion Turbine

MAJOR EQUIPMENT LIST

Dwg. 1950-X-107 (Unit 16B)-Ammonia Recovery

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
DT-1651	1	Phosam Stripper
DT-1653	1	Ammonia Fractionating Column
ES-1650	1	Phosam Solution Exchanger
ES-1651	1	Phosam Solution Cooler
ES-1652	1	Phosam Stripper Primary Condenser
ES-1653	1	Phosam Stripper Secondary Condenser
ES-1654	1	Phosam Stripper Reboiler
ES-1655	1	Ammonia Condenser
FF-1650	1	Phosphoric Acid Tank
FF-1651	1	Caustic Solution Tank
FF-1652	1	Aqua Ammonia Tank
FF-1653	1	Product Ammonia Tank
FF-1654	1	Product Ammonia Tank
FF-1655	1	Contactator
FF-1656	1	Ammonia Reflux Tank
GP-1650	1	Phosphoric Acid Pump
GP-1651	1	Caustic Solution Rotary Pump
GP-1652A/B	2	Aqua Ammonia Pumps
GP-1653A/B	2	Product Ammonia Pumps
GP-1655A/B	2	Ammonia Reflux Pumps

MAJOR EQUIPMENT LIST

Dwg. 1950-X-108A (Unit 21A)-Flue Gas Cleaning & Sulfur Recovery (Citrate Plant)

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
ES-2101	1	Collection Tank Overhead Condenser
FF-2101	1	Condensate Collecting Tank
GP-2101A/B	2	L.P. Condensate Transfer Pumps
SR-1(2)	2	Citrate Sulfur Recovery Plant Package ,

MAJOR EQUIPMENT LIST

Dwg. 1950-X-108B (Unit 21B)-Sulfur Recovery (Claus & Beavon Plant)

<u>ITEM NO</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
CC-2161	1	Inert Gas Compressor
CB-2161(2)	2	Inert Gas Blower
ES-2161	1	Inert Gas Aftercooler
FF-2161	1	Inert Gas Surge Tank
SR-2(2)	2	Claus Sulfur Recovery Plant Package
SR-3	1	Beavon Sulfur Recovery Plant Package

SCIENTIFIC DESIGN COMPANY, INC.

MAJOR EQUIPMENT LIST

Dwg. 1950-X-109 (Unit 22)-Solids Disposal

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
MF-2211	1	Filter Package for Solids Recovery
PC-2211	1	Solids Discharge Conveyor

MAJOR EQUIPMENT LIST

Dwg. 1950-X-110 (Unit 20)-Liquid Effluent Treating

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
DT-2010	1	Sour Water Stripper
ES-2010	1	Stripper Overhead Condenser
ES-2011	1	Stripper Feed-Bottoms Exchanger
ES-2012	1	Stripper Bottoms Cooler
FF-2010	1	Stripper Reflux Drum
FF-2011	1	Waste Water Surge Tank
FF-2012	1	Treated Water Surge Tank
GP-2010A/B	2	Surge Tank Transfer Pumps
GP-2011A/B	2	Treated Water Pumps
GP-2020A/B	2	Sour Water Stripper Reflux Pumps
PR-2010	1	Phenol Removal Unit

MAJOR EQUIPMENT LIST

Dwg. 1950-X-111 (Unit 30)- Steam, Condensate & Power Generation

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
ED-3001(2)	2	Deaerator
ES-3001(2)	2	Turbine Exhaust Condenser
ES-3002(2)	2	Turbine Exhaust Condenser
ES-3003	1	Turbine Exhaust Condenser
ES-3004(2)	2	Blowdown Cooler
FF-3001(2)	2	Boiler Blowdown Tank
GP-3001A/B(2)	4	Turbine Condensate Pumps
GP-3002A/B(2)	4	Turbine Condensate Pumps
GP-3003A/B	2	Turbine Condensate Pumps
GP-3004A/B(2)	4	HP BFW Pumps
GP-3005A/B(2)	4	LP BFW Pumps
NG-3001	1	Turbine-Electric Generator

MAJOR EQUIPMENT LIST

Dwg. 1950-X-112 (Unit 31)- Plant Water System

<u>ITEM NO.</u>	<u>NO. REQ'D</u>	<u>ITEM NAME</u>
FA-3101	1	Water Reservoir
FA-3102	1	Waste Neutralizing Sump
FA-3103	1	Storm Water and Runoff Basin
FI-3101	1	Demineralized Water Tank
GP-3101A/C	3	Filter Feed Pumps
GP-3102A/C	3	Fire Water Pumps
GP-3103	1	Fire Water Jockey Pump
GP-3104A/D	4	L.P. Cooling Water Pumps
GP-3105A/B	2	H.P. Cooling Water Pumps
GP-3106A/B	2	Neutralizing Sump Pumps
GP-3107A/C	3	Demineralized Water Pumps
GP-3108	1	Storm Water Pump
MF-3101A/F	6	Anthracite Pressure Filters
UC-3101	1	Cooling Tower
UD-3101	1	Cooling Water Chemical Feeding System
UD-3102	1	Potable Water Treatment System
UD-3103	1	Demineralized Water Package

BATTELLE AGGLOMERATING ASH BURNER PROCESS

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. No. 1950-X-103 (Units 12 & 14)-Coal Preparation & Coal Feeding

FB-1201

Mill Feed Storage Bin

No. Req'd	:	One
Type	:	3 Bottom Spikes, 70 degree sides
Size	:	30'x160'x70' height
Mat'l of Const.	:	CS

FB-1202

Screen Feed Distribution Bin

No. Req'd	:	One
Type	:	10 Bottom Spikes, 70 degree sides
Size	:	25'x75'x16' height
Mat'l of Const.	:	CS

FB-1203

Dryer Feed Distribution Bin

No. Req'd	:	One
Type	:	3 Bottom Spikes, 70 degree sides
Size	:	25'x75'x16' height
Mat'l of Const.	:	CS

FB-1401

Prepared Coal Bin

No. Req'd	:	One
Type	:	4 Compartments
Size	:	30' x160'x7' vertical height
Mat'l of Const.	:	CS

FF-1401A/B (2)

Coal Feed Injection System

No. Req'd	:	Four units
Capacity	:	150 Ton/Hr of coal
Unit Description	:	Each unit consists of a lockhopper, a primary injector, piping, valves and instruments.
Mat'l of Const.	:	CS

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. No. 1950-X-103 (Units 12 & 14)-Coal Preparation & Coal Feeding (cont'd)

ME-1201

Coal Dryer Package

No. Req'd	:	Four units
Type	:	Flash drying system
Capacity	:	To dry 560 ton/day of wet coal from 10% to 5% moisture.
Unit Description	:	Each unit consists of a coal feeder, a gas-fired air heater, air piping, cyclone collector, vent fan and baghouse.
Mat'l of Const.	:	CS

MS-1201A/J

Screens

No. Req'd	:	Ten
Type	:	Rotex-vibrating
Capacity & Size	:	70 TPH capacity each; 72" diameter
Mat'l of Const.	:	CS; SS elements

SS-1201A/C

Coal Cage Mills

No. Req'd	:	Three
Type	:	2 Row Stedman cage mills
Drive	:	300 HP motors for each mill
Capacity & Size	:	330 TPH capacity each; 5' diameter
Mat'l of Const.	:	CS; SS elements

SPECIFICATIONS OF MAJOR EQUIPMENT

Drawing 1950-X-104 (Unit 15A)-Gasifier, Combustor & Gasifier Heat Recovery

BE-1510 (2)

Gasifier Heat Recovery System

No. Req'd : Two trains, each consisting of two 150 PSIG steam superheaters and two 1250 PSIG steam generator-superheaters.

Duty : 150 PSIG steam superheater: 150×10^6 Btu/Hr. Area: 10,100 Sq.Ft. each train
1250 PSIG steam generator and superheater: 246×10^6 Btu/Hr. Area: 18,300 Sq.Ft., each train.

Mat'l of Const. : CS shell, refractory lined; alloy steel tube coils

BF-1510 (2)

Combustor

No. Req'd : Two

Design Conditions : 125 PSIG, 600°F for vessel shell

Mat'l of Const. : $\frac{1}{2}$ Mo steel shell clad with 18-8 SS; refractory lined.

	Size		
	Inside Dia. Ft. Lining	Inside Dia. Ft. Vessel	Height Ft.
Top section	26'-0"	28'-6"	24'-0"*
Transition			36'-0"
Bottom section	12'-0"	14'-6"	20'-0"
Below bottom T.L.			10'-0"
			90'-0"

*Top spherical head excluded.

DR-1510 (2)

Gasifier

No. Req'd : Two

Design Conditions : 125 PSIG, 600°F for vessel shell

Mat'l of Const. : 18-8SS; refractory lined.

