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EVALUATION OF THE
BATTELLE AGGLOMERATING ASH BURNER
HIGH BTU COAL GASIFICATION PROCESS

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ABSTRACT

The economics of the Battelle Agglomerating Ash Burner Process for the production of high Btu gas (SNG) from coal has been evaluated. A conceptual process design including process flowsheets, heat and material balances and equipment specifications have been prepared. Capital costs, operating costs and gas costs have been developed using the CFBraun "Gas Cost Guidelines". The estimated capital and gas costs for the Battelle process have been compared to the alternative high Btu processes described in CFBraun's "Factored Estimates for Western Coal Commercial Concepts" and were found to be economically unattractive.

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I. INTRODUCTION

This volume presents an evaluation of the Battelle Agglomerating Ash Burner Process for the production of high Btu pipeline gas from coal. The evaluation covers the economics of the process, and its relative attractiveness compared to alternative high Btu gas processes. A review of the technical performance and operability of the process has not been included since operating data on this process is not available. However, a number of potential process problems and requirements have been noted herein, which must be solved or demonstrated if the process is to be technically successful and capable of continuous efficient operation.

The evaluation is based upon preliminary process information prepared by Battelle, including conceptual flowsheets with preliminary material balances. Based on this data, SD developed process flowsheets utilizing the basic Battelle process sequence but adding and modifying operations and equipment to achieve an integrated process facility including the development of balanced steam and plant water systems to service the process unit. The resultant SD process flowsheets were used as the basis for sizing and specifying equipment and making utility take-offs. In this work, the recently published estimates prepared by CFBraun⁽¹⁾ for ERDA were heavily drawn upon, supplemented by SD's own process design experience and judgement in these fields.

Based on SD's flowsheets and equipment specifications, capital and operating cost estimates have been prepared for the Battelle process and comparisons with alternative high Btu coal gasification processes are presented. The comparative economics in this

evaluation were developed in accordance with the CFBraun Guidelines and Factored Estimates, with the degree of definitiveness being appropriate to the time restraints imposed by ERDA.

The work was performed by Scientific Design Company, Inc., (SD) under its contract with ERDA to provide technical and economic evaluations of competing processes, concepts and systems for the production of high Btu gas from coal.

(1) CFBraun "Factored Estimates for Western Coal Concepts" (FE22405) October, 1976.

II. SUMMARY

The process design and economic evaluation of a 250×10^9 Btu/day pipeline quality gas plant has been prepared for the Battelle Agglomerating Ash Burner Process. The process design was based on a preliminary process flowsheet and material balance provided by Battelle, which was modified by SD as required to achieve an integrated facility. In general, the design is consistent with the assumptions in the CFBraun "Coal Gasification Commercial Concepts Gas Cost Guidelines", using an eastern coal, since Battelle's material balance is based on process development work with this coal.

The results of this study have been compared with the alternative high Btu processes described in CFBraun's "Factored Estimates for Western Coal Commercial Concepts" as shown in the following tabulation:

PROCESS	PLANT COST \$MM	GAS COST \$/MM BTU	GAS COST "EX COAL" \$/MM BTU
Battelle Agglom. Ash	\$ 1,260	\$ 5.20	\$ 3.49
With Pre-treatment	1,421	5.67	3.77
IGT HYGAS Steam-O ₂	870	2.71	2.09
IGT HYGAS Steam-Iron	1,180	4.10	3.41
Conoco CO ₂ Acceptor	890	3.38	2.72
BCR BI-GAS	1,020	3.50	2.82
Lurgi (Dual Feed)	1,060	3.30	2.40
Synthane (Export Char)	1,150	3.70	2.86

Since the alternative processes are based on western coal which is lower in cost, i.e. \$7.50 per ton vs. \$25.00 per ton or \$0.43/MM Btu vs. \$1.00/MM Btu the gas costs were recalculated for several alternative processes excluding the cost of coal as shown in the last column.

The results indicate that the capital cost for the Battelle process is higher than all of the other processes, exceeding the lowest cost HYGAS (Steam-Oxygen) and Conoco CO₂ Acceptor by about 50%. The Battelle average gas cost (utility financing method) including the cost of coal is \$5.20/MM Btu or about 50 percent greater than the average of approximately \$3.50/MM Btu for the other processes. On an "ex-coal" basis, the gas cost is still significantly higher, i.e. \$3.49/MM Btu, or about 30% greater than the average of approximately \$2.70/MM Btu for the other processes. Thus the Battelle process appears economically unattractive on the basis of this comparative evaluation.

Since the Battelle estimate is based on eastern coal an attempt was made to establish the capital cost of the plant if a western coal was used. An approximation of the differences in capital cost for the Battelle process indicates that a design based on western coal will reduce the plant cost by about \$35,000,000. This approximate 3% reduction in capital does not significantly alter the ranking of the Battelle process when compared to the other processes.

The above discussion assumes that pre-treatment of the coal is not required. If pre-treatment of the coal is required to prevent caking, the capital and gas cost price would increase as shown in the tabulation above, making the process still less attractive.

Some of the reasons for the higher capital cost and gas cost for the Battelle Process are:

1) Low Gasifier and Combustor Operating Pressure

The gasifier and combustor operating pressure of 100 psia is lower than for most currently proposed processes. This leads to the requirement for a larger number of gasifiers and combustors than for alternative processes.

2) Low Methane Make and Concentration in the Gasifier

The Battelle gasifier effluent methane concentration is quite low in comparison to the other processes in the CFBraun report. The methane concentration for the Battelle process on a wet basis is only 4.8 mol% vs. 5.7 to 11.6% for the other pro-

cesses. On a dry basis the gasifier effluent methane content is 6.8% for Battelle vs. 14.4 to 19.3% for the alternative processes. This leads to higher costs for the methanation step in the Battelle process.

In terms of the total methane make only 26.8% of the product methane is produced in the Battelle gasifier compared to 43.7 to 58.7% for the other processes.

3) Large Volume of Combustion Gases Produced and SO₂ Content

The Battelle process combustion step produces a very large quantity of hot flue gases as compared to the other high Btu processes. In addition, a major portion of the sulfur in the coal is not converted to easily handled H₂S in the gasifier but rather stays with the char and is converted to SO₂ in the combustor. SO₂ is much more expensive to recover from the gas stream and convert to sulfur than H₂S. It is for this reason that the cost of sulfur removal in the Battelle process is very high.

4) Use of Dual CO₂ Removal Units

In the Battelle process, CO₂ is removed both prior to the CO shift reaction and following methanation. Alternate processes remove CO₂ only once, that is, after the shift reaction and prior to methanation. The use of a single CO₂ removal step in the Battelle process has been evaluated in Section IX of this report.

5) Methanation Scheme

The Battelle methanation scheme utilizes a cold gas recycle stream for temperature control. This method of heat removal requires more recycle stream gas and therefore larger more expensive equipment than using an alternative such as hot gas recycle. This alternative is evaluated in Section IX of this report.

III. BACKGROUND OF AGGLOMERATING ASH BURNER PROCESS

The Battelle Agglomerating Ash Burner Gasification Process is a pressurized two-stage, fluidized-bed system involving 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 by the circulation of ash between the burner and the gasifier. It is claimed that the fluidized-bed combustion step operates under temperature/velocity conditions such that the fly ash in the coal is agglomerated or pelletized 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.

Initial development of the Agglomerating Burner Gasification Process was conducted at Battelle in the early and mid-sixties under sponsorship of the Union Carbide Corporation. This work included bench-scale and pilot-scale studies of the self-agglomerating coal burner and bench-scale study of the gasifier operation. The erosion caused by fly ash from the burner at simulated turbine conditions was also studied. All of the Union Carbide supported work was done at atmospheric pressure.

In December 1969, Battelle submitted a proposal to the Office of Coal Research (OCR) to design and construct a Process Development Unit (PDU) to further develop the process.

In September 1972, Battelle initiated work based on an interim authorization provided by the American Gas Association (AGA) in advance of a formal contract award by OCR. In January 1973 a contract was awarded for a project consisting of three phases:

- Phase I

Design, procurement and construction of a 100 PSIG
PDU with a synthesis gas capacity of 800 MSCF per day

- Phase II

Start-up and operation of the PDU gasifier and burner sub-
systems to study the agglomerating behavior of the coals.

- Phase III

Continuous operation of the integrated PDU to establish the
process variables.

The total estimated cost for this three-phase project was \$4,100,000 with completion scheduled for June 1975.

The design, procurement and construction of the PDU (Phase I) was sub-contracted by Battelle to Chemical Construction Corporation (Chemico) in February, 1973. In August 1974 this project was dropped from joint ERDA-AGA funding and ERDA continued to fund on its own. Because of delays in the completion in the PDU the prime contract with Battelle has been amended six times. The current cost of the project is \$12,900,000 and completion date of July 31, 1977 was projected.

IV. BASIS OF DESIGN

TECHNICAL INFORMATION PROVIDED BY BATTELLE

The following information was provided to SD by Battelle via ERDA:

1. A set of preliminary flowsheet sketches (undated). These flowsheets provided a material balance but no equipment sizes.
2. Revisions to the above flowsheets and material balance dated 3/7/77, together with an overall schematic flowsheet, DWG No. 2024003, Rev. 1.
3. Several telephone communications with ERDA (OPPA) to clarify technical questions concerning the process.
4. In addition to the above, SD reviewed the information on the process contained in the 4th, 5th, and 6th Synthetic Pipeline Gas Symposia held in Chicago, Illinois in 1972, 1973 and 1974, respectively.

DESIGN PARAMETERS

The general design parameters used in this work follow the information provided in the "Coal Gasification Commercial Concepts Gas Cost Guidelines" prepared by CFBraun in January, 1976.

The Battelle flowsheets were reviewed, revisions made where required to achieve an integrated design, and equipment was sized and specified.

The specific design assumptions used by Battelle in each plant unit are presented in Section VI, PROCESS DESIGN CRITIQUE together with SD's modifications and suggested improvements.

The major design parameters excerpted from the CFBraun "Guidelines" are summarized below:

COAL PROPERTIES: The plant design is to be based on eastern coal with the following properties:

Proximate Analysis, as received, wt. %

Moisture	6.0
Volatile Matter	31.9
Fixed Carbon	51.5
Ash	10.6
	<u>100.0</u>

Ultimate Analysis (dry), wt. %

Carbon	71.50
Hydrogen	5.02
Nitrogen	1.23
Oxygen	6.53
Sulfur	4.42
Ash	11.30
	<u>100.00</u>

Heating Value of Dry Coal, Btu/lb (HHV) 13,190

Heating Value of Coal as Received, Btu/lb (HHV) 12,400

PLANT SIZE: The design capacity of the commercial plant is 250 billion Btu/stream day (HHV) of synthetic natural gas.

PRODUCT GAS SPECIFICATION: Limits for carbon monoxide and sulfur are as follows:

Cabon Monoxide	0.1 volume percent
Hydrogen Sulfide	0.25 grains/100 SCF
Total Sulfur	10 grains/100 SCF

SERVICE FACTOR: Service factor of the total plant to be 90 percent (7884 hrs/yr).

GRASS ROOTS FACILITY: The plant will be self-contained, except for water and coal. It will have cold start capability. Natural gas is available at start-up for critical

needs, excluding steam generation.

EQUIPMENT SELECTION: The approach to be taken for equipment selection and sizing shall be on the conservative side. This means that wherever possible proven equipment in the same or similar services shall be used. When this is not possible, because of size, pressure, temperature, or other service conditions, the vendor of similar equipment must be contacted to assist in establishing a reasonable extrapolation of current technology.

PLANT SITES: The plant sites for eastern coal will be in the states that are tabulated in the Guidelines. In all cases, the plant will be located near the mine. No barge or other form of water transportation will be available to the plant sites.

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

SEISMIC CONDITIONS: All plant are located in Seismic Zone I.

CLIMATIC CONDITIONS: The plant design is to be based on the following eastern location 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

CONSTRUCTION LABOR CONDITIONS: Average construction labor rates for eastern plant sites are included in the unit costs list. Construction labor, travel costs and living quarters should be added in addition to the normal indirect labor costs for fringe benefits, FICA, FUI, and SUI taxes, and Workmen's Compensation insurance. Labor productivity may vary widely over the eight states covered by the Eastern region. For consistency in developing gas cost comparisons, a labor productivity factor of 1.0 shall be assumed.

COAL FEED: The coal will be received at plot edge on a conveyor belt and will not require any on-plot beneficiation such as washing. The coal will be received on a six days per week and two shifts per day basis.

COAL STORAGE: Storage and handling facilities will be provided for a 14-day supply of coal to the plant.

COAL FEED PREPARATION: Spare mills and classifying equipment will be provided in an amount equivalent to a minimum of 25 percent of total plant capacity. An intermediate coal storage time of eight hours shall be provided for temporary repair of elevators and conveyors. This requirement does not apply if the elevators or conveyors are part of a spare system or line.

WATER STORAGE: A water storage pond containing seven days water needs shall be included.

GASIFICATION: A minimum of two gasifiers will be provided for all processes. For those processes requiring air for gasification, separate air compressors and hot gas expander drives will be provided for each gasifier. Similarly, facilities for gas quench and solids removal will be separate for each gasifier.

GAS TREATMENT: Two trains, 50 percent of total plant capacity each, will be provided for the following major process units. Fixed catalyst beds will be designed for 25 percent of plant capacity, two beds for each 50 percent train.

Shift Conversion
Acid Gas Removal
Methanation

Product Compression
Product Gas Drying

LIQUID EFFLUENT: Process liquid effluent treating will have two trains, 50 percent of the total plant capacity each. One common spare sour water stripper with its auxiliary equipment will be provided.

SOLID EFFLUENT: Solid effluent treating will be one train with multiple equipment items. Spare equipment will be provided to maintain full plant capacity.

GAS EFFLUENT: All gaseous effluents from the plant must meet the effluent standards described in these Guidelines. Facilities for recovering sulfur from effluent gas streams may be provided in single train or in multiple parallel trains. One spare train will be provided to maintain full plant capacity.

WATER WASTES: No water may be discarded to surface runoff. All water wastes must be utilized in the plant. The only water discarded will be with solid waste disposal or for dust control.

BOILER PLANT AND COAL FIRED SUPERHEATERS: To determine total installed capacity, two rules will be followed:

- A. Total installed capacity will be sufficient to handle start-up of second train with one train in operation.
- B. During normal operation of total plant, total boiler capacity will be sufficient to permit shut down of one boiler.

BOILER FLUE GAS: Each boiler, if direct coal fired, will be equipped with a separate absorber. Two regenerators, each 100 percent, will be installed. If fired with low Btu

gas, boiler flue gas cleanup is not required.

POWER GENERATION: Power plant will consist of multiple generators with one spare.

EQUIPMENT SPARING: Equipment will be spared in accordance with the following:

- A. Large compressors and expanders will not be spared.
- B. Other rotating equipment will be spared. The spare may be either a complete spare or a 50 percent spare in cases where two pumps are normally operating.
- C. Spare rotors will be provided for centrifugal compressors and their turbine drivers. Spare rotors for turbine generators will not be included.
- D. Belt conveyors and bucket elevators will not be spared.

INSTRUMENTATION: No computer controlled instrumentation.

MAINTENANCE SHOP: A shop capable of major maintenance will be included in capital costs of the plant.

USE OF AIR-FIN COOLERS: Maximum practical use of air-fin coolers shall be made to minimize plant water requirements. Air-fin units shall be used for all steam turbine surface-condensers, and for all process cooling and condensing services above 130°F.

UNIT COSTS AND PRICES: Unit Costs and Prices as of January 1, 1976 for an eastern location are tabulated below:

Raw Water, Dollars/Thousand Gallons	\$ 0.40
Design Coal, Dollars/Short Ton as Received	25.00
Design Coal, Dollars/Million Btu as Received	1.007
Export Power, Dollars/KWH	0.01
Sulfur, Dollars/Long Ton	25.00
Ammonia (Anhydrous), Dollars/Short Ton	170.00
Light Oil (300-700°F), Dollars/Gallon	0.40
Tar, Dollars/Gallon	0.20
Operating Labor, Dollars/Hour	6.70
Construction Labor, Dollars/Hour	9.50
Ash Disposal, Dollars/Short Ton of Ash	1.00

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

1. 20-year project life
2. 20-year straight line depreciation on plant investment, allowance for funds used during construction, and capitalized portion of start-up costs.
3. Debt-equity ratio of 75/25
4. Percent interest on debt of 9 percent
5. Percent return on equity of 15 percent
6. Federal income tax rate of 48 percent

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

1. 20-year project life
2. 16-year sum of the years digits depreciation on total investment
3. 100 percent equity capital
4. 12 percent DCF return rate
5. 48 percent federal income tax rate

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 items.

The overall process facility is divided into the following major units or sections using a numbering system consistent with CFBraun's work.

<u>Unit No.</u>	<u>Flowsheet Dwg. No. 1887-1-</u>	<u>Description</u>
---	101	Block Flow Diagram
11	102	Coal Storage and Reclaiming
12	102	Coal Preparation
14	102	Coal Feeding
15	103A/B	Gasification and Combustion
16	103A	Raw Gas Quench
17	104	Shift Conversion
18	104	Acid Gas Removal
19A	105	Methanation
19B	106	Final CO ₂ Removal and Gas Compression
20	107	Liquid Effluent Treating
21	104	Sulfur Recovery
22	108	Solids Disposal
23	106	Product Gas Drying
30	110	Steam, Condensate & Power Generation
31	109	Plant Water System

The overall process scheme is illustrated in the Block Flow Diagram, DWG-1887-101 showing the interrelation between the various units. It should be noted that in both the block flow diagram and the Description of Flow which follows, Unit 18-Acid Gas removal precedes Unit 17-Shift Conversion since the sequence of operations in the Battelle process differs from most other processes.

For consistency, the equipment designations used on the flowsheets and the equipment list are the same as in the CFBraun work. The letters following the equipment number indicate the number of duplicates in a single train, while the final number in parentheses indicates the number of parallel trains. Thus CY-1501 A/F (4) indicates the following:

CY	=	Cyclone
1501	=	1st cyclone in Unit 15
A/F	=	6 duplicate cyclones in a train
(4)	=	4 trains in parallel

DESCRIPTION OF FLOW

Unit 11 - Coal Storage and Reclaiming

Raw coal, nominally sized at 1 1/4" x 0" and with nominal 6 percent moisture, enters the plant on a belt conveyor from the mine. Tramp iron is removed by a magnetic separator and the coal is sampled and crushed for lab analysis. The coal is distributed to four storage piles, with 14 days total capacity, by a boomstacker. A bucket reclaimer removes coal from the storage piles and the coal is sent by conveyor to Unit 12 after being subjected to a second electromagnetic tramp iron removal. Runoff from the coal piles is collected in a sump and pumped to effluent treating.

Unit 12 - Coal Preparation

The coal received by conveyor is dried to about 1.5 percent moisture in a direct-fired rotary dryer (D-1201) system. The dry coal is stored in storage bin (V-1201) for metered distribution to a series of Coal Mills (typically, cage mills) GR-1201, and fed to the Screen Feed Bin (V-1202). The oversized mill product is separated on 8 mesh screens (SC-1201) and recycled to the surge bin in feeding the mills. The coal passing 8 mesh is separated in a Air Separator (X-1203) into approximately 83 percent - 8 to + 100, and 17 percent -100 mesh fractions, the former being conveyed to the Gasifier lockhoppers and the latter to the Combustor lockhoppers, both in Unit 14.

Unit 14 - Coal Feeding

The coal feed streams to gasification and combustion are each surged in Distribution Bins V-1223 and V-1203 respectively and metered to four sets of alternating lockhopper systems, one set matched to each Gasifier and each Combustor. The lockhopper system is comprised of Lockhopper Bins (V-1404 & V-1424), Lockhoppers (V-1405 and V-1425) and Feeders (FD-1408 and FD-1428). In each set of lockhoppers one is alternately fed from surge and pressurized with conveying gas while the other is feeding the gasifier or combustor through rotary air lock feeders.

Unit 15 - Gasification and Combustion and Unit 16 - Raw Gas Quench

This section of the plant contains four identical trains. In this description reference will be made to items in only a single train.

The Agglomerating Ash Burner Process utilizes a pressurized (100 psia) two-stage fluidized bed system involving combustion of char and/or coal in a fluidized bed Combustor (R-1502) and the steam gasification of coal in a separate fluidized bed Gasifier (R-1501). The heat for the gasification reaction is provided by the recirculation of ash between the Combustor and Gasifier.

The coal fed to the Combustor will be minus 100 mesh and that to the Gasifier will be minus 8 plus 100 mesh. Fresh coal feed and recycle char from the Gasifier are burned with air in the Combustor in a manner which allows the agglomeration of the ash at a temperature (2100°F) approaching the fusion point.

The hot agglomerated ash particles overflowing the Combustor are transferred continuously to the Gasifier by use of a steam lift. Ash is continuously removed from the Gasifier at the coolest point and recycled back to the Combustor by means of an air lift. A portion of this recycled ash, equal to the ash content on the total coal feed to the Gasifier/Combustor system, is continuously discharged through a lockhopper system for subsequent disposal.

The gas leaving the Gasifier at 1700°F is passed through the Gasifier Cyclones (CY-1501) to remove the major portion of the entrained ash particles and then to the Gasifier Waste Heat Boiler (E-1601) in which it will cool to 300°F before entering the Quench Tower (C-1601) where it is further cooled to 100°F. The lower portion of the Quench Tower serves as a scrubber to remove the fine ash particles. The Quench Tower Bottom Recycle pumps (P-1601) recirculate water condensed in the upper sections through a sparger system over an open gridwork of packing.

Ash slurry from the Quench Tower bottoms is cooled in the Slurry Cooler (E-1604) and pressured to the Ash Slurry Tank (V-1510). In the upper section of the Quench Tower the steam is condensed by the countercurrent flow of hot gas and cooled condensate. Condensed water is trapped in a chimney type tray and circulated by the Quench Tower Top Reflux Pumps (P-1602) through the Reflux Air Cooler (A-1603) and the Reflux Cooler (E-1606) to cool the condensate to 95°F. This cooled condensate is introduced to the Quench Tower middle section comprised of sieve trays for condensing the steam and to the upper section to condense impurities in the gas. Sour water obtained in the upper section is trapped out and sent to the effluent treating unit. The cooled gas at 100°F and 90 psia is introduced to Compressor Suction Drum, V-1603, to remove entrained liquid and compressed in the 1st Stage Compressor, K-1601, to 250 psia and then cooled to 100°F by 1st Stage Air Cooler, A-1601, and 1st Stage Trim Cooler, E-1607. The cooled gas is fed to 1st Stage K.O. Drum, B-1602, to remove condensate and compressed to 500 psia by the 2nd Stage Compressor, K-1602. Sour water effluent from the K.O. Drums is discharged to the effluent treating section. The compressed gas is then introduced to the Acid Gas Removal Unit.

The gas leaving the Combustor (R-1502) at 2100°F will be passed through Cyclones, CY-1502 and a Dust Collector (DC-1501) incorporating small multi-cyclones (CY-1506) to remove the dust. Dust is collected in Dust Bin, V-1501, for intermittent disposal. The hot gas then passes through the No. 1 Waste Heat Boiler (E-1502) to cool it to 1500°F prior to in-

roduction into the Flue Gas Expander (TE-1501) where it will expand and produce power to drive the Air Compressor (K-1502). The expander exit gas at 890°F and 17 psia will be used to generate additional steam and preheat boiler feed water in the No. 2 Combustor Waste Heat Reboiler (E-1506) and Combustor Flue Gas Cooler (E-1508) respectively. The cooled exit gases at 450°F are sent to Sulfur Recovery (Unit 21).

A portion of the expanded flue gas is compressed by the Lift Gas Compressor (K-1501) to 120 psia for use as a conveying gas. As an alternate to flue gas, vent CO₂ from Unit 19B (CO₂ Removal) can be utilized as a conveying gas.

Reject ash will be removed primarily from the Gasifier (R-1501) and collected in Reject Ash Hopper V-1505. It will gravity feed into the Fluidized Ash Cooler (V-1503) where it will be cooled by direct contact with a stream of cooled condensate circulated by Condensate Pump (P-1501). The cooled dry solids will be discharged to Ash Disposal Hopper (V-1504) for removal by pneumatic conveying to solids disposal. The steam generated in the Fluidized Ash Cooler will be condensed with boiler feed water in Cooler Condenser, E-1507. The inerts from this system will be vented to the Sulfur Recovery section. A portion of the condensate containing dust carry-over will be pumped by the Condensate Pump to the Ash Slurry Tank (V-1510) where it is mixed with the bottom slurry from the Quench Tower. The slurry from the Ash Slurry Tank will then be pumped by the Ash Slurry Pump (P-1505) to the Solids Disposal Unit.

Unit 18 - Acid Gas Removal

Prior to shift and methanation the raw gasifier off-gas must be treated in the Acid Gas Removal Unit to reduce the hydrogen sulfide to below 1 ppm by volume. This is accomplished using the proprietary Benfield Process. The regenerated hydrogen sulfide together with carbon dioxide is sent to Sulfur Recovery.

Unit 17 - Shift Conversion

The purified gas from the Acid Gas Removal System is split into two streams. One stream containing approximately 76 percent of the total gas, bypasses Shift Conversion and is sent

directly to Methanation. The remaining 24 percent of the gas is shift converted in a 2-stage conversion system. Four sets of 25 percent fixed bed converters, R-1701 and R-1702 are used. The Shift Converter feed gas is first preheated, in a series of Feed-Effluent Exchangers to 726°F. It is then mixed with 550°F steam and the steam-gas mixture enters the first converter at 650°F. A Start-up Heater (H-1901), common for both the Shift Conversion and Methanation Units, is provided to supply heat during start-up operations.

Gas leaves the First Shift Converter (R-1701) at 829°F, and is cooled to 631°F in Feed-Effluent Exchanger E-1701 before entering the Second Shift Converter (R-1702). The converted gas leaves the Second Converter at 707°F and is cooled to 570°F in Feed-Effluent Exchanger, E-1702. The converted gas then mixes with the by-passed portion of the purified gas from the Acid Gas Removal Unit, and the mixture is delivered to the Methanation Unit at 290°F and 470 psia.

Unit 19A - Methanation

Four Methanation trains (25% of plant capacity each) are required, but only one train will be described. The combined shifted and unshifted effluents enter the ZnO Bed Preheater, (E-1902) where they exchange heat with the Third Methanation Reactor (R-1903) effluent gas. The gas stream at 500°F then enters the ZnO Bed (R-1801). The objective of increasing the gas temperature to 500°F is to maximize the sulfur take-up of the ZnO beds.

The gas stream, now free of sulfur, enters the methanation train at four different points. Each reactor train consists of four massive packed bed reactors in series. A portion of fresh feed gas is combined with the required amount of preheated recycle gas and enters the first Methanation Reactor (R-1901) at 550°F. The effluent from the first reactor is mixed with another portion of fresh feed and cold recycle gas to lower the gas temperature. The mixture is then fed to the Second Methanation Reactor (R-1902). This procedure is continued to the Third Methanation Reactor (R-1903). Part of the third reactor effluent is used to preheat to reaction temperature the recycle gas to the first bed, by means of the

Gas Cooler, E-1901. The third Methanation bed effluent is further cooled, as mentioned earlier, by exchanging heat with the ZnO Bed feed in E-1902. The total stream leaving E-1901 is fed to the Fourth Methanation Reactor (R-1904).

A direct fired start-up heater H-1901 is provided in order to raise the temperature of the feed gas to the ZnO beds to 550°F when the methanation train is starting up.

The effluent gas from each Fourth Methanation Reactor is cooled from about 710°F to 365°F in a Waste Heat Boiler (E-1903) which generates superheated steam at 500 psia and 550°F. The Waste Heat Boiler effluent is cooled to about 240°F by heat exchange with steam condensate which is heated from 134°F to 346°F, in Boiler Feed Water Heater E-1904. Condensed water is separated from the cooled gas in KO drum V-1901 and sent to the plant water system for recovery. The net methanated gas production is sent to the Final CO₂ Removal Unit and the remaining gas is recycled back to the methanator train by means of Recycle Compressor, K-1901, which is designed to handle the total recycle stream of the four methanation trains. The machine is driven by a steam turbine using superheated steam generated in the plant.

Unit 19B - Final CO₂ Removal and Product Gas Compression

The gas from methanation is processed for final CO₂ removal and compression in two parallel 50% trains. The CO₂ absorption system utilizes a hot activated potassium carbonate solution. The feed gas enters the CO₂ Absorber (C-1901) at 240°F and 430 psia and the scrubbed gas leaves at 158°F with 99 percent of the CO₂ in the gas feed removed.

The bottoms from the CO₂ Absorber is pressured to the Carbonate Regenerator (C-1902) for regeneration of the potassium carbonate solution which is returned to the Absorber. The Carbonate Regenerator is provided with Regenerator Reboiler (E-1906) and Regenerator Air Condenser (A-1901). The regenerator overhead is separated in the Regenerator Reflux Drum (V-1902) into a condensate stream for reflux and recovery and a pure hydrogen sulfide free carbon dioxide stream to be used as a conveying gas.

The product gas from the CO₂ Absorbers is compressed in Product Gas Compressor, K-1902, from 420 psia to 1020 psia. Coolers E-1907 and E-1908 and K.O. Drums V-1904 and V-1905 are provided before and after the compressors. The compressed product gas is sent to Unit 23 for drying before delivery to the pipeline.

Unit 20 - Liquid Effluent Treating

Sour water from the Quench Tower (E-1601) and the Acid Gas Removal Unit 18 is sent to the Condensate Oily Water Separator V-2010. After any oils and tars have been separated out and sent to off-site storage, dissolved gases are stripped out in the Condensate Stripper, C-2010. Heat is provided to the Condensate Stripper by the Condensate Stripper Reboiler, E-2016. The overhead from the Condensate Stripper is sent to the Ammonia Recovery Unit (X-2001) which uses the Phosam - W process a proprietary process of USS Engineers, Inc. The bottoms from the Condensate Stripper is pumped to the Biooxidation Unit via the Condensate Stripper Feed Bottoms Exchanger (E-2012), Stripper Condensate Air Cooler (A-2012) and Water Cooler (E-2018). Water recovered from the Biooxidation Unit is reused in the plant.

Unit 21 - Sulfur Recovery

The Sulfur Recovery System treats both the SO₂ - containing combustor flue gas from Unit 15, and the stripped acid gas containing H₂S from the Acid Gas Removal Unit 18, recovering sulfur from these streams using proprietary processes. Unit 21 consists of a Citrate Unit, for SO₂ removal and a Stretford Unit for H₂S removal.

Combustor flue gas leaving the Combustor Flue Gas Cooler (E-1508) at 450°F is cooled to 250°F in the Exchanger E-2101 by interchange with treated flue gas leaving the Acid Gas Removal Section. It then enters the Sulfur Recovery unit where it is treated to produce sulfur by-product and a scrubbed gas with an SO₂ content meeting environmental requirements. The gas is sent to the Stack (S-2101) via E-2101, and the sulfur is sent to storage.