	Size		
	Inside Dia. Ft. Lining	Inside Dia. Ft. Vessel	Height Ft.
Top section	41'-0"	43'-6"	5'-0"*
Transition			28'-0"
Bottom section	30'-0"	32'-6"	25'-0"***
			58'-0"

* Top spherical head excluded.

**Bottom ellipitcal head excluded.

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-104A (Unit 15A)-(cont'd)

ES-1510 (2)

No. 1 Gas Cooler

No. Req'd	:	Two
Type	:	Shell and tube
Duty	:	146×10^6 Btu/Hr, each
Area	:	26,000 Sq. Ft., each
Mat'l of Const.	:	CS shell, 304 SS tubes

ES-1511 (2)

No. 2 Gas Cooler

No. Req'd	:	Two
Type	:	Shell and tube
Duty	:	50×10^6 Btu/Hr, each
Area	:	18,000 Sq. Ft., each
Mat'l of Const.	:	CS shell, 304 SS tubes

MC-1510A/F (2)

Gasifier Cyclones

No. Req'd	:	Twelve
Size	:	6'-0" diam.
Mat'l of Const.	:	CS, Refractory-lined

MC-1511 A/J (2)

Gasifier Multi-cyclones

No. Req'd	:	Twenty
Size	:	3'-6" diam.
Mat'l of Const.	:	CS

MC-1512A/F (2)

Combustor Cyclones

No. Req'd	:	Twelve
Size	:	7'-0" diam.
Mat'l of Const.	:	CS, Refractory-lined

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-104B (Unit 16A)-Raw Gas Scrubbing & Ammonia Removal

DT-1610 (2)

Venturi Scrubber & Knockout Tank

No. Req'd	:	Two
Size	:	Venturi 4'2" ID x 14' long
		Knockout Tank 12'6" ID x 28' T-T
Capacity	:	50,000 ACFM
Mat'l of Const.	:	304 SS clad steel

ES-1612 (2)

No. 3 Gas Cooler

No. Req'd	:	Two
Type	:	Shell and tube
Duty	:	42×10^6 Btu/Hr, each
Area	:	5,300 Sq. Ft., each
Mat'l of Const.	:	CS shell, 304 SS tubes

ES-1613 (2)

No. 4 Gas Cooler

No. Req'd	:	Two
Type	:	Shell and tube
Duty	:	56×10^6 Btu/Hr, each
Area	:	9,000 Sq. Ft., each
Mat'l of Const.	:	CS shell, 304 SS tubes

FF-1610 (2)

K.O. Tank

No. Req'd	:	Two
Type	:	Vertical
Size	:	12-6" OD x 12'-0" T-T
Mat'l of Const.	:	304 SS clad CS

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-105 (Unit 15B)-Flue Gas Energy Recovery, Air Compression & Reject Ash Facilities

BE-1551 (2)

Flue Gas Heat Recovery System

No. Req'd	:	Two trains, each consisting of two 1250 PSIG steam generator-superheaters & two Combustor Air Preheaters
Duty	:	718x10 ⁶ Btu/Hr, total for one train
Mat'l of Const.	:	CS shell, refractory lined; alloy steel tube coils.
Area	:	49,400 Sq.Ft., total for one train

CC-1551 (2)

Air Compressor

No. Req'd	:	Two
Type	:	Axial and centrifugal
Capacity	:	377,500 ACFM each, at suction
BHP	:	97,000
Pressure	:	Suction-14.4 PSIA; discharge-140 PSIA
Drive	:	Flue Gas Expander and condensing turbine
Mat'l of Const.	:	Manufacturer's standard

CC-1552 (2)

Flue Gas Recycle Compressor

No. Req'd	:	Two
Capacity	:	500 ACFM each, at suction
Pressure	:	Suction - 105 PSIA; discharge - 140 PSIA
BHP	:	100
Drive	:	Electric motor 125 HP
Mat'l of Const.	:	Manufacturer's standard

CE-1551 (2)

Flue Gas Expander

No. Req'd	:	Two
Flow Conditions	:	Inlet: flue gas at 498°F and 105 PSIA Outlet: 250°F and 17 PSIA
Flow Rate	:	92,000 ACFM each, at suction
BHP	:	42,100
Mat'l of Const.	:	Manufacturer's standard

CT-1551 (2)

Air Compressor Turbine Drive

No. Req'd	:	Two
Steam Inlet Conditions	:	1200 PSIG and 900°F
Steam Exhaust Pressure	:	2-1/2" Hg, absolute
Steam Flow	:	362,750 Lb/Hr each
BHP	:	54,900 each
Mat'l of Const.	:	Manufacturer's standard

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-105 (Unit 15B) (cont'd)

ES-1551 (2)

Air Compressor Intercooler

No. Req'd	:	Two
Type	:	Shell and tube
Duty	:	54×10^6 Btu/Hr, each
Area	:	37,000 Sq. Ft., each
Mat'l of Const.	:	CS

ES-1552 (2)

HP BFW Preheater

No. Req'd	:	Two
Type	:	Shell and tube
Duty	:	51×10^6 Btu/Hr, each
Area	:	27,000 Sq. Ft., each
Mat'l of Const.	:	CS

FB-1561

Cold Ash Storage Bins

No. Req'd	:	One
Type	:	Four straight compartments with tapered bottoms
Size	:	20'x80'x46' height
Mat'l of Const.	:	CS

MC-1551 A/F (2)

Flue Gas Multi-cyclones

No. Req'd	:	Twelve
Size	:	6'-0" diameter
Mat'l of Const.	:	CS

MC-1561 (2)

Ash Conveyor Cyclones

No. Req'd	:	Two
Size	:	10'-0" diameter
Mat'l of Const.	:	CS

ME-1561 (2)

Rotary Ash Cooler W/Cyclone

No. Req'd	:	Two
Type	:	Rotary, Direct air cooled
Size	:	10' diam. x 55' long
Mat'l of Const.	:	CS

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-106 (Unit 23)-Product Gas Compression, Drying & H₂S Removal

CB-2301 (2)

Coal Dryer Air Blower

No. Req'd	:	Two
Type	:	Centrifugal
Capacity	:	76,000 ACFM @ 28" water
BHP	:	475
Drive	:	Electric Motor 500 HP each
Mat'l of Const.	:	CS

CC-2301 (2)

First Stage Product Gas Compressor

No. Req'd	:	Two
Type	:	Centrifugal
Capacity	:	45,100 ACFM, each
Pressure	:	Suction-100 PSIA; discharge-204 PSIA
BHP	:	20,900 each
Drive	:	Non-condensing Turbine
Mat'l of Const.	:	Manufacturer's standard

CC-2302 (2)

Second Stage Product Gas Compressor

No. Req'd	:	Two
Type	:	Centrifugal
Capacity	:	22,300 ACFM each
Pressure	:	Suction-201 PSIA; discharge-535 PSIA
BHP	:	30,000 each
Drive	:	Condensing Turbine
Mat'l of Const.	:	Manufacturer's Standard

CT-2301 (2)

Product Gas Non-Condensing Turbine Drive

No. Req'd	:	Two
BHP	:	20,900 each
Steam Conditions	:	Inlet: 1200 PSIG & 900 ^o F Outlet: 150 PSIG & 400 ^o F
Mat'l of Const.	:	Manufacturer's standard

CT-2302 (2)

Product Gas Condensing Turbine Drive

No. Req'd	:	Two
BHP	:	30,000 each
Steam Conditions	:	Inlet: 1200 PSIG & 900 ^o F Outlet: 2-1/2" Hg absolute
Mat'l of Const.	:	Manufacturer's standard

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-106 (Unit 23) (cont'd)

Two (2) Selexol systems, one for each train, for removing H_2S from fuel gas and regenerating at about 35% H_2S ; consisting of Selexol Absorbers DT-2301 (2), Regenerators DT-2302, Pumps, Expansion Turbines, Heat Exchangers and Accessory Equipment.

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-107 (Unit 16B) - Ammonia Recovery

One Phosam system for recovery of 12 ton/day of anhydrous ammonia from raw fuel gas; consisting of Phosam Absorbers DT-1611 (2), Phosam Stripper DT-1651, Ammonia Fractionating Column DT-1653, Pumps, Heat Exchangers and Accessory Equipment.

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-108A (Unit 21A) - Flue Gas Cleaning & Sulfur Recovery (Citrate Plant)

SR-1(2)

Citrate Plant

No. Req'd	:	Two package units.
Capacity, total for two units	:	245 LT/D of sulfur, and 30 ST/D of anhydrous sodium sulfate.

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-108B (Unit 21B)-Sulfur Recovery (Claus and Beavon Plant)

SR-2(2)

Claus Plant Package

No. Req'd	:	Two package units
Capacity, total for two units	:	175 LT/D of sulfur

SR-3

Beavon Plant Package

No. Req'd	:	One package unit
Capacity	:	19 LT/D of sulfur and 2 ST/D of anhydrous sodium sulfate.