Unit 22 - Solids Disposal

All solids bearing water streams from the process are thickened by gravity settling, then filtered. The dewatered solids are mixed with the cooled ash purge in a pug mill and delivered to disposal facilities. Recovered water is reused in the plant water system.

Unit 23 - Product Gas Drying

Prior to pipeline entry the cooled final gas product is dehydrated using a recirculating stream of triethylene glycol.

Unit 30 - Steam, Condensate and Power Generation

This system includes equipment for boiler feed water treatment and pumping, condensate and blowdown flash drums, start-up boiler and power generation and condensate recovery.

Unit 31- Plant Water System

The plant water system consists of a water reservoir, firewater system, water filtering system, cooling tower and pumping facilities, demineralizer and chemical treatment facilities.

BATTELLE AGGLOMERATING ASH BURNER PROCESS

SUMMARY MATERIAL BALANCE - BASE CASE

TABLE 5 - 1A

<u>INLET STREAMS</u>	<u>LBS/HR</u>	<u>PERCENT OF TOTAL</u>
Coal, Dry	1,335,000	15.6
Water in Coal	85,000	1.0
Combustion Air, Nitrogen	4,258,000	49.7
Combustion Air, Oxygen	1,281,000	15.0
Raw Water	<u>1,602,000</u>	<u>18.7</u>
Total	8,561,000	100.0
 <u>OUTLET STREAMS</u>		
Product Gas	452,000	5.3
CO ₂ Vent	225,000	2.6
Cooling Tower Losses	600,000	7.0
Flue Gas	6,685,000	78.1
Ammonia	3,700	Nil
Sulfur	58,300	0.7
Waste Solids, Dry	165,000	1.9
Water in Waste Solids	82,000	1.0
Steam and Water Losses	<u>290,000</u>	<u>3.4</u>
Total	8,561,000	100.0

BATTELLE AGGLOMERATING ASH BURNER PROCESS

SUMMARY MATERIAL BALANCE - PRE-TREATMENT ALTERNATE

TABLE 5 - 1B

<u>INLET STREAMS</u>	<u>LBS/HR</u>	<u>PERCENT OF TOTAL</u>
Coal, Dry	1,485,000	14.3
Water in Coal	95,000	0.9
Combustion Air, Nitrogen	5,372,000	51.7
Combustion Air, Oxygen	1,616,000	15.5
Raw Water	<u>1,832,000</u>	<u>17.6</u>
Total	10,400,000	100.0

OUTLET STREAMS

Product Gas	452,000	4.3
CO ₂ Vent	225,000	2.2
Cooling Tower Losses	660,000	6.3
Flue Gas	8,360,000	80.4
Ammonia	4,000	Nil
Sulfur	64,000	0.6
Tars	40,000	0.4
Waste Solids, Dry	184,000	1.8
Water in Waste Solids	91,000	0.9
Steam and Water Losses	<u>320,000</u>	<u>3.1</u>
Total	10,400,000	100.0

BATTELLE AGGLOMERATING ASH BURNER PROCESS

SUMMARY ENERGY BALANCE - BASE CASE

TABLE 5-2

<u>ENERGY INPUT</u>	<u>MM BTU/HR</u>	<u>PERCENT OF TOTAL</u>
Coal to Process, HHV	17,600	100.0
Total Input	<u>17,600</u>	<u>100.0</u>
<u>ENERGY DISTRIBUTION</u>		
Product Gas, HHV	10,420	59.2
By-Products, HHV		
Ammonia	35	0.2
Sulfur	230	1.3
Export Power, HHV	<u>340</u>	<u>1.9</u>
Subtotal, Product By-Products & Export Power	11,025	62.6
Consumption and Losses	<u>6,575</u>	<u>37.4</u>
Total Distribution	17,600	100.0
<u>COLD GAS EFFICIENCY, PERCENT</u>		<u>59.2</u>
<u>PLANT THERMAL EFFICIENCY, PERCENT</u>		<u>62.6</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

SUMMARY ENERGY BALANCE - PRE-TREATMENT ALTERNATE

TABLE 5-3

<u>ENERGY INPUT</u>	<u>MM BTU/HR</u>	<u>PERCENT OF TOTAL</u>
Coal to Process, HHV	19,600	100.0
Total Input	<u>19,600</u>	<u>100.0</u>
<u>ENERGY DISTRIBUTION</u>		
Product Gas, HHV	10,420	53.2
By-Products, HHV		
Ammonia	37	0.2
Sulfur	256	1.3
Tars	798	4.0
Export Power, HHV	440	2.2
Subtotal Product, By-Products & Export Power	11,951	60.9
Consumption & Losses	7,649	39.1
Total Distribution	<u>19,600</u>	<u>100.0</u>
<u>COLD GAS EFFICIENCY, PERCENT</u>		<u>53.2</u>
<u>PLANT THERMAL EFFICIENCY, PERCENT</u>		<u>60.9</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

MAJOR EQUIPMENT LIST

Unit 11 - Coal Storage and Reclaiming

<u>Item No.</u>	<u>Item Name</u>
CH-1101	Sample Rejects Chute
CH-1102	Diverting Chute
CH-1103	Secondary Sample Rejects Chute
CO-1101	Cross Conveyor No. 1
CO-1102	Cross Conveyor No. 2
CO-1103	Cross Conveyor No. 3
CO-1104A/B	Bedding and Reclaim Conveyors
CO-1105	Collecting Conveyor
FD-1101	Vibrating Feeder
FD-1102	Belt Feeder
GR-1101	Sample Crusher
P-1101A/B	Storage Area Sump Pump
MS-1101	Electronic Separator
MS-1102	Electronic Separator
V-1101	Surge Hopper
V-1102	Emergency Reclaim Hopper
W-1101	Belt Scale
X-1101	Electric Hoist
X-1102	Calibration Chain and Reel
X-1103	Automatic Belt Sampler
X-1104	Rotary Sampler
X-1105A/B	Belt Trailers No. 1 and 2
X-1106	Travelling Boomstacker
X-1107	Bucket Wheel Reclaimer
X-1108	Transfer Car
X-1109	Sample Bucket Elevator
X-1110	Secondary Belt Sampler
X-1111	Electric Hoist Driven Trolley
X-1112	Lightning Arrestors
X-1113	Sprinklers
X-1114	Front End Loader

MAJOR EQUIPMENT LIST

Unit 12 - Coal Preparation

GRINDING SECTION

Item No.

Item Name

CH-1201A/D
CH-1202A/D
CH-1203A/Z
CH-1204A/Z
CH-1206
CH-1207
CH-1215A/B
CH-1217

Cage Mill Feed Chute
Cage Mill Discharge Chute
Screen Feed Chute
Screen Oversize Chute
Oversize Return Conveyor Chute
Oversize Transfer Conveyor Chute
Oversize Discharge Conveyor Chute
Oversize Discharge Cross Conveyor Chute

CO-1201
CO-1202
CO-1214
CO-1215

Plant Feed Belt Conveyor
Mill Discharge Conveyor
Oversize Discharge Cross Conveyor
Oversize Return Conveyor

D-1201A/B/C

Dryer Package

FD-1201A/D

Belt Feeder

GR-1201A/D

Coal Mill

SC-1201A/Z

Screen

V-1201
V-1201

Storage Bin
Screen Feed Bin

W-1201A/D

Belt Scale

X-1202

Electric Overhead Bridge Crane

COMBUSTOR TRAIN

CH-1205A/Z
CH-1208A/B
CH-1209(4)
CH-1213A/B/C

Screen Undersize Chute
Undersize Collecting Conveyor Chute
Distribution Bin Discharge Chute
Sampler Discharge Chute

CO-1205A/B
CO-1206

Undersize Collecting Conveyor
Distribution Bin Feed Conveyor

FD-1202 (4)
FD-1207(4)

Distribution Bin Discharge Vibratory Feeders
Belt Feeder

MAJOR EQUIPMENT LIST

Unit 12 - Coal Preparation (Con't)

V-1203

W-1202(4)

X-1203A/D

X-1204(4)

X-1206A/B/C

X-1207A/B/C

Item No.

CH-1229(4)

CH-1233A/B

CO-1226

FD-1222(4)

FD-1227(4)

V-1223

W-1222(4)

X-1224(4)

X-1226A/B/C

X-1227A/B/C

BL-1201A/B

DC-1201A/B

P-1201A/B

P-1202A/B/C

TK-1201

X-1215A/B/C

COMBUSTOR TRAIN (Con't).

Distribution Bin

Belt Scale

Air Separator Package

Bin Gates

Primary Sampler

Secondary Sampler

GASIFIER TRAIN

Item Name

Distribution Bin Discharge Chute

Sampler Discharge Chute

Distribution Bin Feed Conveyor

Distribution Bin Discharge Vibratory Feeders

Belt Feeder

Distribution Bin

Belt Scales

Bin Gates

Primary Sampler

Secondary Sampler

MISCELLANEOUS

Dust Collector Fan

Dust Collector

Dust Collector Slurry Pump

Cleanup Sump Pump

Pump Sump

Space Heaters

MAJOR EQUIPMENT LIST

Unit 14 - Coal Feeding

COMBUSTOR TRAIN

Item No.

Item Name

FD-1408A/B(4)

Combustor Feeders

V-1404(4)

Lockhopper Bin

V-1405A/B(4)

Combustor Lockhoppers

GASIFIER TRAIN

FD-1428A/B(4)

Gasifier Feeders

V-1424(4)

Lockhopper Bin

V-1425A/B(4)

Gasifier Lockhoppers

PRE-TREATMENT ALTERNATIVE

The following items will be added to Unit 14 if coal pre-treatment is required.

Item No.

Item Name

A-3002

Pre-treatment Turbine Condenser

BL-1401

Pre-treatment Air Blower

BL-1402(4)

Recycle Gas Blower

DC-1401(4)

Venturi Scrubber

E-1401

Incinerator Waste Heat Boiler Package

H-1401(4)

Pre-treater Startup Heater

H-1402

Pre-treatment Waste Incinerator

P-1401(4)

Venturi Circulation Pump

R-1401(4)

Pre-treater

TG-1401

Pre-treatment Turbine/Generator Package

V-1401(4)

Pre-treater Tar/Water Separator

MAJOR EQUIPMENT LIST

Unit 15 - Gasification & Combustion

<u>Item No.</u>	<u>Item Name</u>
CY-1501A/F(4)	Gasifier Cyclones
CY-1502A/H(4)	Combustor Cyclones
CY-1503(4)	Dust Cyclone (in V-1501)
CY-1504(4)	Cooler Cyclone (in V-1503)
CY-1506(4)	Dust Collector Cyclone (in DC-1501)
DC-1501(4)	Dust Collector
E-1502(4)	No. 1 Combustor Waste Heat Boiler
E-1504(4)	Lift Gas Cooler
E-1506(4)	No. 2 Combustor Waste Heat Boiler
E-1507(4)	Cooler Condenser
E-1508(4)	Combustor Flue Gas Cooler
F-1501(4)	Filter/Silencer
FD-1501A/B(4)	Rotary Feeders
K-1501(4)	Lift Gas Compressor
K-1502(4)	Air Compressor
P-1501A/B(4)	Condensate Pump
P-1505(4)	Ash Slurry Pump
R-1501(4)	Gasifier
R-1502(4)	Combustor
TE-1501A/D(4)	Flue Gas Expanders
V-1501(4)	Dust Bin
V-1502(4)	No. 1 Combustor WHB Steam Drum
V-1503(4)	Fluidized Ash Cooler
V-1504(4)	Ash Disposal Hopper
V-1505A/B(4)	Reject Ash Hoppers
V-1510(4)	Ash Slurry Tank
V-1511(4)	No. 2 Combustor WHB Steam Drum

MAJOR EQUIPMENT LIST

Unit 16 - Raw Gas Quench

<u>Item No.</u>	<u>Item Name</u>
A-1601(4)	First Stage Air Cooler
A-1603(4)	Reflux Air Cooler
C-1601(4)	Quench Tower
E-1601(4)	Gasifier Waste Heat Boiler
E-1604(4)	Slurry Cooler
E-1606(4)	Reflux Cooler
E-1607(4)	First Stage Trim Cooler
K-1601(4)	First Stage Compressor
K-1602(4)	Second Stage Compressor
P-1601A/B(4)	Quench Bottom Recycle Pump
P-1602A/B(4)	Quench Top Reflux Pump
V-1601(4)	Gasifier WHB Steam Drum
V-1602(4)	First Stage K.O. Drum
V-1603(4)	Compressor Suction Drum

Unit 17 - Shift Conversion

E-1701(4)	First Feed-Effluent Heat Exchanger
E-1702(4)	Second Feed-Effluent Heat Exchanger
R-1701(4)	First Stage Shift Converter
R-1702(4)	Second Stage Shift Converter

Unit 18 - Acid Gas Removal

R-1801(4)	ZnO Bed
X-1801	Proprietary Benfield Unit

MAJOR EQUIPMENT LIST

Unit 19A - Methanation

<u>Item No.</u>	<u>Item Name</u>
E-1901(4)	Gas Cooler
E-1902(4)	Zinc Oxide Bed Preheater
E-1903(4)	Methanation Waste Heat Boiler
E-1904(4)	BFW Heater
H-1901	Start-up Heater
K-1901	Recycle Compressor
R-1901(4)	First Methanation Reactor
R-1902(4)	Second Methanation Reactor
R-1903(4)	Third Methanation Reactor
R-1904(4)	Fourth Methanation Reactor
V-1901(4)	KO Drum
V-1902(4)	Methanator WHB Steam Drum

Unit 19B - Final CO₂ Removal and Product Gas Compression

A-1901(2)	Regenerator Condenser
C-1901(2)	CO ₂ Absorber
C-1902(2)	Carbonate Regenerator
E-1905(2)	Lean Solution Cooler
E-1906(2)	Regenerator Reboiler
E-1907(2)	Product Gas Cooler
E-1908(2)	Compressor After Cooler
K-1902(2)	Product Gas Compressor
K-1903(2)	Third Stage Vapor Compressor
P-1901A/B(2)	Lean Solution Pump
P-1902A/B(2)	Regenerator Reflux Pump
V-1902(2)	Regenerator Reflux Drum
V-1904(2)	Product Gas KO Drum
V-1905(2)	After Cooler KO Drum

MAJOR EQUIPMENT LIST

Unit 20 - Effluent Treating

Item No.

A-2012(2)

C-2010(3)

E-2012(3)

E-2016(3)

E-2018(2)

P-2010(3)

TK-2001A/B/C

V-2010(2)

X-2001

CHEMICALS REMOVAL

Item Name

Stripped Condensate Air Cooler

Condensate Stripper

Condensate Stripper/Feed Bottom Exchanger

Condensate Stripper Reboiler

Stripped Condensate Water Cooler

Condensate Stripper Bottoms Pump

Ammonia Storage Tank

Condensate Oily Water Separator

Ammonia Recovery Unit

BIOLOGICAL POND

P-2020A/B(2)

P-2021A/B(2)

P-2022A/B(2)

P-2023A/B

P-2024A/B

P-2025A/B(2)

P-2026A/B(2)

P-2027A/B

P-2028A/B(2)

P-2029A/B(2)

P-2030A/B(2)

P-2031A/B

P-2032A/B

P-2033A/B(2)

P-2034A/B(2)

Primary Flotator Feed Pump

First Clarifier Recycle Pump

Primary Flotator Effluent Pump

Flotator Thickener Feed Pump

Digester Effluent Pump

Third Aerator Feed Pump

Filtered Water Pump

Thickener Float Pump

Second Clarifier Recirculation Pump

Second Aerator Feed Pump

Third Clarifier Recycle Pump

Filtrate Recirculation Pump

Sludge Thickener Float Pump

Primary Flotator Float Pump

Basin Float Pump

MAJOR EQUIPMENT LIST.