CC-2161

Inert Gas Compressor

No. Req'd	:	One
Type	:	Centrifugal
Capacity	:	4,000 CFM at suction
BHP	:	1200
Pressure	:	Suction 0 PSIG, discharge 300 PSIG
Drive	:	1250 HP motor
Mat'l of Const.	:	Manufacturer's standard

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-109 (Unit 22) - Solids Disposal

MF-2211

Filter Package for Solids Recovery

No. Req'd

:

One Packaged Unit, for recovery
of solids from slurry streams:

	<u>Lb/Hr</u>
Fine coal particles	1000
Fine char particles	200
Fine ash particles	<u>200</u>
	1400

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-110 (Unit 20) - Liquid Effluent Treatment

DT-2010

Sour Water Stripper

No. Req'd	:	One
Type	:	20 Trays
Size	:	4' dia. x 40' T-T
Mat'l of Const.	:	CS

FF-2011

Waste Water Surge Tank

No. Req'd	:	One
Type	:	Vertical
Size	:	14' dia. x 10' T-T
Mat'l of Const.	:	CS

FF-2012

Treated Water Surge Tank

No. Req'd	:	One
Type	:	Vertical
Size	:	14' dia. x 10' T-T
Mat'l of Const.	:	CS

PR-2010

Phenol Removal Unit

No. Req'd	:	One Packaged unit, to remove approximately 500 Lb/Hr of phenols from waste water as a 72 wt. % aqueous solution for use as fuel in plant.
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SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-111 (Unit 30)-Steam, Condensate & Power Generation

ED-3001 (2)

Deaerator

No. Req'd	:	Two
Type	:	Manufacturer's standard
Size	:	50,000 gallon storage capacity, each 2200 GPM water flow, each
Mat'l of Const.	:	CS

ES-3001 (2)

Turbine Exhaust Condenser

No. Req'd	:	Two
Type	:	Surface condenser
Duty	:	363×10^6 Btu/Hr, each
Area	:	44,000 Sq. Ft., each
Mat'l of Const.	:	CS

ES-3002 (2)

Turbine Exhaust Condenser

No. Req'd	:	Two
Type	:	Surface condenser
Duty	:	97.5×10^6 Btu/Hr, each
Area	:	11,800 Sq. Ft., each
Mat'l of Const.	:	CS

ES-3003

Turbine Exhaust Condenser

No. Req'd	:	One
Type	:	Surface condenser
Duty	:	202×10^6 Btu/Hr
Area	:	24,500 Sq. Ft.
Mat'l of Const.	:	CS

FF-3001 (2)

Boiler Blowdown Tank

No. Req'd	:	Two
Type	:	Vertical
Size	:	5'I.D. x6' T-T
Mat'l of Const.	:	CS

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-111 (Unit 30) (cont'd)

GP-3001 (2)

Turbine Cond. Pumps

No. Req'd	:	Four
Type	:	Centrifugal
Capacity	:	680 GPM @ 160 Ft. TDH
Drive	:	40 HP motor
Mat'l of Const.	:	CS

GP-3002A/B (2)

Turbine Cond. Pumps

No. Req'd.	:	Four
Type	:	Centrifugal
Capacity	:	365 GPM @ 160 ft. TDH
Drive	:	25 HP motor
Mat'l of Const.	:	CS

GP-3003A/B

Turbine Cond. Pumps

No. Req'd	:	Two
Type	:	Centrifugal
Capacity	:	172 GPM @ 160 ft. TDH
Drive	:	15 HP motor
Mat'l of Const.	:	CS

GP-3004A/B (2)

HP BFW Pumps

No. Req'd	:	Four
Type	:	Horizontal centrifugal
Capacity	:	2300 GPM @ 4470 ft. TDH
Drive	:	3000 HP motor
Mat'l of Const.	:	CS

GP-3005A/B(2)

LP BFW Pumps

No. Req'd	:	Four
Type	:	Horizontal centrifugal
Capacity	:	75 GPM @ 60 ft. TDH
Drive	:	3 HP motor
Mat'l of Const.	:	CS

NG-3001

Turbine Electric Generator

No. Req'd	:	One
Size	:	25 MW,
Drive	:	Steam Turbine
		172,100 Lb/hr steam
		1265 PSIA, 905°F
		condensing pressure 2.5" Hg absolute

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-112 (Unit 31)-Plant Water System

FA-3101

Water Reservoir

No. Req'd	:	One
Size	:	360'x360'x15' Deep
Capacity	:	14,000,000 gallon, 2 Day supply
Mat'l of Const.	:	Concrete

FA-3102

Waste Neutralizing Sump

No. Req'd	:	One
Size	:	25'x25'x15' Deep
Capacity	:	60,000 gallon
Mat'l of Const.	:	Concrete

FI-3101

Demineralized Water Tank

No. Req'd	:	One
Type	:	API
Size	:	775,000 gallons, 60' D x 38' height 8 hour inventory
Mat'l of Const.	:	CS

GP-3101A/C

Filter Feed Pumps

No. Req'd	:	Three
Type	:	Vertical
Drive	:	75 HP motor
Capacity	:	800 GPM @ 230 ft. TDH
Mat'l of Const.	:	Cast iron

GP-3102A/C

Fire Water Pumps

No. Req'd	:	Three
Type	:	Vertical
Capacity	:	5000 GPM @ 340 ft. TDH
Drive	:	2 motors at 700 HP, 1 Diesel
Mat'l of Const.	:	Cast iron

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-112 (Unit 31) (cont'd)

GP-3104A/C

LP Cooling Water Pumps

No. Req'd	:	Four
Type	:	Vertical turbine
Capacity	:	43,300 GPM @ 70 ft. TDH
Drive	:	900 HP motor
Mat'l of Const.	:	CS

GP-3105A/B

HP Cooling Water Pumps

No. Req'd	:	Two
Type	:	Vertical turbine
Capacity	:	38,000 GPM @ 175 ft. TDH
Drive	:	2,000 HP Motor
Mat'l of Const.	:	CS

GP-3107A/C

Demineralized Water Pumps

No. Req'd	:	Three
Type	:	Centrifugal
Capacity	:	800 GPM @ 230 ft. TDH
Drive	:	75 HP motor
Mat'l of Const.	:	CS

MF-3101 A/F

Anthracite Pressure Filters

No. Req'd	:	Six
Size	:	12 ft. diam. flow capacity 1600 GPM
Mat'l of Const.	:	Manufacturer's standard

UC-3101

Cooling Tower

No. Req'd	:	One
Type	:	12 cells, induced draft; crossflow
Duty	:	170,000 GPM, $1,375 \times 10^6$ Btu/Hr
		12 fans @ 200 HP each - wet bulb temperature 77°F

SPECIFICATIONS OF MAJOR EQUIPMENT

Dwg. 1950-X-112 (Unit 31) (cont'd)

UD-3101

Cooling Water Chemical Feeding System

No. Req'd	:	One
Capacity	:	Polyphosphate Feed System
		Capacity 15 Lb/Hr

UD-3103

Demineralized Water Package

No. Req'd	:	One
Capacity	:	1600 GPM total capacity;
Size	:	Four trains of 10 ft. diam. x
		6 ft. T-T
		Cation and anion units
Mat'l of Const.	:	Rubber lined vessels

The diagram illustrates the process flow of a coal gasification plant, starting from raw water and coal delivery, through various processing units, and ending with product gas and waste streams.

Units and Processes:

- UNIT 31 PLANT WATER SYSTEM:** Receives raw water and produces treated water.
- UNIT 20 STEAM, COND. & POWER:** Receives treated water and produces HP steam.
- UNIT 11 COAL STORAGE & RECLAIMING:** Receives coal from delivery.
- UNIT 12 COAL PREPARATION:** Receives coal from Unit 11.
- UNIT 14 COAL FEEDING:** Receives coal from Unit 12.
- UNIT 15A GASIFICATION COMBUSTION:** Receives coal from Unit 14 and produces raw gas, hot ash, and flue gas.
- UNIT 15B REJECT ASH FACILITIES:** Receives hot ash from Unit 15A.
- UNIT 15B ENERGY RECOVERY & AIR COMPRESSION:** Receives flue gas from Unit 15A and produces superheated L.P. steam and compressed air.
- UNIT 15A GAS HEAT RECOVERY:** Receives raw gas from Unit 15A and produces steam.
- UNIT 16A RAW GAS QUENCH:** Receives raw gas from Unit 15A and produces sour water.
- UNIT 16A AMMONIA SCRUBBING:** Receives raw gas from Unit 16A and produces rich and lean phosphoric acid solutions.
- UNIT 16B AMMONIA RECOVERY:** Receives rich phosphoric acid solution from Unit 16A and produces lean phosphoric acid solution and acid gas.
- UNIT 23 PRODUCT GAS COMPRESSION:** Receives raw gas from Unit 16A and produces L.P. steam and H.P. steam.
- UNIT 23 H₂S REMOVAL & GAS DRYING:** Receives product gas from Unit 23 and produces medium BTU fuel gas.
- UNIT 23 SOLIDS DISPOSAL:** Receives coal fines slurry from Unit 12 and produces waste water.
- UNIT 21A Sulfur Recovery:** Receives ash slurry and char fines slurry from Unit 15A and produces sulfur.
- UNIT 20 SOUR WATER STRIPPER:** Receives sour water from Unit 16A and produces acid gas.
- UNIT 20 PHENOL REMOVAL:** Receives acid gas from Unit 20 and produces waste water.