Unit 20 - Effluent Treating (Continued)

BIOLOGICAL POND (Con't.)

Item No.

Item Name

X-2020(2)	Primary Flash Mixer
X-2021(2)	Primary Flocculator
X-2022(2)	Primary Flotator
X-2023(2)	First Stage Aerator
X-2024(2)	First Stage Clarifier
X-2025	Aerobic Digester
X-2026	Sludge Flotator Thickener
X-2027	Sludge Cake Accumulator
X-2028A/B/C	Vacuum Filter
X-2029(2)	Second Stage Aerator
X-2030(2)	Second Stage Clarifier
X-2031(2)	Third Stage Aerator
X-2032(2)	Third Stage Clarifier
X-2033A/B/C	Gravity Sand Filter
X-2034(2)	Equalization Basin

MAJOR EQUIPMENT LIST

Unit 21 - Sulfur Recovery

<u>Item No.</u>	<u>Item Name</u>
E-2101A/B/C/D	Exchanger
S-2101	Stack
X-2101	Proprietary Sulfur Recovery Units

Unit 22 - Solids Disposal

CO-2203	Belt Conveyor
F-2201A/B	Sand Filters
F-2203A/B	Rotary Vacuum Filter
P-2201A/B	Sand Filter Feed Pump
P-2202A/B	Vacuum Filter Feed Pump
P-2203A/B	Sand Filter Backwash Pump
TH-2201	Filtered Water Surge Tank
TK-2201	Filtered Water Surge Tank
X-2202A/B	Pug Mill
X-2204A/B/C	Reject Ash Cyclone System
X-2250	Slurry Cooler

Unit 23 - Product Drying

C-2301(2)	Glycol Absorber
C-2302(2)	Regenerator Column
E-2301(2)	Regenerator
F-2301(2)	Filter
P-2301A/B(2)	Glycol Pump
V-2301(2)	Storage Tank
V-2302(2)	Flash Drum

MAJOR EQUIPMENT LIST

Unit 30 - Steam, Condensate and Power Generation

<u>Item No.</u>	<u>Item Name</u>
A-3001A/B/C	Turbine Generator Steam Condensers
B-3001	Start-up Boiler
D-3001	Packaged Air Dryer
E-3001	Blowdown Exchanger
K-3001A/B	Plant Air Compressors
P-3001A/B/C	HP BFW Pumps
P-3002A/B/C	MP BFW Pumps
P-3003A/B/C	LP BFW Pumps
P-3004A/B/C	Condensate Return Pumps
P-3005	O ₂ Scavenger Feed Pump
P-3006	Condensate Corrosion Inhibitor Feed Pump
P-3007	Dispersant Feed Pump
P-3008	Spare Chemical Feed Pump
TG-3001A/B/C	Turbine Generators
TK-3001	O ₂ Scavenger Day Tank
TK-3002	Condensate Corrosion Inhibitor Bulk Tank
TK-3003	Condensate Corrosion Inhibitor Day Tank
TK-3004	Dispersant Bulk Tank
TK-3005	Dispersant Day Tank
V-3001	Deaerator and Storage
V-3002	Condensate Return Flash Drum
V-3003	Blowdown Flash Drum
V-3004	HP Condensate Flash Drum
V-3005A/B	Plant Air Receivers

MAJOR EQUIPMENT LIST

Unit 31 - Plant Water System

<u>Item No.</u>	<u>Item Name</u>
F-3101A/DD	Anthracite Pressure Filters
P-3101A/B/C	Filter Feed Pumps
P-3102A/B/C/D	Fire Water Pumps
P-3103	Fire Water Jockey Pump
P-3104A/B/C	Cooling Water Pumps
P-3105A/B/C	Demineralized Water Pumps
P-3106A/B	Neutralizing Sump Pumps
P-3107A/B	Sulfuric Acid Pumps
P-3108	Inhibitor Pump
P-3109	Dispersant Pumps
P-3110	Spare Chemical Pump
P-3111A/B	Sulfuric Acid Transfer Pumps
P-3112A/B	Caustic Transfer Pumps
TK-3101	Demineralized Water Tank
TK-3102	Sulfuric Acid Day Tank
TK-3103	Corrosion Inhibitor Bulk Tank
TK-3104	Corrosion Inhibitor Day Tank
TK-3105	Dispersant Bulk Tank
TK-3106	Dispersant Day Tank
TK-3107	Sulfuric Acid Storage Tank
TK-3108	Caustic Storage Tank
X-3101	Water Reservoir
X-3102	Cooling Tower
X-3103	Demineralizer
X-3104	Waste Neutralizing Sump
X-3105	Waste Water Evaporator System
X-3106	ANDCO Chromate Removal Package
X-3107	Chromate Clarifier & Thickener Package
X-3108	Chlorinator Package

BATTELLE AGGLOMERATING ASH BURNER PROCESS

SPECIFICATIONS OF MAJOR ITEMS

Unit 12 - Coal Preparation

D-1201

No. Req'd :
Type :
Mat'l of Const. :
Drive :
Size :

Dryer Package

Three
Rotary Direct Fired
CS
Motor - 100 HP
13'-6" Dia. x 70' long

GR-1201

No. Req'd :
Type :
Mat'l of Const. :
Drive :
Size :

Coal Mill

Four
2 Row Stedman Cage Mill
CS, SS Elements
Motors-200 HP, 300 HP
300 TPH Capacity, 5' Diameter

V-1201

No. Req'd :
Type :
Mat'l of Const. :
Size :

Storage Bin

One
4 Bottom Spikes, 70 degree sides
CS
30' x 125' x 70' total height

V-1202

No. Req'd :
Type :
Mat'l of Const. :
Size :

Screen Feed Bin

One
26 Bottom Spikes, 70 degree sides
CS
18' x 104' x 10' vertical height

V-1203

No. Req'd :
Type :
Mat'l of Const. :
Size :

Combustor Distribution Bin

One
4 Bottom Spikes, 70 degree sides
CS
12' x 50' x 10' vertical height

V-1223

No. Req'd :
Type :
Mat'l of Const. :
Size :

Gasifier Distribution Bin

One
4 Bottom Spikes, 70 degree sides
CS
12' x 90' x 16' vertical height

SPECIFICATIONS OF MAJOR ITEMS

Unit 14 - Coal Feeding

V-1404

No. Req'd :
Type :
Mat'l of Const. :
Size :

Combustor Lockhopper Bin

Four
Vertical, cone bottom
CS
34' ID x 40' T-T top, 60° bottom

V-1405

No. Req'd :
Type :
Mat'l of Const. :
Size :

Combustor Lockhoppers

Eight
Vertical, cone bottom
CS
6' ID x 9' T-T top, 60° bottom

V-1424

No. Req'd :
Type :
Mat'l of Const. :
size :

Gasifier Lockhopper Bin

Four
Vertical, cone bottom
CS
32' ID x 35' T-T, 60° bottom

V-1425

No. Req'd :
Type :
Mat'l of Const. :
Size :

Gasifier Lockhoppers

Eight
Vertical, cone bottom
CS
12' 6" ID x 19' T-T top, 60° bottom

SPECIFICATIONS OF MAJOR ITEMS

Unit 15 - Gasification and Combustion

E-1502

No. Req'd :
Duty :
Mat'l of Const. :
Area :

No. 1 Combustor Waste Heat Boiler

Four
250 x 10⁶ Btu/hr.
Tubes - Alloy 800, Shell - Refractory
lined alloy
12,000 Sq. Ft.

E-1506

No. Requ'd :
Duty :
Mat'l of Const. :
Area :

No. 2 Combustor Waste Heat Boiler

Four
110 x 10⁶ Btu/hr.
Tubes - SS, Shell - SS clad
11,000 Sq. Ft.

E-1507

No. Req'd :
Type :
Duty :
Mat'l of Const. :
Area :

Cooler Condenser

Four
FTS
13 x 10⁶ Btu/hr.
Tubes - SS, Shell - CS
5,000 Sq. Ft.

E-1508

No. Req'd :
Type :
Duty :
Mat'l of Const. :
Area :

Combustor Flue Gas Cooler

Four
FTS
55 x 10⁶ Btu/hr.
Tubes - Alloy, Shell - CS
17,000 Sq. Ft.

SPECIFICATIONS OF MAJOR ITEMS

Unit 15 - Gasification and Combustion (Con't)

K-1501

Lift Gas Compressor

No. Req'd	:	Four
Type	:	Centrifugal
Capacity	:	900 ACFM @ suction
Pressure	:	Suction - 17 psia, discharge 120 psia
Drive	:	250 HP motor
Mat'l of Const.	:	316 SS

K-1502

Air Compressor

No. Req'd	:	Four
Type	:	Centrifugal
Capacity	:	337,000 ACFM @ suction
BHP	:	86,000
Pressure	:	Suction - 14 psia, discharge-120 psia
Drive	:	Flue Gas Expander
Mat'l of Const.	:	Manufacturer's standard

TE-1501

Flue Gas Expander

No. Req'd	:	Four
Capacity	:	173,000 ACFM @ suction
Pressure	:	Suction-95 psia, discharge 17 psia
BHP	:	95,000 generated
Mat'l of Const.	:	Manufacturer's standard

SPECIFICATIONS OF MAJOR ITEMS

Unit 15-Gasification and Combustion (Cont'd)

R-1501

Gasifier

No. Req'd	:	Four
Design Conditions	:	150 psia, 650°F for vessel
Mat'l of Const.	:	Vessel 18-8 Clad, 1/2 Mo. Steel overlay Lining 2ft. thickness total made up of firebrick, insulating brick, super block

SIZE

	<u>Inside Dia. Ft.</u> <u>Lining</u>	<u>Inside Dia. Ft.</u> <u>Vessel</u>	<u>Height, Ft.</u>
Top Section	50'-0"	54'-0"	24'-0"
Transition			10'-0"
Intermediate Section	37'-6"	42'-0"	20'-0"
Transition			8'-0"
Bottom Section	18'-0"	22'-0"	16'-0"
Below Support Plate			10'-0"
			<u>88'-0"</u>

R-1502

Combustor

No. Req'd	:	Four
Design Conditions	:	150 psig, 650° for vessel
Mat'l of Const.	:	Vessel 18-8 Clad; 1/2 Mo. Steel overlay Lining 2'-3' thickness total made up of firebrick, insulating brick, super block

	<u>Inside Dia. Ft.</u> <u>Lining</u>	<u>Inside Dia. Ft.</u> <u>Vessel</u>	<u>Height, Ft.</u>
Top Section	48'-6"	53'-0"	24'-0"
Transition			10'-0"
Bottom Section	26'-0"	31'-0"	20'-0"
Below Support Plate			10'-0"
			<u>64'-0"</u>

SPECIFICATIONS OF MAJOR ITEMS

Unit 16 - Raw Gas Quench

A-1601

No. Req'd :
Duty :
Mat'l of Const. :
Area :

First Stage Air Cooler

Four
 37×10^6 Btu/hr.
Tubes - Alloy Steel
10,000 Sq. Ft. bare tube

A-1603

No. Req'd :
Duty :
Mat'l of Const. :
Area :

Reflux Air Cooler

Four
 40×10^6 Btu/hr.
Tubes - Alloy Steel
28,000 Sq. Ft. bare tube

C-1601

No. Req'd :
Type :
Size :
Mat'l of Const. :

Quench Tower

Four
Vertical, trayed cone bottom
18' OD x 40' T-T
304 SS Clad

SPECIFICATIONS OF MAJOR ITEMS

Unit 16 - Raw Gas Quench (Cont'd)

E-1601

No. Req'd :
Duty :
Mat'l of Const. :
Area :

Gasifier Waste Heat Boiler

Four
 434×10^6 Btu/Hr.
Tubes-Alloy 800, Shell-Refractory
lined alloy
129,000 Sq. Ft.

E-1604

No. Req'd :
Duty :
Type :
Mat'l of Const. :
Area :

Slurry Cooler

Four
 8.6×10^6 Btu/hr.
FTS
CS
1,000 Sq. Ft.

E-1606

No. Req'd :
Type :
Duty :
Mat'l of Const. :
Area :

Reflux Cooler

Four
FTS
 96×10^6 Btu/hr.
CS
30,000 Sq. Ft.

E-1607

No. Req'd :
Type :
Duty :
Mat'l of Const. :
Area :

First Stage Trim Cooler

Four
FTS
 6×10^6 Btu/hr.
CS
10,000 Sq. Ft.

SPECIFICATIONS OF MAJOR ITEMS

Unit 16 - Raw Gas Quench (Cont'd)

K-1601

No. Req'd
Type
Capacity
Pressure
BHP
Drive
Mat'l of Const.

:
:
:
:
:
:
:

First Stage Compressor

Four
Centrifugal
34,000 ACFM @ Suction
Suction-80 psia, discharge- 250 psia
20,500
Motor
Manufacturer's standard

K-1602

No. Req'd
Type
Capacity
Pressure
BHP
Drive
Mat'l of Const.

:
:
:
:
:
:
:

Second Stage Compressor

Four
Centrifugal
11,000 ACFM @ suction
Suction - 245 psia, discharge-500 psia
11,700
Motor
Manufacturer's standard

SPECIFICATIONS OF MAJOR ITEMS

Unit 17 - Shift Conversion

E-1701

First Feed - Effluent Heat Exchanger

No. Req'd	:	Four
Type	:	FTS
Duty	:	16×10^6 Btu/hr
Mat'l of Const.	:	CS
Area	:	4,700 Sq. Ft.

E-1702

Second Feed Effluent Heat Exchangers

No. Req'd	:	Four
Type	:	FTS
Duty	:	11×10^6 Btu/hr
Mat'l of Const.	:	CS
Area	:	1,300 Sq. Ft.

R-1701

First Stage Shift Converter

No. Req'd	:	Four
Type	:	Vertical, with grate
Mat'l of Const.	:	CS, brick lined
Size	:	8'-6" ID lining, 10'-ID Shell, 18' T-T

R-1702

Second Stage Shift Converter

No. Req'd	:	Four
Type	:	Vertical, with grate
Mat'l of Const.	:	CS, brick lined
Size	:	7'-6" ID lining, 9' ID shell, 15' T-T

Unit 18 - Acid Gas Removal

R-1801

ZnO Beds

No. Req'd	:	Four
Type	:	Vertical, packed with catalyst
Mat'l of Const.	:	SS Clad
Size	:	15' ID x 30' T-T

SPECIFICATIONS OF MAJOR ITEMS

Unit 19A - Methanation

E-1901

Gas Cooler

No. Req'd	:	Four
Type	:	FTS
Duty	:	138×10^6 Btu/hr
Mat'l of Const.	:	1/2 Mo Steel
Area	:	9,300 Sq. Ft.

E-1902

Zinc Oxide Bed Preheater

No. Req'd	:	Four
Type	:	FTS
Duty	:	53×10^6 Btu/hr
Mat'l of Const.	:	1/2 Mo Steel
Area	:	5,000 Sq. Ft.

E-1903

Methanation Waste Heat Boiler

No. Req'd	:	Four
Duty	:	375×10^6 Btu/hr
Mat'l of Const.	:	CS
Area	:	69,000 Sq. Ft.

E-1904

BFW Heater

No. Req'd	:	Four
Type	:	FTS
Duty	:	243×10^6 Btu/hr
Mat'l of Const.	:	CS Shell, 304SS tubes
Area	:	50,000 Sq. Ft.

H-1901

Start-up Heater

No. Req'd	:	One
Type	:	Direct-Fired Heater
Mat'l of Const.	:	Refractory lined; C. S. Shell
		1-Cr-1/2 Mo Tubes
Duty	:	40×10^6 Btu/hr
Firing	:	Oil or Gas

SPECIFICATIONS OF MAJOR ITEMS

Unit 19A - Methanation (cont'd)

K-1901

No. Req'd :
Type :
Capacity :
Pressure :
BHP :
Drive :
Mat'l of Const. :

Recycle Compressor

One
Centrifugal
2,100,000 SCFM
Suction 430 psia, discharge 470 psia
20,400
Steam Turbine
To be specified by vendor

R-1901

No. Req'd :
Type :
Mat'l of Const. :
Size :

First Methanation Reactor

Four
Spherical with catalyst bed
RL (1/2 Mo Steel) with refractory lined
20' ID

R-1902

No. Req'd :
Type :
Mat'l of Const. :
Size :

Second Methanation Reactor

Four
Spherical with catalyst bed
RL(1/2 Mo Steel) with refractory lined
26' ID

R-1903

No. Req'd :
Type :
Mat'l of Const. :
Size :

Third Methanation Reactor

Four
Spherical with Catalyst bed
RL(1/2 Mo Steel) with refractory lined
30' ID

R-1904

No. Req'd :
Type :
Mat'l of Const. :
Size :

Fourth Methanation Reactor

Four
Spherical with catalyst bed
RL(1/2 Mo Steel) with refractory lined
30' ID

V-1901

No. Req'd :
Type :
Mat'l of Const. :
Size :

K. O. Drum

Four
Vertical
CS
15' OD x 15" T-T

SPECIFICATIONS OF MAJOR ITEMS

Unit 19B - Final CO₂ Removal and Product Gas Compression

A-1901

Regenerator Condenser

No. Req'd	:	Two
Type	:	Air Cooler
Duty	:	78×10^6 Btu/hr
Mat'l of Const.	:	CS
Area	:	27,500 Sq. Ft. bare tubes

C-1901

CO₂ Absorber

No. Req'd	:	Two
Type	:	Vertical, 3 beds
Mat'l of Const.	:	CS; packing CS and SS
Size	:	90' T-T, 9' Dia. top bed, 11'-9" Dia. bottom (2beds)

C-1902

Carbonate Regenerator

No. Req'd	:	Two
Type	:	Vertical, 3 beds
Mat'l of Const.	:	CS; packing CS and SS
Size	:	14' Dia. x 110' T-T

E-1905

Lean Solution Cooler

No. Req'd	:	Two
Type	:	FTS
Mat'l of Const.	:	CS
Duty	:	29×10^6 Btu/hr
Area	:	2,800 Sq. Ft.

E-1906

Regenerator Reboiler

No. Req'd	:	Two
Type	:	FTS
Mat'l of Const.	:	CS
Duty	:	73×10^6 Btu/hr
Area	:	10,500 Sq. Ft.

E-1907

Product Gas Cooler

No. Req'd	:	Two
Type	:	FTS
Duty	:	20×10^6 Btu/hr
Mat'l of Const.	:	CS
Area	:	23,000 Sq. Ft.

SPECIFICATIONS OF MAJOR ITEMS

Unit 19B - Final CO₂ Removal Product Gas Compression (cont'd)

E-1908

Compressor Aftercooler

No Req'd	:	Two
Type	:	FTS
Duty	:	46 x 10 ⁶ Btu/hr
Mat'l of Const.	:	CS
Area	:	24,000 Sq. Ft.

K-1902

Product Gas Compressor

No. Req'd	:	Two
Type	:	Centrifugal
Capacity	:	94,000 SCFM
Pressure	:	Suction-425 discharge-1020 psia
Drive	:	9000 HP Motor
Mat'l of Const.	:	CS

V-1902

Regenerator Reflux Drum

No. Req'd	:	Two
Type	:	Vertical
Mat'l of Const.	:	CS
Size	:	9' ID x 10' T-T

V-1904

Product Gas K.O. Drum

No. Req'd	:	Two
Type	:	Vertical
Mat'l of Const.	:	CS

V-1905

Aftercooler K. O. Drum

No. Req'd	:	Two
Type	:	Vertical
Mat'l of Const.	:	CS
Size	:	5' ID x 6' T-T

Unit 21 - Sulfur Recovery

E-2101

Exchanger

No. Req'd	:	Four
Type	:	FTS
Duty	:	80 x 10 ⁶ Btu/hr
Mat'l of Const.	:	316 SS
Area	:	28,000 Sq. Ft.

SPECIFICATIONS OF MAJOR ITEMS

Unit 30 - Steam, Condensate and Power Generation

A-3001

Turbine Generator Condenser

No. Req'd	:	Three
Type	:	Air Cooler
Duty	:	540×10^6 Btu/hr
Mat'l of Const.	:	CS
Area	:	200,000 Sq. Ft., bare tube

B-3001

Startup Boiler

No. Req'd	:	One
Type	:	Direct Fired
Mat'l of Const.	:	Manufacturer's standard
Duty	:	$1,000 \times 10^6$ Btu/hr
Firing	:	Oil

P-3001

HP BFW Pumps

No. Req'd	:	Three
Type	:	Horizontal Centrifugal
Mat'l of Const.	:	CS
Capacity	:	2,600 GPM @ 3700 Ft. TDH
Drive	:	3300 HP Motor

P-3002

MP BFW Pumps

No. Req'd	:	Three
Type	:	Horizontal Centrifugal
Mat'l of Const.	:	CS
Capacity	:	1,700 GPM @ 1400 FT. TDH
Drive	:	800 HP Motor

P-3003

LP BFW Pumps

No. Req'd	:	Three
Type	:	Horizontal Centrifugal
Mat'l of Const.	:	CS
Capacity	:	500 GPM, @ 460 Ft. TDH
Drive	:	75 HP Motor

SPECIFICATIONS OF MAJOR ITEMS

Unit 30 - Steam Condensate and Power Generation (Cont'd)

P-3004

Condensate Return Pumps

No. Req'd	:	One
Type	:	Centrifugal
Mat'l of Const.	:	CS
Capacity	:	1600 GPM @ 260 Ft. TDH
Drive	:	250 HP motor

TG-3001

Turbine Generator

No. Req'd	:	Three
Size	:	130 MW, 150 MVA
Drive	:	Steam Turbine

V-3001

Deaerator

No. Req'd	:	One
Type	:	Manufacturers standard
Mat'l of Const.	:	CS
Size	:	9' Dia. x 30' T-T

V-3002

Condensate Return Flash Drum

No. Req'd	:	One
Type	:	Vertical
Mat'l of Const.	:	CS
Size	:	12' Dia. x 35' T-T

V-3003

Blowdown Flash Drum

No. Req'd	:	One
Type	:	Vertical
Mat'l of Const.	:	CS
Size	:	8' Dia. x 16' T-T

V-3004

HP Condensate Flash Drum

No. Req'd	:	One
Type	:	Vertical
Mat'l of Const.	:	CS
Size	:	5' Dia. x 8' T-T

SPECIFICATIONS OF MAJOR ITEMS

Unit 31 - Plant Water System

F-3101

Anthracite Pressure Filters

No. Req'd	:	Thirty
Mat'l of Const.	:	Manufacturer's standard
Size	:	10' Dia.

P-3101

Filter Feed Pumps

No. Req'd	:	Three
Type	:	Vertical
Mat'l of Const.	:	Cast Iron
Drive	:	500 HP Motor
Capacity	:	3,000 GPM, @ 350 Ft. TDH

P-3102

Fire Water Pump

No. Req'd	:	Four
Type	:	Vertical
Mat'l of Const.	:	Cast Iron
Drive	:	2 Motors 700 HP, 2 Diesel Motors 650 HP
Capacity	:	5,000 GPM, @ 340 Ft. TDH

P-3104

Cooling Water Pump

No. Req'd	:	Three
Type	:	Vertical
Mat'l of Const.	:	CS
Capacity	:	23,000 GPM @ 200 Ft. TDH
Drive	:	1750 HP Motor

P-3105

Demineralized Water Pumps

No. Req'd	:	Three
Type	:	Centrifugal
Mat'l of Const.	:	316 SS
Capacity	:	3,000 GPM @ 300 Ft. TDH
Drive	:	400 HP Motor

SPECIFICATIONS OF MAJOR ITEMS

Unit 31 - Plant Water System (Cont'd)

TK-3101

No. Req'd :
Type :
Mat'l of Const. :
Size :

Demineralized Water Tank

One
A.P.I.
CS, Plasite 7155 HHB lining
160' Dia. x 40' High

TK-3107

No. Req'd :
Type :
Mat'l of Const. :
Size :

Sulfuric Acid Storage Tank

One
A.P.I.
CS
30,000 Gal

TK-3108

No. Req'd :
Type :
Mat'l of Const. :
Size :

Caustic Storage Tank

One
A.P.I.
CS
30,000 Gal.

X-3102

No. Req'd :
Type :
Duty :
Design :

Cooling Tower

One
Forced Draft
45000 GPM, 900×10^6 Btu/hr
Five cells @ 200 HP each -
Wet bulb temperature 75°F.

X-3103

No. Req'd :
Capacity :

Demineralizer System

One
6,000 GPM

VI. PROCESS DESIGN CRITIQUE

This section presents an analysis and review of the Battelle Agglomerating Ash Burner Process Design. For each of the major plant units the following information is provided:

- Summary of Battelle Design Basis
- Deficiencies in Battelle Design Basis
- SD Revisions to Design Basis
- SD Suggestions for Further Improvement or Study

In carrying out this analysis and review, no attempt was made to optimize the process by investigation of a broad range of variables or alternate schemes. Battelle's processing schemes were retained unless they were found to be deficient, in which case corrections or modifications were made with as little change as possible to the overall scheme.

UNIT 11 - COAL STORAGE AND RECLAIMING

Battelle Design Basis

The Battelle flowsheets indicate coal stockpiling and recovery by means of trucks and frontloaders, respectively. Crushing of the coal is also indicated by Battelle.

Revised SD Design Basis

SD considers the use of highly labor intensive trucks and frontloaders to be uneconomical for high capacity operation. For the SD design this method has been replaced by a boom-stacker/bucket wheel reclaimer system.

In accordance with the CFBraun "Guidelines" the eastern coal has been assumed to be received as 1 1/4" x 0" sized from the mine or previous preparation step and therefore, further crushing is not required. Removal of tramp iron has been allowed for.

UNIT 12 - COAL PREPARATION

Battelle Design Basis

The Battelle design calls for drying the coal from the stockpile followed by grinding by an air-swept roller mill. Battelle requires approximately 82.5% of the total coal to be ground to -8+100 mesh for feeding the gasifier and 17.5% to be ground to -100 mesh for feeding the Combustor.

Revised SD Design Basis

The subject of coal milling, which is the pivotal step of this section, was discussed with various vendors and a consensus emerged that an air swept roller mill which does internal size classification would not be a proper selection for reliably and efficiently producing the required coal size distribution at these capacities.

The Battelle requirement for sized coal cuts of -8 +100 mesh is a rather specialized coarse grading of prepared coal. Cage mills and impactor (hammer) mills have been used extensively for the reliable preparation of similar coal sizes for coke ovens and other chemical process users. These systems are generally designed with a recycle of oversized material to reduce the production of fines ("closed circuit" vs once-through "open circuit"). Air-swept roller mills, as called for by Battelle, on the other hand, are commonly used for the preparation of finer sized distributions such as for pulverized coal fines for combustion firing in the utility and chemical process industries and do not appear appropriate for the present application. Cage mills were therefore selected for this evaluation.

We believe that the use of a cage or hammer mill does not require predried coal. Over 8 mesh material is separated on a screen for recycle in a closed system. It is likely that any such mill would require between 5 and 30% recycle to reliably produce the required size distribution. However, since screens cannot be used to separate the -100 mesh from the -8 +100 mesh mill product fractions (wet or dry), air separation is indicated, and mechanically driven air separators have been provided for this service. Since air

separators require a dry powder regime, drying of the coal is necessary. Drying coal in fluid bed dryers would effect a separation of sorts, by elutriation of fines, but fluid bed dryers based on the CFBraun estimate for the CO₂ Acceptor process, seemed prohibitively expensive. Therefore, direct-fired rotary coal dryers were included ahead of the mills to dry the 1 1/4" x 0" coal conveyed from the storage piles.

Drying the coal before milling as indicated by Battelle, seems correct also from the point of view of the relative ease of removing water from the larger coal sizes which have less surface to retain moisture and from the point of being able to use higher gas velocities without excessive loss of fines. The rotary dryer uses -100 mesh coal as a source of fuel equivalent to approximately 1.5% of the product gas heating value.

At present, our estimate is based on modification of the CFBraun CO₂ Acceptor process flowsheets, implying mechanical conveying of -100 mesh coal from the air separator to the combustor feed system.

SD Suggestions for Further Improvement or Study

The entire coal preparation system is clearly open to further study and optimization.

Points for further study would include:

1. Clarification of the maximum and minimum moisture limits for coal from the mine and from the pile.
2. Drying rate data to be sought for the specific coal to be used.
3. Cage mill and impactor mill tests on eastern coal to produce the required size distribution.
4. The cost and practicality of fluidized transport of -100 mesh coal versus mechanical conveying should be checked with the vendors.
5. Battelle should establish as early as possible whether the total product from the mills can be split and fed to both the combustor and gasifier trains, without

separation as discussed in their proposals. This would simplify the system by greatly reducing the type and number of coal handling items and perhaps even eliminate the need for coal drying.

UNIT 14 - COAL FEEDING

Battelle Design Basis

Battelle's intention seems to be to simultaneously convey and pressurize the coal for the gasifier and combustor by means of solids fluidizing pumps, and to store the coal in pressurized bins.

Revised SD Design Basis

The Battelle method of feeding the gasifiers and combustors was not considered to be attractive or safe and we have substituted a lockhopper system adapted from the CFBraun CO₂ Acceptor flowsheets. There is the possibility of solids hangup or flow problems in feeding the -100 mesh material from the lockhopper to the combustor; however, we believe that a proper mechanical design can minimize and breakup such blockages.

UNIT 14 - COAL FEEDING (WITH PRE-TREATMENT ALTERNATIVE)

Battelle Design Basis

Battelle has not included pre-treatment in their design for eastern caking coals. They believe that the high circulation of ash in the gasifier will serve to dilute the coal particles and separate them sufficiently to prevent gross caking. They therefore, expect that pre-treatment will not be required, and have accordingly not shown any pre-treating facilities in the flowsheet presented to Scientific Design Company. However, they have included a pre-treatment system in their PDU in Columbus, Ohio, should pre-treating prove necessary, in practice.

Deficiencies in Battelle Design Basis

It has not been firmly established whether or not coal fed to the Battelle gasifier needs to be pre-treated by partial oxidation.

Until the requirement for pre-treatment is definitely ruled out, SD feels that the effect of its inclusion on the process economics should be evaluated.

SD Revisions to Design Basis

If pre-treatment should prove to be required, we have provided an alternative design including a fluidized bed pre-oxidizer vessel with a fresh air blower combined with recycle of the off-gases to provide a 5-10 volume percent oxygen stream to the pre-oxidizers. Pre-oxidizer operating conditions are 750°F and 2 hours residence time. Off-gases from the pre-oxidizer are quenched in a venturi scrubber, with a purge of hot gases taken upstream of the scrubber and sent to an incinerator to completely oxidize all organics to CO₂ and water prior to entering the SO₂ removal unit. Tars and oils are condensed in the venturi scrubber and separated in a decanter. A portion of the oil is used as auxillary fuel in the incinerator and the remainder exported for sale.