Streams and Outputs:

- RAW WATER:** Input to Unit 31.
- COAL FROM DELIVERY:** Input to Unit 11.
- AIR:** Input to Unit 15B.
- TREATED WATER:** Output from Unit 31 to Unit 20.
- HP STEAM:** Output from Unit 20 to various units.
- LP STEAM:** Output from Unit 20 to various units.
- COAL:** Output from Unit 11 to Unit 12.
- COAL FINES SLURRY:** Output from Unit 12 to Unit 23.
- RAW GAS:** Output from Unit 15A to Unit 15A Gas Heat Recovery, Unit 16A, and Unit 23.
- HOT ASH:** Output from Unit 15A to Unit 15B Reject Ash Facilities.
- FLUE GAS:** Output from Unit 15A to Unit 15B Energy Recovery & Air Compression.
- COMpressed AIR:** Output from Unit 15B to Unit 15A.
- SOUR WATER:** Output from Unit 16A to Unit 20 and Unit 16B.
- RICH PHOSPHORIC ACID SOLUTION:** Output from Unit 16A to Unit 16B.
- LEAN PHOSPHORIC ACID SOLUTION:** Output from Unit 16B to Unit 16A.
- ACID GAS:** Output from Unit 16B to Unit 20 and Unit 20 Phenol Removal.
- WASTE WATER:** Output from Unit 23 Solids Disposal and Unit 20 Phenol Removal.
- CLEAN FLUE GAS TO ATMOSPHERE:** Output from Unit 15B.
- BY-PRODUCT SULFUR:** Output from Unit 21A.
- BY-PRODUCT ANHYDROUS SODIUM SULFATE:** Output from Unit 20.
- DRY ASH TO DISPOSAL:** Output from Unit 15B.

SCIENTIFIC DESIGN COMPANY INC.
 DESIGN • TWO PARK AVENUE • CONSTRUCTION
 NEW YORK

BATTELLE AGGLOMERATING ASH PROCESS
BLOCK FLOW DIAGRAM
PLANT CAPACITY 250 X 10³ BTU/DAY
FUEL GAS

DRAWN BY TC/JM DATE 12/178 SCALE NONE
CHECKED BY DATE APPROVED BY *AKW* DATE 5/15/8
APPROVED FOR CONSTRUCTION BY DATE No. 1950-X-101

NO.	DESCRIPTION	DATE	BY	CHECKED
REVISIONS				

NOTES

<u>PC-1113</u>	<u>PC-1118</u>	<u>PC-1114</u>	<u>GP-1110A/B</u>	<u>PC-1115</u>	<u>MS-1111</u>
<u>BUCKET WHEEL</u>	<u>COLLECTING</u>	<u>RECLAIMER</u>	<u>STORAGE AREA</u>	<u>MILL GUARD</u>	<u>MILL GUARD</u>
<u>RECLAIMER</u>	<u>& TRANSFER</u>	<u>STACKER</u>	<u>SUMP PUMPS</u>	<u>ELECTRIC HOIST</u>	<u>ELECTRO MAGNETIC</u>
	<u>CONVEYOR</u>	<u>TRANSFER CAR</u>		<u>& MONORAIL</u>	<u>SEPARATOR</u>

PC-1110	PX-1110	MS-1110	PC-1112	PC-1113	PC-1114	SX-1110	PC-1115	PS-1110	PC-1116	PC-1117A,B	PX-1117A,B	PX-1112
COAL RECEIVING
CONVEYOR	SAMPLER GUARD
ELECTRIC HOIST
AND MONORAIL	SAMPLER GUARD
ELECTROMAGNETIC
SEPARATOR	CALIBRATION CHAIN
AND REEL	BELT SCALE	CROSS CONVEYOR
No 1	AUTOMATIC SAMPLER
PACKAGE	CROSS CONVEYOR
No 2	DIVERTING CHUTE	CROSS CONVEYOR
No. 3	BEDDING & RECLAIM
CONVEYORS	BELT TRAILERS	TRAVELING
BOOM STACKER

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NEW YORK

BATTELLE AGGLOMERATING ASH PROCESS

COAL STORAGE & RECLAIMING UNIT II

PLANT CAPACITY: 250 X 10⁹ BTU/DAY

FUEL GAS

DRAWN BY ER/J.M. DATE 7/25/78 SCALE _____

CHECKED BY _____ DATE _____ APPROVED BY ARW DATE 5/15/72

APPROVED FOR CONSTRUCTION BY _____ DATE _____ No. 1950-X-102

No. 1950-X-102

NOTES:
1. ONE (1) TRAIN IS REQUIRED.

SCIENTIFIC DESIGN COMPANY INC.	
DESIGN • TWO PARK AVENUE • CONSTRUCTION NEW YORK	
BATTELLE AGGLOMERATING ASH PROCESS	
COAL PREPARATION (COAL FEEDING) UNIT 15	
PLANT CAPACITY 250 X 10 ³ TON/DAY	
FUEL GAS	
DRAWN BY ER/RJC DATE 5-29-78 SCALE 1/4" = 1'-0"	
CHECKED BY DATE APPROVED BY DATE 5/15/78	
APPROVED FOR CONSTRUCTION BY DATE	
No. 1950-X-103	

A		B		C		D		E		F		G		H		I		J		K		L		M		N		O			
STREAM NO.		4001		4002		4003		4004		4005		4006		4007		4008		4009		4010		4011		4012		4013					
SERVICE		COAL		TOTAL STEAM FEED		STEAM TO COAL FEED		ASH		GASIFIER EFFLUENT GAS		CHAR		ASH/AIR		COMPRESSED AIR		REJECT ASH		ASH FROM GASIFIER		FLUE GAS		RAW GAS TO SCRUBBER		RECYCLE FLUE GAS					
COMPONENT	M.W.	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%	LB/HR	MOLE/HR	WT%
HYDROGEN	2										49.516	24.758	51.23									49.516	24.758	51.23							
CARBON MONOXIDE	28										49.516	1.755	3.73									49.516	1.755	3.73							
CARBON DIOXIDE	44										13.932	1.081	3.47									13.932	1.081	3.47	1000	39.1	19.52				
METHANE	16										24.154	1.509	3.12									24.154	1.509	3.12							
HYDROGEN SULFIDE	34										18.200	535.3	1.11									18.200	535.3	1.11							
AMMONIA	17										5.5	30.5	.06									5.5	30.5	.06							
NITROGEN	28										13.3590	605.8	78.06				13.3590	605.8	78.06			13.3590	605.8	78.06							
WATER (VAPOR)	18										12.141	678.5	1.19				12.141	678.5	1.19			12.141	678.5	1.19	90.11	136.1	18.06				
OXYGEN	32										316.749	11.710	9				316.749	11.710	9			316.749	11.710	9							
SULFUR DIOXIDE	64																														
COAL (DRY)		509245		95.0																											
WATER (LIQUID)		26012		5.0																											
CARBON																															
SULFUR																															
ASH																															
TOTAL		532245		100.00	310.470	17803.9	100.00	49.510	2.750.6	12.99.01	482.176	48215.5	100.00	139.167	100.00	16.10.480	56.44.1.1	100.00	62.6.41	12.99.01		17803.9	56.144.7	100.00	48215.5	100.00	94.17	302.5	100.00

NOTES:
1. TWO (2) TRAINS ARE REQUIRED
2. MATERIAL BALANCES ARE FOR ONE OF TWO - 50% CAPACITY TRAINS

DRAWING NO. 1950-X-104A
KEY
◇ STREAM NO.
□ TEMPERATURE °F
○ PRESSURE PSIA (UNLESS OTHERWISE NOTED)
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DESIGN • TWO PARK AVENUE • CONSTRUCTION
NEW YORK
BATTELLE AGGLOMERATING ASH PROCESS
GASIFIER, COMBUSTOR & GASIFIER HEAT RECOVERY-UNIT
FOR
PLANT CAPACITY 250 x 10⁹ BTU/HR
FUEL GAS
DRAWN BY HJ/J.M. DATE 7/25/78 SCALE 1/4"=1'-0"
APPROVED BY [Signature] DATE 8/5/78
NO. 1950-X-104A

STREAM NO.	5001/4011	5002	5003/4013
SERVICE	FLUE GAS	FLUE GAS TO EXHAUST	RECYCLE FLUE GAS
COMPONENT	LB/HR	LB/HR	LB/HR
HYDROGEN	7.8	19.91	7.600
CARBON MONOXIDE	4.4	11.018	4.4
CARBON DIOXIDE	16	40.754	16.0
METHANE	18	45.901	18.06
NITROGEN	18	45.901	18.06
WATER (VAPOUR)	52	133.0	52.0
OXYGEN	64	164.9	64.0
SULFUR DIOXIDE	64	164.9	64.0
TOTAL	246.25	624.7	246.25

NOTES:
1. TWO (2) TRAINS ARE REQUIRED.
2. MATERIAL BALANCES ARE FOR ONE OF TWO - 50% CAPACITY TRAINS.