The gases leaving the incinerator at about 2000°F are cooled in a waste heat boiler to 450°F and then discharged to the SO₂ recovery unit. The steam raised in the waste heat boiler is expanded in a turbine power generating unit which supplies power to the fresh air and recycle compressors and the venturi scrubber recirculating pumps, with the excess power being exported for sale. The hot pre-treated coal at 750° is fed to the lockhoppers and then to the gasifiers, with one pre-treatment unit per train. The equipment list for this unit is shown in Section V.

SD Suggestions for Further Improvement or Study

If pre-treatment is determined to be necessary, the pre-treatment procedure required for the Battelle process should be studied in greater depth and additional process design work should be done to more closely integrate this unit with the rest of the plant. Such integration could lead to improved overall energy utilization and somewhat lower costs

then cited in Section VII.

UNIT 15 - GASIFICATION & COMBUSTION

<u>Battelle Design Basis</u>	<u>Gasifier</u>	<u>Combustor</u>
Pressure, psia	100	100
Exit Gas Temperature, °F	1,700	2,100
Calculated Gas Velocity-Upper Section, Ft/sec	1.25	2.0
Calculated Gas Velocity-Lower Section, Ft/sec	--	7.0
Steam/Carbon Weight Ratio	2.0	--

Deficiencies in Battelle Design Basis

SD is unable to comment on the expected technical performance or operability of the gasifier as a commercial unit since performance data is not available. However, the following comments are offered concerning the design:

A. Gasifier/Combustor Operating Pressure

The gasifier operating pressure of 100 psia is lower than most currently proposed gasification processes. This leads to the requirement for a larger number of gasifiers (and combustors) than for alternative processes, in order to maintain the desired linear velocity in the vessels. In addition, low pressure is unfavorable for methane production in the gasifier, leading to increased methanation costs downstream. Finally, compression costs to reach the desired product gas pressure of 1000 psig are greater when starting at a lower gasification pressure.

B. High Steam/Carbon Ratio in Gasifier

The steam/carbon ratio of 2.0 for the Battelle process is appreciably higher than for the Conoco CO₂ Acceptor Process ratio of 1.4. This results in:

1. Greater steam requirement for Battelle
2. Larger gas flow in the Battelle reactors leading to a requirement for larger reactors for the same final gas production. A comparison of the gasifier effluent for Battelle process and the CO₂ Acceptor process illustrates this significant difference:

<u>Component</u>	<u>CO₂ Acceptor</u>	<u>Battelle</u>
CO	12,204	25,908
H ₂	50,460	59,338
CH ₄	11,458	7,247
CO ₂	4,328	13,020
N ₂	196	478
H ₂ S	26	751
H ₂ O	19,838	43,216
NH ₃	392	---
TOTAL	98,902	149,958
P, psia	161	100
T, °F	1,545	1,700
ACFM	219,940	560,090

3. Greater condensing load downstream, leading to larger quench and cooler-condenser units and greater energy requirements for air or water cooling circulation.
4. Larger size of liquid effluent treating units to handle the contaminated condensate including additional sour water stripping, etc.

As a result of the low gasifier pressure and the high steam/carbon ratio four gasifiers are required, each larger than 50 ft. in diameter; with an external alloy metal shell of 54 ft. I.D. The ability of vessel manufacturers to build a vessel of this size and material was discussed with Wyatt Engineering of Houston who indicated that they had successfully built vessels of this size for catalytic crackers and distillation columns.

C. Gasifier & Combustion Operation Pressure

There is a need to demonstrate operability and process conditions at 100 psia since laboratory data were obtained at atmospheric pressure.

D. Circulation of Ash Agglomerates

The process requires transfer of a large quantity of hot ash agglomerate between gasifier and combustor. The ash must be non-sticky in order for such a transfer to be practical and feasible. The ash may possibly solidify if it is allowed to be stagnant during the upsets, startups and shutdowns. It will also be difficult to measure and control the flow.

E. Effect of Particulates on Gas Turbine

It is necessary to minimize the amount of solid particulates present in the hot flue gas that enters the gas turbines. Data must be obtained to demonstrate whether the Agglomerating Ash Process will result in a gas low enough in particulates for direct use in a gas turbine. Battelle has not shown any dust collecting device beyond the combustor cyclone. To assure protection of the gas turbine our design includes a dust collector incorporating small multicyclones.

F. Scrubber

The Battelle flowsheet shows a conventional packed bed water scrubber used to remove dust carryover from the gasifier. A large amount of cooling water is used with once through circulation. This is not practical since the packing would be subject to fouling. In addition an excessive burden would be placed on the effluent treating section because of the very large quantity of water to be handled. A quench tower is therefore substituted which employs an open grid-work in the bottom section to scrub the gas and an integral cooler-condenser section in the upper section to condense out water and discharge a sour water effluent. The bottom slurry is assumed to have a solids content of 5 weight percent.

G. Reject Ash

The Battelle flowsheet did not indicate any design for the handling of hot reject ash from the gasifier. A system was provided to cool the ash in a fluidized ash cooler prior to discharge.

UNIT 18 - ACID GAS REMOVAL

Battelle Design Basis

Feedgas:

Temperature = 257°F
Pressure = 490 psia

Treated Gas:

CO₂ Removal = 98 percent
H₂S - shown as nil in material balance

The regenerator pressure was specified as 480 psia with the regenerated gas delivered in two streams, one stream containing 15% H₂S in CO₂ and the other "pure" CO₂ for use in the lockhoppers.

Deficiencies in Battelle Design Basis

- A. The H₂S concentration in the purified gas should be specified and should be very low because the cost of final H₂S removal using ZnO beds before methanation is extremely high.
- B. The regenerator pressure of 480 psia specified by Battelle is contrary to conventional design and would increase the difficulty and cost of regeneration. For our design a pressure slightly above atmospheric has been used.
- C. To provide the regenerated acid gas in two streams as specified by Battelle would require a "sloppy" CO₂-H₂S absorber, where about one third of the CO₂ and the bulk of H₂S are removed, to be followed by a high-purity absorber in which the H₂S is purified to 1 ppm in the outlet gas.

Revised SD Design Basis

SD's design assumes use of the proprietary Benfield hot carbonate system. The gas conditions at the inlet to the Acid Gas Removal Unit and the percent CO₂ recovery are the same as shown by Battelle. For the H₂S we have assumed an outlet concentration of less than 1 ppm (by volume). The regenerated acid gas, consisting of only one stream at 17 psig containing about 5.5 percent H₂S (dry basis), and cooled to 140° for delivery

to the Sulfur Recovery Unit, has been assumed in the SD design.

SD Suggestions for Further Improvements

Due to the large quantities involved, acid gas removal for coal gasification processes is very expensive. The Battelle scheme requires acid gas removal before shift conversion and final CO₂ removal after methanation.

It is believed that a less expensive scheme will be to remove the acid gases only once at a location between the shift conversion and the methanation, as is done in alternative synthetic pipeline gas processes.

UNIT 17 - SHIFT CONVERSION

Battelle Design Basis

Shift Converter inlet gas temperature = 550°F
Acid washed gas - 23.7 percent to shift converters
 - 76.3 percent by-passed to methanation
Shift Conversion - 80.0 percent of the CO entering the shift converter is converted to H₂.

The Battelle process includes a two stage shift conversion scheme with a feed - effluent heat exchanger between stages. An external heat loss of 5 percent of the total enthalpy of the shift feed gas (21.6 MM Btu/hr) was assumed.

Deficiencies in Battelle Design Basis

A. Based on the Battelle design the shift catalyst has to be active at 550°F with a feed gas containing 1 ppm H₂S, and should have an operating temperature range of 500 to 750°F. Discussions with Girdler Corp. indicate that they have no shift catalyst which can meet these conditions. The available Girdler shift catalysts include:

G-66 Copper Zinc LTS Catalyst -

This catalyst cannot be used because its operating temperature range is limited to 400-500°F.

G-93 Cobalt-Moly Catalyst -

This catalyst has an operating temperature range of 500-900°C and is suitable as far as temperature is concerned. But, according to Girdler, it needs a minimum of 50 ppm of H₂S in the feed gas to keep the catalyst in the sulfided state. The cost of final removal of 50 ppm H₂S by ZnO before methanation will be prohibitively high, due to the high usage and non-regenerability of the ZnO.

G-3 Chrome-Iron Catalyst -

To use this catalyst the temperature of the feed gas to shift conversion must be raised from 550°F to about 650°F. There is no requirement for minimum H₂S content in the feed gas and the temperature range will be satisfactory.

- B. The heat loss of 21.6 MM Btu/hr for shift conversion assumed by Battelle is too high. A more reasonable estimate is 5% of the heat of reaction, which amounts to about 4.1 MM Btu/hr.

Revised SD Design Basis

The feed gas temperature has been increased from 550°F to 650°F by the addition of a feed - effluent exchanger after the second stage converter.

SD Suggestions for Further Improvements

If acid gas removal is carried out between shift conversion and methanation, the cost of shift conversion will be increased due to the higher CO₂ content in the converter feed gas and the need for using the more expensive Cobalt-Moly catalyst on account of the high H₂S content of 0.7 percent. It is believed, however, that the increase in the cost of shift conversion will be more than off-set by the decrease in the cost of CO₂ removal which will be accomplished in one step rather than two steps as in the Battelle scheme.

UNIT 19A - METHANATION

Battelle Design Basis

The Battelle scheme presented is a multi-stage process which uses sufficient cold product gas recycle to cool the effluent gas from each stage and assure a limited temperature rise in each bed under adiabatic conditions. However, there are some differences from other recycle processes one of them being that in the Battelle scheme, the recycle gas used is partially dried by cooling, as compared with hot gas recycle schemes.

Battelle used the following assumptions in their design:

- a) Carbon monoxide conversion of 99.9%
- b) Hydrogen conversion of 99.0%
- c) Recycle ratio (moles of recycle gas per mole of fresh feed) of 3:1
- d) Recycle gas temperature of 239°F.
- e) Feed gas temperature (mix temperature) of 450°F.
- f) Gas by-passing shift at 230°F.
- g) Shift effluent at 825°F.
- h) Heat loss: 1.4% of heat of reaction.

Battelle indicates a split of the fresh feed stream and recycle stream between beds but does not show the temperature or gas quantities to each bed. The calculations are on an overall basis rather than stage to stage.

Deficiencies in Battelle Design Basis

Several operating conditions are required in this section which are not met by the Battelle design:

- A. The fresh feed streams containing about 1 ppm H_2S should be heated to a temperature of 550°F in order to maximize the H_2S take-up by the ZnO beds. In the Battelle design the gas temperature is only 450°F, based on mixing the shifted stream at 825°F and the unshifted stream at 230°F.

- B. The required initiation temperature for the methanation reaction is approximately 500°F. Battelle's large recycle stream at 239°F mixed with fresh feed stream at 450°F leads to a mixture temperature of 290°F, which is much too low.
- C. Battelle's heat balance was found to be in error due to an incorrect value of the heat of methanation reaction being used.

Revised SD Design Basis

A. Reactors

Our original approach was simply to scale up the number of methanators from the Lurgi scheme shown in the CFBraun study based on overall gas flow. However, this procedure was felt to be too inaccurate, and therefore heat and material balances for each stage were made using a 50°F approach to equilibrium. Based on the results of this analysis inlet and exit temperatures for each bed were calculated as well as gas flows between beds. Using these gas flows resulted in larger beds than the original procedure.

Spherical vessels containing a shallow catalyst bed were selected for each stage. It is believed that this vessel configuration would minimize equipment costs. The spherical vessel is the best approach since for pressure drop reasons, the vessel size (diameter) is controlled primarily by the gas flow and not by the catalyst volume. A total of four trains with four reactor stages in series for each, were specified.

B. Gas-Gas Exchangers

The recycle stream entering the first reactor will be heated to the reaction temperature of 550°F by exchanging heat in E-1901 with a portion of the gas effluent from the Third Methanation Reactor. In addition fresh feed to the ZnO beds will be brought to reaction temperature by heat exchange with the Third Methanation Reactor effluent in E-1902.

C. Waste Heat Boiler

SD recalculated the waste heat boiler duty and resultant steam generation.

Our revised version calls for generating steam at one higher pressure level (500 psia) rather than two levels as Battelle shows in their calculations and flowsheets (this could only be accomplished by having two different waste heat boiler systems).

SD Suggestions for Further Improvement or Optimization

- A. Study the Battelle system in more depth, emphasizing the thermodynamic and kinetic aspects.
- B. Study the use of alternate methanation schemes which may be superior i.e., hot gas recycle, tubewall, fluid bed or liquid phase methanation.

UNIT 19B - FINAL CO₂ REMOVAL AND PRODUCT GAS COMPRESSION

Battelle Design Basis

Gas feed to final CO₂ Removal

Temperature	= 230°F
Pressure	= 440 psia
Regenerator Pressure	= 440 psia
CO ₂ Removal	= 99 percent
Product Gas Compression	= 435 psia to 1020 psia

Deficiencies in Battelle Design Basis

The regenerator pressure of 440 psia is much too high.

Revised SD Design Basis

All per Battelle design basis except that the regenerator will be operated at slightly higher than atmospheric pressure.

SD Suggestions for Further Improvements

- A. The CO₂ from the final CO₂ regenerator will be H₂S free and can be used as conveying gas in the lockhoppers.
- B. If the acid gas removal is done between the shift conversion and the methanation, there will be no need for the final CO₂ removal.

UNIT 20 - EFFLUENT TREATMENT

After the quench system in Unit 16 was redesigned the liquid flow was nearly the same as for the Synthane process as shown by CFBraun. Thus the entire Unit 20 of that process, including the biological pond, has been adopted.

UNIT 21 - SULFUR REMOVAL

Battelle Design Basis

The Battelle flowsheets show the H_2S stream from the Acid Wash Unit going to a Claus Unit and the SO_2 containing combustor flue gas stream going to an SO_2 removal unit of unspecified type.

Revised SD Design Basis

For H_2S streams, conventional well proven commercial technology exists to bring the gas concentration down to required levels and convert the H_2S to liquid sulfur. These include:

- A. Claus process followed by Stretford or Scot tail gas clean up units.
- B. Stretford or Scot processes by themselves.

For SO_2 streams, the technology is much less well developed since most of the effort has gone into alkaline scrubbing processes which produce alkali sulfites and sulfates rather than sulfur as end product. The two proprietary SO_2 treating processes investigated in the study were:

- A. Citrate process licensed by Arthur G. McKee
- B. Stackpol 150 process of Institute Francaise du Petrole (IFP).

Vendors of each of the above processes for H_2S and SO_2 treating were contacted and preliminary budget estimates and utility consumptions were obtained from each.

The processes selected for this estimate were the Citrate process for SO_2 removal and the Stretford process for H_2S , since together they provide the best combination from the point of view of capital cost, operating cost and expected performance.

It should be noted that the fuel gas usage for these units are quite considerable with a fuel requirement of 520 MM Btu/hr or the equivalent of 5% of the fuel value of the pipeline gas product (10,400 MM Btu/hr).

SD Suggestions for Future Study

It should be pointed out that a possible alternate to these separate processes for H_2S and SO_2 removal is a combined single unit process of the Citrate or Stackpol 150 type in which both the SO_2 and H_2S are introduced into different sections of the same unit. This could present an advantage since in the above processes, H_2S must be internally generated to react with the SO_2 for the production of sulfur. However, both vendors indicated lack of experience with this scheme and the probability of little saving at this stage of commercial development because of the large volume of inerts included in the H_2S stream.

UNIT 22 - SOLIDS DISPOSAL

No indication of a solids disposal scheme was provided on the Battelle flowsheets. SD adapted a suitable scheme from the CFBraun design for the CO_2 Acceptor process.