DRAWING NO. KEY
◇ STREAM NO.
□ TEMPERATURE °F
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BATTILL AGGLOMERATING ASH PROCESS
FLUE GAS ENERGY RECOVERY
AIR COMPRESSION (PROJECT ASH FACILITIES UNIT 19B)

PLANT CAPACITY 250 x 10³ BTU/DAY

FUEL GAS

DRAWN BY: BN/WR DATE: 1-18-78 SCALE: NONE
CHECKED BY: DATE: APPROVED BY: DATE: 5/6/78
APPROVED FOR CONSTRUCTION BY: DATE: No. 1950-X-105

A	B	C	D	E
STREAM NO.	6001	6002	6003	6004
SERVICE	GAS FEED TO COMPRESSOR	GAS TO SELEXOL ABSORBER	SELEXOL ABSORBER OVERHEAD GAS	SELEXOL REGENERATOR STRIPPING GAS
COMPONENT	MMB	MMB	MMB	MMB
HYDROGEN	44.44	24.81	24.81	24.81
CARBON MONOXIDE	4.4	2.48	2.48	2.48
CARBON DIOXIDE	4.4	2.48	2.48	2.48
METHANE	4.4	2.48	2.48	2.48
HYDROGEN SULFIDE	4.4	2.48	2.48	2.48
AMMONIA	4.4	2.48	2.48	2.48
NITROGEN	4.4	2.48	2.48	2.48
WATER (VAPOR)	4.4	2.48	2.48	2.48
TOTAL	62.44	34.81	34.81	34.81

F	G	H	I	J
STREAM NO.	6007	6008	6009	6010
SERVICE	PRODUCT GAS TO PIPELINE	GAS TO BEAVON PLANT	GAS TO COAL DRYER	GAS TO COAL FEED INJECTION PACKAGE
COMPONENT	MMB	MMB	MMB	MMB
HYDROGEN	44.44	24.81	24.81	24.81
CARBON MONOXIDE	4.4	2.48	2.48	2.48
CARBON DIOXIDE	4.4	2.48	2.48	2.48
METHANE	4.4	2.48	2.48	2.48
HYDROGEN SULFIDE	4.4	2.48	2.48	2.48
AMMONIA	4.4	2.48	2.48	2.48
NITROGEN	4.4	2.48	2.48	2.48
WATER (VAPOR)	4.4	2.48	2.48	2.48
TOTAL	62.44	34.81	34.81	34.81

NOTES:
1. TWO (2) TRAINS ARE REQUIRED
2. MATERIAL BALANCES ARE FOR ONE OF TWO 50% CAPACITY TRAINS.

DRAWING NO. 1950-X-106
KEY
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□ TEMPERATURE °F
○ PRESSURE PSIA
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DESIGN • TWO PARK AVENUE • CONSTRUCTION
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BATTLE AGGLOMERATING ASH PROCESS
PRODUCT GAS COMPRESSION, DRYING (H₂S REMOVAL) UNIT 23
FOR
PLANT CAPACITY 250 X 10³ BTU/HR
FUEL GAS
DRAWN BY: B.V. NO. DATE: 7/25/78 SCALE: 1/2"=1'-0"
CHECKED BY: DATE: APPROVED BY: DATE: 8/1/78
APPROVED FOR CONSTRUCTION BY: DATE: No. 1950-X-106

STREAM NO.	B	C	D	E	F	G	H	I	K	L	M	N	O
COMPONENT	M.W.	LB/HR	MOL/HR	MOL%	LB/HR	MOL/HR	MOL%	LB/HR	MOL/HR	MOL%	LB/HR	MOL/HR	MOL%
HYDROGEN	2												
CARBON MONOXIDE	28												
CARBON DIOXIDE	44	284,854	11,014.4	14.52	484,854	11,014.4	18.67	28,214	647.6	15.90			
METHANE	16							159	10.0	0.57			
NITROGEN	28							11,734	521.5	35.10			
WATER (VAPOR)	18	17,341	674.5	1.19	59,886	3,321.0	5.64	1,482	107.0	1.96			
OXYGEN	32												
SULFUR DIOXIDE	64	8,589	134.2	0.24	837	13.0	0.02						
TOTAL		1,157,008	54,442.2	100.00	1,796,996	58,938.5	100.00	53,445	1,481.0	100.00			

- NOTES:
1. THE CITRATE PLANT, ITEM NO. 32-1, SHOWN IN THIS DRAWING WILL BE SUPPLIED BY THE VENDOR AS A PACKAGE.
 2. THE CITRATE PLANT HAS TWO IDENTICAL PARALLEL TRAINS.
 3. MATERIAL BALANCES ARE FOR ONE OF TWO 50% CAPACITY TRAINS.

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SCIENTIFIC DESIGN COMPANY INC.
DESIGN : TWO PARK AVENUE • CONSTRUCTION
NEW YORK

BATTELLE AGGLOMERATING ASH PROCESS

FLUE GAS CLEANING & SULFUR RECOVERY (UNIT 21A)

PLANT CAPACITY 250 X 10⁶ BTU/DAY - UNIT 21A

FUEL GAS

DRAWN BY: JC/ JM DATE: 7/23/78 SCALE: NONE

CHECKED BY: DATE: APPROVED BY: 22/24 DATE: 5/1/79

APPROVED FOR CONSTRUCTION BY: DATE: No. 1950-X-108A

NO.	DESCRIPTION	DATE	BY	CHKD BY

STEAM NO.	8051/2 X 6008	8052/2013	8053/2 X 6004
SERVICE	FUEL GAS TO BURNER	SOUR GAS TO STEELWORK ABSORBER	INERT GAS TO SELEXOL REBOILER
COMPONENT	LB/HR	MOL/HR	MOL %
HYDROGEN	2	212	56.69
CARBON MONOXIDE	28	171.6	35.74
CARBON DIOXIDE	4.6	10.3	3.33
METHANE	16	16.6	0.34
HYDROGEN SULFIDE	14	4	0.14
AMMONIA	14	10	0.6
NITROGEN	28	28.36	5.94
WATER	14	140	80.8
TOTAL	6,418	480.0	100.00

NOTE:
1. TWO (2) CLAUß PLANT PACKAGES REQUIRED.
2. ONE (1) BEAVON PLANT PACKAGE REQUIRED.

DRAWING NO. KEY

- ◇ STREAM NO.
- TEMPERATURE °F
- PRESSURE PSIA (UNLESS OTHERWISE NOTED)

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SCIENTIFIC DESIGN COMPANY INC.
DEBORAH J. TWO PARK AVENUE NEW YORK

BATTLE-ACQUILONEATING ASH PROCESS
SULFUR RECOVERY (CLAUß AND BEAVON PLANT) - UNIT 21B
PLANT CAPACITY - 250 X 10³ BTU/DAY
FUEL GAS

DRAWN BY TC/JP DATE 7/25/78 SCALE 1/8" = 1'-0"
CHECKED BY DATE APPROVED BY DATE
APPROVED FOR CONSTRUCTION BY DATE

No. 1950-X-108B

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SCIENTIFIC DESIGN COMPANY, INC.
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BATTELLE AGGLOMERATING ASH PROCESS

SOLIDS DISPOSAL - UNIT - 22

PLANT CAPACITY 250 X 10³ BTU/DAY

FUEL GAS

DRAWN BY ER/BJC DATE 7-25-78 SCALE

CHECKED BY DATE APPROVED BY DATE

APPROVED FOR CONSTRUCTION BY DATE No. 1950-X-109

VI. CAPITAL COST ESTIMATE

BASIS OF ESTIMATE

The capital cost estimate presented is order of magnitude type, comparable to CFBraun's work, based on engineering data taken from the process flow diagrams as presented in Section V and equipment specifications.

In preparing the equipment specifications, use was made of technical information from vendors, especially for the proprietary processes. In addition SD's judgement and experience were utilized. A complete major equipment list and specifications for the major items are included in Section V. The Total Plant Investment Summary, Table VI-1 provides a breakdown of costs for each process flowsheet. The estimate summaries in Tables VI-1A through L provide a breakdown of costs by type of equipment.