UNIT 30 - STEAM, CONDENSTATE AND POWER GENERATION

Battelle Design Basis

Battelle supplied a brief line sketch showing the flow of boiler feed water and steam.

This balance is summarized below:

<u>WATER IN</u>	<u>(LB/HR)</u>	<u>WATER OUT</u>	<u>(LB/HR)</u>
Make-up Water	2,260,000	Steam (275°F, 45 psia)	260,000
Condensate from		Methanation Unit	
Turbine-Generator	2,270,000	Steam (400°F, 150 psia)	1,150,000
		Methanation WHB	
		Steam (400°F, 150 psia)	460,000
		No. 2 Combustion WHB	
		Steam (550°F, 485 psia)	390,000
		Methanation WHB	
		Steam (990°F, 1250 psia)	1,070,000
		No. 1 Combustion WHB	
		Steam (900°F, 1250 psia)	1,200,000
		Gasifier WHB	
TOTALS	<u>4,530,000</u>		<u>4,530,000</u>

Deficiencies in Battelle Design Basis

Battelle supplied a brief line sketch showing the flow of boiler feed water and steam.

No reference was made to the usual steam system accessories such as deaerator, boiler feed water pumps, flash drums, heat exchangers, etc.

The steam turbines used for power generation show no provision for extraction and reheat of steam, without which the moisture content in the final stages of the turbine would be excessive. The units were depicted to operate at 2" Hg condensing pressure which can only be achieved by using cooling water.

Revised SD Design Basis

The boiler feed water flows to the Gasification Waste Heat Boiler and No. 2 Combustion Waste Heat Boiler have not been changed. The boiler feed water flow to the No. 1 Combustion Waste Heat boiler has been modified to permit steam reheat by superheating of the steam extracted from an intermediate point in the Generator Steam Turbine (TG-3001 A/B/C). This avoids the problem of excessive moisture in the final stages. A bleed stream of 470,000 lb/hr. steam is extracted from the steam turbine to augment the 70 psia steam level. An Air Cooler (A-3001 A/B/C) is used to condense the steam at 2.7 psia and 137°F in order to reduce cooling water usage and conform to the CFBraun "Guidelines". The boiler feed water balance is as follows:

<u>INLET</u>	<u>(LB/HR)</u>	<u>OUTLET</u>	<u>(LB/HR)</u>
Condensate	2,325,000	BFW to No. 1 Combustor	2,180,000
Water Make-up	2,150,000	WHB and Gasifier WHB	
		BFW to Methanation WHB	1,420,000
		BFW to No. 2 Combustor WHB	450,000
		BFW to Sulfur Recovery	50,000
		Condensate Return Flash	230,000
		Deaerator Flash Steam	<u>145,000</u>
TOTALS	<u>4,475,000</u>		4,475,000

It can be seen that the water make-up and BFW requirements are approximately the same as Battelle. However, there is a different distribution pattern for the steam flow (see DWG 1887-1-110). It should be noted that Scientific Design has not optimized the steam grid to take into consideration optimum pressure levels, extraction of steam and use of steam for turbine drives.

The power obtained from the turbine generator was somewhat lower (260,000 KW) than that obtained by Battelle (280,000 KW) due to turbine operation at a higher condensing pressure and the extraction of steam.

SD Suggestions for Further Improvement

The steam pressure levels and preheating of boiler feed water and condensate streams should be optimized to minimize capital and operating costs. The use of steam turbine drives for large pumps and air cooler fans should be considered.

UNIT 31 - PLANT WATER SYSTEM

Battelle Design Basis

The cooling system depicted by Battelle was a simple line sketch with the following cooling water users indicated:

	<u>GPM</u>
Cooling of Gas before Compression.	14,900
Compressor Intercooler	3,200
H ₂ S Removal	<u>3,600</u>
TOTAL	21,700

The cooling water requirements were based on an inlet temperature of 80°F and an outlet temperature of 180°F. On that basis the cooling water tower load is approximately 1.1×10^9 Btu/hr. No air coolers were used. Condensate from methanation (1340 GPM) was returned to the cooling tower presumably to allow for evaporation losses.

Revised SD Design Basis

The cooling water system shown on drawing 1887-1-109 is based on an inlet temperature of 80°F and a maximum 120°F return, to minimize scaling of the heat exchangers.

The approximate users are as follows:

	<u>GPM</u>
Unit 16 Raw Gas Quench	8,700
Unit 19 Methanation	5,800
Unit 18 Acid Gas Removal	12,100
Unit 20 Effluent Treating	5,000
Unit 21 Sulfur Recovery	5,800
- Miscellaneous	<u>5,600</u>
TOTAL	43,000

The heat load to be dissipated by the cooling tower is approximately 1.0×10^9 Btu/hr.

The evaporative losses are estimated at 1200 GPM and are replaced by water from the biological treatment unit and the ammonia treatment unit. The flow of 43,000 GPM established by SD is greater than that used by Battelle due to the smaller temperature rise in the exchangers and the additional process requirements (methanation, effluent treating, sulfur recovery etc.). However, the use of air cooled exchangers reduces the cooling water flow requirements and the load on the cooling water tower.

VII. CAPITAL COST ESTIMATE

BASIS OF ESTIMATE

The capital cost estimates presented are 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, the CFBraun work for similar processing steps and materials of construction, and SD's judgement and experience are utilized. A complete major equipment list and specifications for the major items are included in Section V. The Total Plant Investment Summary, Table 7-1 provides a breakdown of costs for each process unit. The unit estimate summaries in Tables 7-A through N provide a breakdown of costs by type of equipment. Table 7-1 also includes a delta capital for the pre-treatment alternate.

Major Equipment

The major equipment costs were obtained from supplier rough price quotations, by adjusting prices from similar equipment in the Braun report or from 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

In general the CFBraun installed cost factors were adjusted for the complexity and equipment size of the specific Battelle plant section. These ratios were generally found to be consistent with SD's experience.

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 covers 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 in accordance with the "Gas Cost Guidelines." Contingency covers unknowns such as engineered equipment sizing and pricing, materials of construction, installation factors, and general facilities and contractor's home office costs.

Escalation

The capital cost estimates for all processes are based on January 1976 costs with no forward escalation included, in accordance with the CFBraun "Guidelines."

INVESTMENT COST COMPARISONS WITH ALTERNATIVE PROCESSES

The total plant capital cost for the Battelle process and the investment for each unit may be compared with other processes, evaluated in the CFBraun estimates for western coal, shown in the summary tabulation, Table 7-2. As can be observed from Table 7-2 in comparison with the Battelle estimate, Table 7-1, the Battelle process investment is significantly greater than all other process with the exception of two of the Synthane Cases. The Battelle capital cost is 50% greater than the lowest investment processes, Hygas (Steam-Oxygen) and Conoco CO₂ Acceptor.

Since the Battelle estimate was prepared for eastern coal and the others are based on western coal, an approximation of the difference in capital cost for the Battelle process on eastern vs. western basis was made. The major differences would be in the coal storage, preparation and feeding section, Unit Nos. 11, 12 and 14, and the Sulfur Recovery Unit No. 21. The estimated differences in installed cost are as follows:

Millions of Dollars			
<u>Unit</u>	<u>Eastern Coal</u>	<u>Western Coal</u>	<u>Difference</u>
11 Coal Storage	11	13	+ 2
12 Coal Prep.	32	60	+28
14 Coal Feed	<u>22</u>	<u>34</u>	<u>+12</u>
Sub-total	65	107	+42
21 Sulfur Recovery	165	95	<u>-70</u>
NET DIFFERENCE IN INSTALLED COST			-28

If contractors home office costs and fee at 10% and contingency at 15% of installed cost are added, the total difference is \$35,000,000 resulting in an overall plant cost of approximately \$1,225,000,000 on western coal basis or about 3% less than for eastern coal. This figure is still substantially higher than for the alternative gasification process in the CFBraun report.

Although a quantitative comparison of the various pipeline gas processes to determine the technical and economic factors leading to the higher capital cost for the Battelle Agglomerating Ash Burner Process has not been made, some of the major points are as follows:

1) Low Gasifier and Combustor Operating Pressure

As indicated in Section VI, Design Critique, the gasifier and combustor operating pressure of 100 psia is lower than most currently proposed processes. This leads to the requirement for a larger number of gasifiers and combustors than for alternate processes, and consequently to a significantly higher capital cost for these units.

2) Low Methane Make and Concentration in the Gasifier

The Battelle gasifier effluent methane concentration is quite low in comparison to the other processes in the CFBraun report. The methane concentration for the Battelle process on a wet basis is only 4.8 mol% vs. 5.7 to 11.6% for the other processes. On a dry basis the gasifier effluent methane content is 6.8% for Battelle vs. 14.4 to 19.3% for the alternative processes. This leads to higher cost for the methanation step in the Battelle process.

In terms of the total methane make only 26.8% of the product methane is produced in the Battelle gasifier compared to 43.7 to 58.7% for the other processes.

3) Large Volume of Combustion Gases Produced and SO₂ Content

The Battelle process combustion step produces a very large quantity of hot flue gases, as compared to the other high Btu processes. In addition, a major portion of the sulfur in the coal is not converted to easily handled H₂S in the gasifier but rather stays with the char and is converted to SO₂ in the Combustor.

SO₂ is much more expensive to recover from the gas stream and convert to sulfur than H₂S. It is for this reason that the cost of sulfur removal in the Battelle process is very high.

4) Use of Dual CO₂ Removal Units

In the Battelle process, CO₂ is removed both prior to the CO shift reaction and following methanation. Alternative processes remove CO₂ only once, that is, after the shift reaction and prior to methanation. The use of a single CO₂ removal step in the Battelle process has been evaluated in Section IX of this report.

5) Methanation Scheme

The Battelle methanation scheme utilizes a cold gas recycle stream for temperature control. This method of heat removal requires more recycle gas and therefore larger more expensive equipment than using an alternative such as hot gas recycle. This alternative is evaluated in Section IX of this report.

BATTELLE AGGLOMERATING ASH BURNER PROCESS

TOTAL PLANT INVESTMENT SUMMARY

TABLE 7 - 1

<u>UNIT</u>		<u>BASE CASE</u>	<u>"DELTA" FOR COAL PRE-TREATMENT</u>
11	Coal Storage & Reclaiming	\$ 11,000,000	+\$ 1,000,000
12	Coal Preparation	32,000,000	+ 1,000,000
14	Coal Feeding	22,000,000	+ 25,000,000
15	Gasification & Combustion	252,000,000	-
16	Raw Gas Quench	82,000,000	-
17	Shift Conversion	7,000,000	-
18	Acid Gas Removal	54,000,000	-
19	Methanation, Purification & Comp.	80,000,000	-
20	Liquid Effluent Treatment	42,000,000	+ 5,000,000
21	Sulfur Recovery	145,000,000	+ 30,000,000
22	Solids Disposal	6,500,000	+ 500,000
23	Product Gas Drying	800,000	=
30	Steam Cond. & Power Generation	127,000,000	+ 48,000,000
31	Plant Water System	30,000,000	+ 2,500,000
--	General Facilities	<u>98,700,000</u>	+ <u>13,000,000</u>
TOTAL INSTALLED COST		\$ 990,000,000	+\$126,000,000
CONTRACTORS COST'S & FEE		<u>110,000,000</u>	+ <u>14,000,000</u>
SUBTOTAL		\$1,100,000,000	+\$140,000,000
CONTINGENCY		<u>160,000,000</u>	<u>21,000,000</u>
<u>TOTAL PLANT INVESTMENT</u>		<u>\$1,260,000,000</u>	<u>+\$161,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 11 - COAL STORAGE AND RECLAIMING

TABLE 7 - 1A

Major Equipment

Pumps	\$ 20,000
Conveying Equipment	
Belt	1,150,000
Elevators	7,000
Feeders	6,000
Chutes	23,000
Bins and Hoppers	19,000
Mills (Sample Crusher)	7,000
Weight Scales	6,000
Other Large Equipment	
Belt Trailers	240,000
Traveling Boomstacker	735,000
Bucket Wheel Reclaimer	980,000
Transfer Car	100,000
Magnetic Separators	26,000
Samplers	10,000
Allowance for Miscellaneous	<u>231,000</u>
Total Major Equipment	\$ 3,560,000
Installed Cost	<u>\$ 11,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 12 - COAL PREPARATION

TABLE 7 - 1B

Major Equipment

Pumps	\$ 20,000
Dryer Package	2,000,000
Conveying Equipment	
Belts	1,630,000
Feeders	329,000
Gates and Chutes	506,000
Bins	3,765,000
Mills	510,000
Weigh Scales	53,000
Separation Equipment	
Dust Collectors	100,000
Screens	650,000
Air Separator Package	480,000
Samplers	30,000
Allowance for Miscellaneous	<u>527,000</u>
Total Major Equipment	\$ 10,600,000
Installed Cost	<u>\$ 32,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 14 - COAL FEEDING

TABLE 7 - 1C

Major Equipment

Pressure Vessels	\$ 2,690,000
Hoppers	4,200,000
Conveying	170,000
Allowance for Miscellaneous	<u>280,000</u>
Total Major Equipment	\$ 7,340,000
Installed Cost	<u>\$ 22,000,000</u>

to the Sulfur Recovery Unit, has been assumed in the SD design.

SD Suggestions for Further Improvements

Due to the large quantities involved, acid gas removal for coal gasification processes is very expensive. The Battelle scheme requires acid gas removal before shift conversion and final CO₂ removal after methanation.

It is believed that a less expensive scheme will be to remove the acid gases only once at a location between the shift conversion and the methanation, as is done in alternative synthetic pipeline gas processes.

UNIT 17 - SHIFT CONVERSION

Battelle Design Basis

Shift Converter inlet gas temperature = 550°F
Acid washed gas - 23.7 percent to shift converters
 - 76.3 percent by-passed to methanation
Shift Conversion - 80.0 percent of the CO entering the shift converter is converted to H₂.

The Battelle process includes a two stage shift conversion scheme with a feed - effluent heat exchanger between stages. An external heat loss of 5 percent of the total enthalpy of the shift feed gas (21.6 MM Btu/hr) was assumed.

Deficiencies in Battelle Design Basis

A. Based on the Battelle design the shift catalyst has to be active at 550°F with a feed gas containing 1 ppm H₂S, and should have an operating temperature range of 500 to 750°F. Discussions with Girdler Corp. indicate that they have no shift catalyst which can meet these conditions. The available Girdler shift catalysts include:

G-66 Copper Zinc LTS Catalyst -

This catalyst cannot be used because its operating temperature range is limited to 400-500°F.