Major Equipment

The major equipment costs were obtained from supplier budget price estimates and in-house estimates. An allowance for miscellaneous equipment and facilities is included to cover known cost items that are required for a completely engineered process unit, but are individually too small to justify the time to develop them for an order of magnitude type estimate.

Installation Factors

The installed cost was determined by applying installation factors to the

major equipment costs. These factors are based on SD experience with similar equipment modified to reflect the special requirements of this technology.

General Facilities

This item covers costs for site preparation, electrical distribution, yard piping, buildings, flare system, tankage, shipping and receiving facilities, sulfur storage, potable and sanitary water, holding ponds, plant roads, and the like. These costs are estimated at approximately ten percent of the total installed cost exclusive of home office cost, fee and contingency.

Contractor's Home Office Costs and Fee

The home office costs cover project management, engineering, purchasing, estimating cost control, scheduling, accounting, and other home office and consulting costs. Approximately ten percent of the total plant investment, exclusive of contingency, is included to cover contractor's home office costs and fee.

Contingency

This is included as approximately fifteen percent of the total cost.

Contingency covers unknowns such as engineered equipment sizing and pricing, materials of construction, installation factors, general facilities and contractor's home office costs.

Escalation

The capital cost estimates for all processes are based on May 1978 costs with no forward escalation included.

ESTIMATED CAPITAL COST

The estimated total investment of a plant to product 250 billion Btu/Stream day of medium Btu fuel gas based on the flowsheets in Section V is \$370,000,000 (mid-1978 dollars).

BATTELLE AGGLOMERATING ASH BURNER PROCESS

TOTAL PLANT INVESTMENT SUMMARY

TABLE VI-1

<u>DWG. NO.</u> <u>1950-X-</u>	<u>UNIT NO.</u>	<u>DESCRIPTION</u>	<u>COST</u>
102	11	Coal Storage & Reclaiming	\$13,000,000
103	12 & 14	Coal Preparation & Coal Feeding	32,800,000
104A	15A	Gasifier, Combustor & Gasifier Heat Recovery	67,000,000
104B	16A	Raw Gas Scrubbing & Ammonia Removal	5,400,000
105	15B	Flue Gas Energy Recovery, Air Compression & Reject Ash Facilities	41,000,000
106	23	Product Gas Compression, Drying & H ₂ S Removal	34,300,00
107	16B	Ammonia Recovery	2,500,000
108A&B	21A&B	Flue Gas Cleaning & Sulfur Recovery (Citrato Plant); Sulfur Recovery (Claus & Beavon Plant)	34,500,000
109	22	Solids Disposal	2,600,000
110	20	Liquid Effluent Treating	2,300,000
111	30	Steam, Cond. & Power Gen.	19,000,000
112	31	Plant Water System	10,200,00
	-----	General Facilities	25,400,000
		Contractors Costs & Fee	<u>32,000,000</u>
		Subtotal	322,000,000
		Contingency	<u>48,000,000</u>
		<u>Total Plant Investment</u>	<u>\$370,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

Dwg. No. 1950-X-102 (Unit 11)-COAL STORAGE & RECLAIMING

TABLE VI-1A

Major Equipment

Pumps	\$ 20,000
Conveying Equipment	1,060,000
Weight Scales	6,000
Other Large Equipment	
Belt Trailers	220,000
Traveling Boomstacker	700,000
Bucket Wheel Reclaimer	800,000
Transfer Car	110,000
Magnetic Separators	25,000
Samplers	15,000
Rotary Car Dumper	1,200,000
Allowance for Miscellaneous	<u>274,000</u>
Total Major Equipment	\$ 4,430,000
Installed Cost	<u>\$13,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-103(Units 12 & 14)-COAL PREPARATION &
COAL FEEDING

TABLE VI-1B

Major Equipment

Pumps	\$ 15,000
Dryer Package	2,200,000
Coal Feed Injection Systems	3,000,000
Conveying Equipment	
Belts	975,000
Feeders	595,000
Gates and Chutes	260,000
Bins	2,800,000
Mills	225,000
Weigh Scales	45,000
Separation Equipment	
Dust Collectors	55,000
Screens	175,000
Samplers	5,000
Allowance for Miscellaneous	<u>550,000</u>
Total Major Equipment	\$10,900,000
Installed Cost	<u>\$32,800,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-104A(Unit 15A)-GASIFIER, COMBUSTOR &
GASIFIER HEAT RECOVERY

TABLE VI-1C

Major Equipment

Gasifiers & Combustors	\$18,100,000
Pressure Vessels	2,610,000
Exchangers	1,130,000
Pumps	4,300,000
Waste Heat Boiler Package	6,250,000
Allowance for Miscellaneous	<u>710,000</u>
Total Major Equipment	\$33,100,000
Installed Cost	<u>\$67,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-104B(Unit 16A)- RAW GAS SCRUBBING &
AMMONIA REMOVAL

TABLE VI - 1D

Major Equipment

Columns	\$ 880,000
Pressure Vessels	185,000
Exchangers	425,000
Pumps	130,000
Hydroclones	10,000
Allowance for Miscellaneous	<u>110,000</u>
Total Major Equipment	\$1,740,000
Installed Cost	<u>\$5,400,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-105 (Unit 15B)-FLUE GAS ENERGY RECOVERY,
AIR COMPRESSION & REJECT ASH FACILITIES

TABLE VI - 1E

Major Equipment

Cyclones	\$ 610,000
Pressure Vessels	30,000
Exchangers	840,000
Compressors & Expanders	12,360,000
Rotary Feeders	15,000
Waste Heat Boiler Package	8,825,000
Rotary Ash Cooler	610,000
Bins & Hoppers	480,000
Conveyors	180,000
Allowance for Miscellaneous	<u>950,000</u>
Total Major Equipment	\$24,900,000
Installed Cost	<u>\$41,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-106 (Unit 23)-PRODUCT GAS COMPRESSION,
DRYING & H₂S REMOVAL

TABLE VI - 1F

Compressors & Blowers	\$ 8,370,000
Columns	1,655,000
Heat Exchangers	1,880,000
Pressure Vessels	370,000
Pumps	1,660,000
Allowance for Miscellaneous	<u>665,000</u>
Total Major Equipment	\$ 14,600,000
Installed Cost	<u>\$34,300,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-107 (Unit 16B)-AMMONIA RECOVERY

TABLE VI - 1G

Major Equipment

Columns	\$ 255,000
Pressure Vessels	235,000
Exchangers	100,000
Pumps	20,000
Allowance for Miscellaneous	<u>60,000</u>
	\$ 670,000
Installed Cost	<u>\$2,500,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-108A&B (Unit 21A&B)-FLUE GAS CLEANING &
SULFUR RECOVERY (CITRATE PLANT); SULFUR RECOVERY (CLAUS
& BEAVON PLANT)

TABLE VI - 1H

Major Equipment

Proprietary Packages (Installed)	\$30,700,000
Compressors & Blowers	915,000
Pressure Vessels	140,000
Exchangers	6,000
Pumps	5,000
Allowance for Miscellaneous	<u>234,000</u>
Total Major Equipment	\$32,000,000
Installed Cost	<u>\$34,500,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS
DWG. NO. 1950-X-109 (Unit 22) - SOLIDS DISPOSAL

TABLE VI - 1I

Major Equipment

Filter Package for Solids Recovery	\$ 500,000
Conveyors	25,000
Allowance for Miscellaneous	<u>35,000</u>
Total Major Equipment	\$ 560,000
Installed Cost	<u>\$2,600,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-110 (Unit 20) - LIQUID EFFLUENT TREATING

TABLE VI - 1J

Major Equipment

Columns	30,000
Pressure Vessels	8,000
API Tanks	75,000
Exchangers	95,000
Pumps	40,000
Phenol Removal Unit	1,745,000
Allowance for Miscellaneous	<u>67,000</u>
Total Major Equipment	\$2,060,000
Installed Cost	<u>\$2,300,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-111 (UNIT 30) - STEAM, CONDENSATE & POWER GENERATION

TABLE VI - 1K

Major Equipment

Tanks	\$ 15,000
Exchangers	3,455,000
Deaerator	260,000
Pumps	810,000
Power Generator Sets (Installed)	1,600,000
Allowance for Miscellaneous	<u>260,000</u>
Total Major Equipment	\$ 6,400,000
Installed Cost	<u>\$19,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

DWG. NO. 1950-X-112 (Unit 31) - PLANT WATER SYSTEM

TABLE VI - 1L

Major Equipment

Tanks	\$ 750,000
Pumps	800,000
Cooling Tower, including chlorination	2,150,000
Cooling Water Chemical Feeding System	25,000
Filters and Demineralizers	1,160,000
Potable Water Treatment System	6,000
Allowance for Miscellaneous	<u>309,000</u>
Total Major Equipment	\$ 5,200,000
Installed Cost	<u>\$10,200,000</u>

VII. OPERATING COST ESTIMATES AND GAS COST CALCULATIONS

The calculation of annual operating costs and product gas costs was carried out in accordance with the CFBraun "Gas Cost Guidelines". The results are shown in Tables VII-1 through VII-4.

Table VII-1 shows the Net Annual Operating Costs, allowing for by-product credit. For the raw materials and water, the quantities estimated were multiplied by the unit costs in the "Basis of Design", while chemicals costs were estimated or obtained from vendors. An annual on-stream time of 90% (7884 operating manhours) was used.

For labor, 50 men/shift were assumed, consistent with the CO₂ Acceptor and other similar processes. Estimating methods for the other elements of annual operating cost are shown in Table VII-1, and the calculation of Annual Maintenance Cost is presented in Table VII-2. The breakdown of Total Capital Requirement is given in Table VII-3. Based on the above capital and operating costs, the gas costs based on both utility and private financing have been calculated in accordance with the "Guidelines". The results of these calculations are presented in Table VII-4.

BATTELLE AGGLOMERATING ASH BURNER PROCESS

ANNUAL OPERATING COSTS

(IN MILLIONS OF DOLLARS)

BASIS: 90% OPERATING FACTOR (7884 hours/yr)

TABLE VII - 1

RAW MATERIALS		
Coal (532.2 tons/hr, @ \$1.00/MM BTU)		94.64
CATALYSTS AND CHEMICALS		1.80
PURCHASED WATER (4700 GPM @ \$0.40/1000 Gal)		.99
ASH DISPOSAL (63 ton/hr @ \$1.00/ton)		.49
LABOR		
Process Operating Labor (50 men/shift @ \$7.00/hr)	2.90	
Maintenance Labor (60% of Annual Maintenance Costs)	8.33	
Supervision (20% of Direct Labor)	<u>2.25</u>	
		13.48
ADMINISTRATION & GENERAL OVERHEAD (60% of Labor)		8.09
SUPPLIES		
Operating (30% of Process Operating Labor)	0.87	
Maintenance (40% of Annual Maintenance Costs)	<u>5.56</u>	
		6.43
LOCAL TAXES AND INSURANCE (2.7% of Total Plant Investment)		9.99
<u>TOTAL GROSS ANNUAL OPERATION COSTS</u>		<u>135.91</u>
BY-PRODUCT CREDITS		
Sulfur (18.3 long ton/hr @ \$25/long ton)	(3.60)	
Ammonia (0.51 ton/hr @ \$170/ton)	(0.68)	
Anhydrous Sodium Sulfate (32 ton/day @ \$52/ton)	(0.51)	
Export Steam @ \$1.00/MM BTU	<u>(0.18)</u>	
		(4.97)
<u>TOTAL NET ANNUAL OPERATION COSTS</u>		<u>130.94</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESSANNUAL MAINTENANCE COSTS(IN MILLIONS OF DOLLARS)TABLE VII-2

<u>DWG. NO.</u> <u>1950-X-</u>	<u>UNIT NO.</u>	<u>DESCRIPTION</u>	<u>UNIT COST</u>	<u>MAINTENANCE</u> <u>FACTOR % COST</u>
102	11	Coal Storage & Reclaiming	16.60	6 1.00
103	12 & 14	Coal Preparation & Coal Feeding	41.85	6 2.51
104A	15A	Gasifier, Combustor and Gasifier Heat Recovery	85.48	4.5 3.85
104B	16A	Raw Gas Scrubbing & Ammonia Removal	6.89	4.5 0.31
105	15B	Flue Gas Energy Recovery Air Compression & Reject Ash Facilities	52.30	4.5 2.35
106	23	Product Gas Compression, Drying and H ₂ S Removal	43.76	3 1.31
107	16B	Ammonia Recovery	3.19	3 0.10
108A/B	21A/B	Flue Gas Cleaning and Sulfur Recovery	44.02	3 1.32
109	22	Solids Disposal	3.32	3 0.10
110	20	Liquid Effluent Treating	2.93	3 0.09
111	30	Steam, Condensate Power Generation	24.24	1 0.24
112	31	Plant Water System	13.01	3 0.39
		General Facilities	<u>32.41</u>	<u>1 0.32</u>
-	-	TOTAL PLANT	<u>370.00</u>	<u>3.75 13.89</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

TOTAL CAPITAL REQUIREMENT

(MILLIONS OF DOLLARS)

TABLE VII - 3

TOTAL PLANT INVESTMENT	370.00
INITIAL CHARGE OF CATALYSTS & CHEMICALS	1.02
PAID UP ROYALTIES	1.53
ALLOWANCE FOR FUNDS USED DURING CONSTRUCTION (Total Plant Investment x Average Spending Period in years x 0.09)	58.28
START-UP COSTS (20 percent of Total Annual Gross Annual Operating Costs)	27.18
WORKING CAPITAL (Sum of (1) Raw Materials Inventory of 14 days at full rate, (2) Materials and Supplies at 0.9 percent of Total Plant Investment, and (3) Net receivables at 1/24 annual gas and by-products revenue at calculated sales prices.)	20.23
<u>TOTAL CAPITAL REQUIREMENTS</u>	<u>478.24</u>

NOTE: The gas sales price used in the above is based
on the average cost with utility financing.

BATTELLE AGGLOMERATING ASH BURNER PROCESS

GAS COSTS, \$/MM BTU

TABLE VII-4

First Year Gas Cost by the Utility Financing Method	2.67
Average Gas Cost by the Utility Financing Method	2.30
Constant Gas Cost by the Private Investor's Financing Method	2.82

VIII. SUGGESTIONS FOR FUTURE DEVELOPMENT ACTIVITIES

In order to substantiate Battelle's current claims for the Agglomerating Burner Process it is suggested that as a minimum, the development program outlined below be carried out using the existing PDU at West Jefferson, Ohio.

A PDU program incorporating the latest Battelle Gasifier-Combustor design should be implemented. The principal objective of this work will be to confirm the feasibility of Battelle's new mode of Gasifier operation, in which the hot ash agglomerates "rain" through the fluidized char bed to supply the heat for the coal gasification, rather than having both the ash and the char particles fluidized. The PDU must demonstrate the ability of the Gasifier to handle caking coals by dilution with large quantities of steam, char, and agglomerated ash without external pretreatment. Also, runs for the Combustor should be made to demonstrate the steady agglomeration of the ash particles and to prove the capability of the classifying section at the top of the Combustor to supply ash agglomerates of the correct size and uniformity required by the Gasifier.

Sustained, integrated runs of the PDU Gasifier-Combustor will be necessary to determine the optimum operating conditions. Materials of construction should be tested to insure uninterrupted operation of the gasification plant.

Analytical data must be collected during the PDU runs. The bottom feeding of the coal and the high operating temperature used in the Battelle Gasifier are known to yield raw gases relatively free of impurities. However, it is important to have reliable gas analyses to confirm this fact, especially for such minor components as ammonia, phenols and other organic compounds, if any. Reliable analytical data will help confirm many of the design assumptions used in this evaluation. These include:

- The correct split of the coal sulfur into H_2S in the raw fuel gas and SO_2 in the flue gas. Battelle's latest thermobalance experiments indicated that the actual split may be even higher than the 80/20 (H_2S/SO_2) assumed for the present study. A higher H_2S/SO_2 ratio will mean more sulfur recovery by the Claus plant and less by the Citrate plant. Since the Citrate plant is more costly than the Claus plant in both initial and operating costs, a higher H_2S/SO_2 ratio will result in further economy of the Battelle process.
- The Gasifier steam and carbon conversions.
- The production of phenols and ammonia and the identification of the minor organic components which may be present in the steam stripped sour water condensed out of the raw fuel gas.

The Battelle process appears to be quite competitive for producing a medium Btu fuel gas. It is believed that this process may also be suitable for making synthesis gas for chemical plants, especially for the synthesis of methanol. The additional power required to boost the product gas from 500 psig to that required by the modern low-pressure methanol processes is very moderate and the fuel gas composition can be easily adjusted to that required for methanol synthesis. Process studies should be carried out to establish a conceptual process flowsheet for the generation of syngas for use in downstream methanol production.

IX. CONCLUSIONS

This study was undertaken to study the economics of the Battelle Agglomerating Ash Process by way of a conceptual design of a commercial scale plant for producing a medium Btu fuel gas. Although the Battelle PDU tests did not produce enough data for a proper evaluation of the process and an assumed design was used for the Gasifier/Combustor, it is apparent that the process has certain important inherent advantages. The principal advantages of the Battelle process, if fully developed, would be as follows:

1. An expensive air separation plant for oxygen production is not required.
2. High overall carbon conversion for the Gasifier/Combustor unit.
3. Bottom feed of coal to the Gasifier and relatively high Gasifier operating temperature results in a raw fuel gas that is free of tar and light oils and contains very little ammonia and phenols.
4. The possible capacity for handling caking coals without external pretreatment because of the dilution by inert particles of agglomerated ash and char present in the fluidized bed Gasifier.
5. Low CO₂ content in the raw gas because heat for the Gasifier is not furnished by in-situ burning of carbon with oxygen. CO₂ removal is not required if the raw gas is used as a fuel.

6. Low $\text{CO}_2/\text{H}_2\text{S}$ ratio in the raw fuel gas makes sulfur recovery by the Claus process easily accomplished.
7. The raw gas composition appears to be suitable for methanol synthesis with only minor adjustment and treatment required.
8. The calculated thermal efficiency, 83.7%, of the process is relatively high. It must be pointed out that the fuel gas is delivered at 500 psig for distribution within a radius of about 250 miles. A competing process producing medium Btu gas at low pressure must have a thermal efficiency of close to 90% in order to allow for the energy required to compress the product gas to 500 psig.

However, it is also recognized that the Battelle process has the following disadvantages:

1. The process requires the circulation of a large quantity of hot ash between the Gasifier and Combustor. The control of this stream may present operating difficulties.
2. In order to produce a suitable agglomerate the fluidized bed Combustor must be operated within a narrow range of superficial velocity and bed temperature. Here again this may present operating problems.
3. Two streams must be treated to remove sulfur. In addition to the product gas, flue gas from the Combustor must be treated to remove SO_2 ; the process for SO_2 removal is more expensive than is the removal of H_2S .

SD's evaluation of the Agglomerating Ash Burner process as presently conceived by Battelle is that the process can be developed to produce a medium Btu gas. Furthermore, estimates of the capital (\$370,000,000 mid-'78 dollars) and gas costs (\$2.82/MM Btu) derived from the conceptual design of a commercial scale plant are attractively low. We therefore conclude that the technology warrants further study and that a development program (as outlined in Section VIII) should be carried out to demonstrate the operability and performance of the Gasifier-Combustor system. Under no circumstances should a costly pilot or demonstration plant be constructed without successful PDU operation.

Name	Approved	Date

Project Number G-6995-0003

Internal Distribution

HF Feldmann

WG Steedman

Contract/Files

February 26, 1979

Mr. Allan S. West
Project Manager
Scientific Design Company, Inc.
2 Park Avenue
New York, New York 10016

Dear Allan:

This letter is in response to the major questions raised by DOE/ORNL about your draft report on the Battelle Agglomerating Burner Process. As you know, I cannot release reports documenting some of the supporting data without approval from the sponsor. However, I have supplied the required information in as much detail as possible.

The responses to the questions listed in T. Y. Chang's memo dated February 1, 1979, are as follows:

I Gasifier

- (1) Bench-scale and limited pilot-scale operations have demonstrated that pretreatment of caking coals is not required to prevent agglomeration and defluidization in the gasifier. Although the coal particles become plastic and "sticky" as they are heated, they do not form "clumps" because they are sufficiently diluted by the hard inert particles present.
- (2) Supporting evidence for 80 percent steam decomposition comes from two sources. In bench-scale experimentation, a steam decomposition of 87 percent was achieved at conditions similar to those used in your study. Furthermore, COGAS achieves an 80 percent steam decomposition. It appears that steam-blown gasifiers, achieve relatively higher steam decompositions as a result of the water-gas shift. (Less CO₂ is formed directly from the coal, causing the water-gas shift reaction to proceed in the direction in which steam is consumed.)

- (3) The raw fuel gas composition was calculated based on the following conditions and assumptions:

- Operating Temperature - 1750 F
- Operating Pressure - 100 psig
- Steam/Carbon - 1.0
- Carbon Conversion (total) - 61.9 percent
- Steam Decomposition - 80 percent
- Sulfur in coal released as H_2S - 80 percent
(based on TGA tests)
- Nitrogen in coal released as ammonia - 10 percent
(average of similar gasifiers)
- Assumed water-gas shift equilibrium
- Calculated methane concentration from compilation of data from similar systems (Methane concentration is a function of hydrogen concentration, temperature, and pressure.)

- (4) a. Ash flux densities in the range of 19,000-47,000 lb/hr ft² were obtained in experiments by Dr. D. Geldhart. A copy of a paper presenting his findings is enclosed

b. The actual bed densities at these locations were not calculated. (It is not required for the design procedure used.) The average bed density for the entire bed is approximately 26 lb/ft³

c. The average char particle size was determined to be approximately 40 mesh. The required fluidizing velocity for such a particle would be approximately 0.043 ft/sec

d. The choice of coal grind is not too coarse for segregation and stripping as supported by the design calculations. Even though the agglomerates are nearly the same size as the largest char particles, the particle densities are quite different, 162 and 34 lb/ft³ respectively. Furthermore, most of the char particles are considerably smaller than the maximum.

- (5) "Complete" stripping of char from the ash in the char stripping zone is assured. This section operates near the minimum fluidizing velocity of the agglomerates, therefore, the bed behaves similar to a moving, packed bed of agglomerates. However, the velocity in the interstitial spaces between adjacent agglomerates would be sufficiently high to entrain even the larger particles of the less dense char.

February 26, 1979

- (6) We have never encountered clinkering in bench-scale tests, so we don't believe this is a problem. Since no oxygen is injected, there is little likelihood of hot spots.

II. Combustor

- (1) The design used in your study has not been proven in practice, however, there is no reason to suspect that it would not actually be an improvement to our original design. The statement that "Growth only occurs in the flame zone at the base where char particles burn at 2400 to 2800 F" is apparently made on an inaccurate concept of the self-agglomeration phenomenon.

Bench-scale experiments and our experience in operating the PDU clearly shows a correlation between bed height and solids capture. If agglomeration occurred only at the base of the bed, solids capture would necessarily be independent of bed height.

A direct comparison between the Agglomerating Burner and agglomerating gasifiers like Westinghouse and U-Gas does not exist. The agglomerating gasifiers must limit agglomeration to the bottom section of the bed because carbon, present in large amounts in the gasifying section of these vessels, interferes with the self-agglomeration of coal ash. However, the Agglomerating Burner is essentially free of carbon throughout. Therefore, the entire bed is available for self-agglomeration.

- (2) The PDU combustor was operated with 5 percent excess air, therefore, we would anticipate no problem in operating with this amount of excess air.
- (3) The bed will "clinker" in the combustor if defluidized before the coal feed is stopped. However, no such problems occurred in the transfer lines during normal operation.

I hope you find these responses adequate for your needs. If you have any additional questions, please call me at (614)424-4931.

Sincerely yours,

W. G. Steedman

W. G. Steedman
Research Engineer
Chemical Process Development Section

WGS:sb

Enc.

SCIENTIFIC DESIGN COMPANY, INC

MEMORANDUM *from* TYChang

TO: ASWest

DATE February 1, 1979

PROJECT NO.: 1950 BATTELLE EVALUATION

SUBJECT: QUESTIONS BY DOE/ORNL

At our January 11, 1979 meeting with DOE/ORNL a number of questions were raised about our draft report on the Battelle Agglomerating Ash Burner Process. The following is a list of the major questions and comments raised by DOE/ORNL concerning the assumed design of the Gasifier-Combustor. These points should be raised with Battelle at our forthcoming meeting.

I. Gasifier

1. Is pretreating of Illinois No. 6 coal needed before feeding into the Gasifier?
2. Supporting evidence for 80% steam decomposition in the Gasifier.
3. How was raw fuel gas composition arrived at?
4. a. Ash flux densities are 9000 Lb/Hr-Ft² (at 41' - diam. section) and 17,000 Lb/Hr-Ft² (at 30' - diam. section). What is the supporting evidence for these flux densities?
b. What are the bed densities at sections where the ash flux densities are 9000 and 17,000 Lb/Hr-Ft²?
c. What is the required fluidization velocity of the char in the Gasifier?
d. Is the choice of coal grind (minus 6 mesh) too coarse for the segregation of char and ash (8 mesh) in the Gasifier and for the complete stripping of the char from ash in the bottom stripper section?
5. Design of the char stripping section at bottom of Gasifier. Is "complete" stripping of char from ash assured?
6. What about the possibility of clinkering in the Gasifier?

II. Combustor

1. Classifying section at top of combustor - Is it proven in practice? ORNL stated - "Growth only occurs in the flame zone at the base where char particles burn at 2400 to 2800°F. Intense agitation may be needed to avoid agglomeration," and "complete redesign of combustor should be examined.....Design

Continued . . .

SCIENTIFIC DESIGN COMPANY, INC.

Memo - TYChang
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February 1, 1979
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II. Combustor (Continued)

1. (Continued)

close to Westinghouse or U-Gas should be employed where a small, intensely fluidized flame zone permits burning of the char under conditions where the particles melt and stick to existing agglomerates."

2. What is evidence for operating the PDU combustor at 5% excess air?
3. What is the possibility of clinkering the combustor?
"Sticky" agglomerates may give trouble in the transfer line from Combustor to Gasifier.

T. Y. Chang
TYChang

TYC:ASW:gac