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 16 - RAW GAS QUENCH

TABLE 7 - 1E

Major Equipment

Columns	\$	3,400,000
Pressure Vessels		1,465,000
Exchangers		10,400,000
Pumps		255,000
Compressor		8,200,000
Allowance for Miscellaneous		<u>1,222,000</u>
	\$	24,942,000
Installed Cost	\$	<u>82,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 17 - SHIFT CONVERSION

TABLE 7 - 1F

Major Equipment

Reactors	\$ 1,450,000
Exchangers	400,000
Allowance for Miscellaneous	<u>150,000</u>
Total Major Equipment	\$ 2,000,000
Installed Cost:	<u>\$ 7,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 18 - ACID GAS REMOVAL

TABLE 7 - 1G

Major Equipment

Columns	\$	5,500,000
Reactors		2,500,000
Pressure Vessels		505,000
Exchangers		
Shell and Tube		1,293,000
Air Coolers w/Jet Ejectors		1,275,000
Pumps		936,000
Compressors		1,645,000
Allowance for Miscellaneous		<u>546,000</u>
Total Major Equipment	\$	14,200,000
Installed Cost	\$	<u>54,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 19 - METHANATION, FINAL CO₂ REMOVAL & GAS COMPRESSION

TABLE 7 - 1H

Reactors	\$ 11,000,000
Pressure Vessels	2,287,000
Exchangers	7,505,000
Pumps	391,000
Compressors	4,014,000
Start-up Heater	350,000
Allowance for Miscellaneous	<u>1,353,000</u>
Total Major Equipment	\$ 26,900,000
Installed Cost	<u>\$ 80,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 20 - LIQUID EFFLUENT TREATING

TABLE 7 - II

Major Equipment

Columns	\$	390,000
Pressure Vessels		114,000
Ammonia Storage Spheres		1,370,000
Exchangers		822,000
Pumps		35,000
Ammonia Removal Package		1,290,000
Bio-Oxidation Package		
Equipment		3,305,000
Tanks		285,000
Pumps		146,000
Allowance for Miscellaneous		<u>393,000</u>
Total Major Equipment	\$	8,150,000
Installed Cost	\$	<u>42,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 21 - SULFUR RECOVERY

TABLE 7 - 1J

Major Equipment

Proprietary Packages (Installed)	\$ 130,000,000
Combustion Includes Heaters, Burners, Stack	1,000,000
Exchangers	1,600,000
Allowance for Miscellaneous	<u>200,000</u>
Total Major Equipment	\$ 132,800,000
Installed Cost	<u>\$ 145,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 22 - SOLIDS DISPOSAL

TABLE 7 - 1K

Major Equipment

Thickener	\$ 105,000
Exchangers	300,000
Pumps	20,000
Conveyors	450,000
Pug Mill	185,000
Separation Equipment	
Rotary Vacuum Filters	300,000
Cyclones	500,000
Allowance for Miscellaneous	<u>105,000</u>
Total Major Equipment	\$ 1,965,000
Installed Cost	<u>\$ 6,500,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 20 - PRODUCT GAS DRYING

TABLE 7 - 1L

Major Equipment

Glycol Dehydration Package	<u>\$ 500,000</u>
Installed Cost	<u>\$ 800,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS
UNIT 30 - STEAM CONDENSATE & POWER GENERATION

TABLE 7 - 1M

Major Equipment

Pressure Vessel	\$	88,000
Tanks		25,000
Exchangers		1,500,000
Deareator		400,000
Pumps		611,000
Compressors and Blowers		900,000
Plant Air Dryer		45,000
Boilers		13,000,000
Electrostatic Precipitators		1,200,000
Power Generator Sets (Installed)		32,500,000
Export Power Distribution (Installed)		15,000,000
Allowance for Miscellaneous		<u>1,231,000</u>
Total Major Equipment	\$	66,500,000
Installed Cost	\$	<u>127,000,000</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

UNIT 31 - PLANT WATER SYSTEM

TABLE 7 - 1N

Major Equipment

Tanks	\$ 915,000
Pumps	713,000
Cooling Tower, including chlorination	570,000
Cold Lime Clarification Package	110,000
Filters and Demineralizers	3,190,000
Waste Water Evaporation Package	11,200,000
Allowance for Miscellaneous	<u>802,000</u>
Total Major Equipment	\$ 17,500,000
Installed Cost	<u>\$ 30,000,000</u>

TABLE 7-2

ERDA-A.G.A.		TABLE III PLANT INVESTMENT SUMMARY, WESTERN SUB-BITUMINOUS COAL FACTORED ESTIMATES						Project 4568-NW
UNIT	COAL GASIFICATION PLANT							
	IGT HYGAS STEAM-OXYGEN	IGT HYGAS STEAM-IRON	CONOCO CO ₂ ACCEPTOR	BCR BI-GAS	ERDA-PERC SYNTHIANE EXPORT POWER		EXPORT CHAR SLURRY FEED	LURGI
					DRY FEED	SLURRY FEED		
11 Coal Storage & Reclaiming	\$ 13,000,000	\$ 13,000,000	\$ 13,000,000	\$ 13,000,000	\$ 14,000,000	\$ 14,000,000	\$ 14,000,000	\$ 14,000,000
12 Coal Preparation	17,000,000	25,000,000	59,000,000	34,000,000	63,000,000	22,000,000	22,000,000	26,000,000
Lurgi Proprietary Units								191,000,000
14 Coal Feed	26,000,000	28,000,000	27,000,000	17,000,000	81,000,000	16,000,000	16,000,000	Included
15 Gasification & Power Recovery	45,000,000	459,000,000	193,000,000	44,000,000	48,000,000	54,000,000	54,000,000	Included
16 Raw Gas Quench	19,000,000	80,000,000	64,000,000	50,000,000	41,000,000	38,000,000	38,000,000	Included
17 Shift Conversion	31,000,000			32,000,000	52,000,000	52,000,000	52,000,000	Included
18 Acid Gas Removal	108,000,000	75,000,000	36,000,000	114,000,000	127,000,000	116,000,000	116,000,000	117,000,000
19 Methanation	29,000,000	26,000,000	36,000,000	29,000,000	33,000,000	33,000,000	33,000,000	58,000,000
20 Sour Water Stripping, Ammonia Recovery, & Bio-Oxidation	58,000,000	68,000,000	18,000,000	37,000,000	42,000,000	42,000,000	42,000,000	19,000,000
21 Sulfur Recovery	60,000,000	76,000,000	52,000,000	59,000,000	77,000,000	77,000,000	61,000,000	54,000,000
22 Solids Disposal	8,900,000	7,600,000	8,500,000	8,900,000	9,200,000	9,200,000	9,200,000	3,300,000
23 Product Gas Drying	800,000	800,000	800,000	800,000	800,000	800,000	800,000	800,000
20 Steam & Utility Systems	136,000,000	12,000,000	100,000,000	184,000,000	350,000,000	364,000,000	255,000,000	178,000,000
21 Plant Water Systems	17,000,000	34,000,000	18,000,000	25,000,000	20,000,000	20,000,000	20,000,000	22,000,000
22 Oxygen Plant	45,000,000			81,000,000	97,000,000	80,000,000	80,000,000	65,000,000
General Facilities	68,000,000	100,000,000	69,000,000	81,000,000	117,000,000	104,000,000	90,000,000	83,000,000
Site Preparation, Electrical Distribution, Yard Piping, Buildings, Flare System, Tankage, Shipping & Receiving Facilities, Sulfur Storage, Potable & Sanitary Water, Holding Ponds, Plant Roads								
Installed Cost	\$681,700,000	\$1,004,400,000	\$694,300,000	\$ 809,700,000	\$1,172,000,000	\$1,042,000,000	\$ 903,000,000	\$ 831,100,000
Contractors Home Office Cost & Fee	75,300,000	111,600,000	76,700,000	90,300,000	130,000,000	116,000,000	100,000,000	92,900,000
Subtotal	\$757,000,000	\$1,116,000,000	\$771,000,000	\$ 900,000,000	\$1,302,000,000	\$1,158,000,000	\$1,003,000,000	\$ 924,000,000
Contingency	113,000,000	164,000,000	119,000,000	130,000,000	198,000,000	172,000,000	147,000,000	136,000,000
Total Plant Investment	\$870,000,000	\$1,280,000,000	\$890,000,000	\$1,030,000,000	\$1,500,000,000	\$1,330,000,000	\$1,150,000,000	\$1,060,000,000
		EXPORT POWER CASE						DUAL FEED CASE

VIII. 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" in order to produce a consistent set of data for comparison with alternative processes. The results are shown in Tables 8-1 through 8-5.

Tables 8-1A & 8-1B show the Net Annual Operating Costs, allowing for by-product credit, for the Base Case and Pre-treatment alternate respectively. For the raw materials and water, the quantities estimated were multiplied by the unit costs in the "Guidelines" for eastern coal, while catalyst and chemicals costs were estimated or obtained from vendors. An annual on-stream time of 90% (7884 operating manhours) was used as suggested in the "Guidelines."

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 Tables 8-1A&B, and the calculation of Annual Maintenance Cost is presented in Table 8-2. The breakdown of Total Capital Requirement is given in Table 8-3.

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 8-4, together with gas costs for other processes calculated by CFBraun using the same methods. The alternate case based on the pre-treatment of the coal to the gasifier is also shown in Table 8-4. It can be observed that the Battelle process gas cost on each basis is appreciably higher than all other

processes evaluated. For comparison, the Battelle average gas cost with utility financing of \$5.20/MM Btu is noted to be almost twice the cost of the lowest process evaluated, HYGAS (steam-oxygen), and 50% greater than the Conoco CO₂ Acceptor, even though the latter process is burdened by the high operating cost for the acceptor as pointed out by CFBraun. Using an average gas cost of \$3.50/MM Btu for the other processes, the Battelle gas cost is almost 50% greater.

It should be noted that the difference, while highly capital related, also reflects the unit cost differences between western and eastern coals. To partially compensate for this unit cost difference, the gas cost for the Battelle Process and other processes have been calculated "ex coal," that is excluding the cost of coal from each. These figures are shown in Table 8-5.

The results indicate that the gas costs, even on this adjusted basis, are distinctly higher for the Battelle process than for the other processes. Thus the Battelle process cost "ex coal" of \$3.49/MM Btu is 65% higher than the HYGAS steam-oxygen route and 30% greater than the CO₂ Acceptor. Taking an average gas cost for the other processes of approximately \$2.70/MM Btu "ex coal", the Battelle gas cost is about 30% higher.

The cold gas and thermal efficiencies for the Battelle Process have been calculated and are compared to the alternate processes in Table 8-6. As can be seen from this table the Battelle efficiencies are significantly lower than the most favorable processes, HYGAS (steam-oxygen) and CO₂ Acceptor.

BATTELLE AGGLOMERATING ASH BURNER PROCESS

ANNUAL OPERATING COSTS

(IN MILLIONS OF DOLLARS)

TABLE 8-1A BASE CASE

RAW MATERIALS		\$139.94
Coal (710 tons/hr @ \$25.00/ton)		
CATALYSTS AND CHEMICALS		8.33
PURCHASED WATER (3200 GPM @ \$.40/1000 Gal)		0.61
ASH DISPOSAL (124 ton/hr @ 1.00/ton)		0.98
LABOR		
Process Operating Labor (50 men/shift @ \$6.70/hr)	2.78	
Maintenance Labor (60% of Annual Maintenance Costs)	25.48	
Supervision (20% of Direct Labor)	<u>5.65</u>	
		33.91
ADMINISTRATION & GENERAL OVERHEAD (60% of Labor)		20.35
SUPPLIES		
Operating (30% of Process Operating Labor)	0.83	
Maintenance (40% of Annual Maintenance Costs)	<u>16.99</u>	
		17.82
LOCAL TAXES AND INSURANCE (2.7% of Total Plant Investment)		34.02
<u>TOTAL GROSS ANNUAL OPERATING COSTS</u>		<u>\$255.96</u>
BY-PRODUCT CREDITS		
Sulfur (29.1 tons/hr @ \$25/long ton)	(5.12)	
Ammonia (1.85 tons/hr @ \$170/ton)	(2.48)	
Surplus Electrical Power(100,000 kwh/hr @ 0.01 kwh)	<u>(7.88)</u>	
		<u>(15.48)</u>
<u>TOTAL NET ANNUAL OPERATING COSTS</u>		<u>\$240.48</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

ANNUAL OPERATING COSTS

(IN MILLIONS OF DOLLARS)

TABLE 8-1B PRE-TREATMENT ALTERNATE

RAW MATERIALS		\$ 155.71
Coal (790 tons/hr @ 25.00/ton)		
CATALYSTS AND CHEMICALS		8.63
PURCHASED WATER(3700 GPM @ \$.40/1000 Gal)		0.70
ASH DISPOSAL (139 ton/hr @ 1.00/ton)		1.10
LABOR		
Process Operating Labor (50 men/shift @ \$6.70/hr)	2.78	
Maintenance Labor (60% of Annual Maintenance Costs)	27.98	
Supervisoin (20% of Direct Labor)	<u>6.15</u>	
		36.91
ADMINISTRATION AND GENERAL OVERHEAD (60% of Labor)		22.15
SUPPLIES		
Operating (30% of Process Operating Labor)	0.83	
Maintenance (40% of Annual Maintenance Costs)	<u>18.65</u>	
		19.48
LOCAL TAXES & INSURANCE (2.7% of Total Plant Investment)		38.37
TOTAL GROSS ANNUAL OPERATING COSTS		<u>\$ 283.05</u>
BY-PRODUCT CREDITS		
Sulfur (32.0 short tons/hr. @ \$25/long ton)	(5.63)	
Ammonia (2.00 tons/hr. @ \$170/ton)	(2.68)	
Surplus Electrical Power (120,000 kwh/hr @ 0.01 kwh)	(10.25)	
Tars (500 gallons/hr @\$0.20/gal)	(9.00)	
		<u>(27.56)</u>
<u>TOTAL NET ANNUAL OPERATING COSTS</u>		<u>\$ 255.49</u>

BATTELLE AGGLOMERATING ASH BURNER PROCESS

ANNUAL MAINTENANCE COSTS

(IN MILLIONS OF DOLLARS)

TABLE 8-2

NO.	UNIT	UNIT COST	MAINTENANCE	
			FACTOR %	COST
11	Coal Storage & Reclaiming.	\$ 14.00	6	\$ 0.84
12	Coal Preparation	40.73	6	2.44
14	Coal Feeding	28.00	6	1.68
15	Gasification & Combustion	320.73	4.5	14.43
16	Raw Gas Quench	104.36	6	6.26
17	Shift Conversion	8.91	3	0.27
18	Acid Gas Removal	68.73	3	2.06
19	Methanation, Purif. & Comp.	101.82	3	3.05
20	Liquid Effluent Treatment	53.45	3	1.60
21	Sulfur Recovery	184.55	3	5.54
22	Solids Disposal	8.27	3	0.25
23	Product Gas Drying	1.02	3	0.03
30	Steam, Cond. & Power Gen.	161.64	1	1.62
31	Plant Water System	38.18	3	1.14
--	General Facilities	<u>125.61</u>	<u>1</u>	<u>1.26</u>
	TOTAL PLANT	\$ 1,260.00	3.37	\$ 42.47

NOTE: The annual Maintenance Cost shown above is for the Base Case. If Pre-treatment of the coal is required the Annual Maintenance cost will increase to \$46.63 Million.

BATTELLE AGGLOMERATING ASH BURNER PROCESS

TOTAL CAPITAL REQUIREMENT

(MILLIONS OF DOLLARS)

TABLE 8-3

	DELTA FOR COAL	
	<u>BASE CASE</u>	<u>PRE-TREATMENT</u>
TOTAL PLANT INVESTMENT	\$1,260.00	+ \$161.00
INITIAL CHARGE OF CATALYSTS & CHEMICALS	6.20	Nil
PAID-UP ROYALTIES	1.75	+ 0.00
ALLOWANCE FOR FUNDS USED DURING CONST. (Total Plant Investment x Average Spending Period in Years x 0.09)	198.45	+ 25.36
START-UP COSTS (20 Percent of Total Annual Gross Annual Operating Costs)	51.19	+ 5.42
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 cal- culated sales prices.	35.74	+ 4.24
<u>TOTAL CAPITAL REQUIREMENT</u>	<u>\$1,553.33</u>	<u>+ \$ 196.02</u>

NOTE: Gas Sales Price used in above is based on average cost with utility financing.

COMPARISON OF GAS COSTS
BATTELLE PROCESS VS. OTHER PROCESSES
GAS COSTS, \$/MM BTU

TABLE 8-4

	<u>Battelle Without Coal Pre- treatment</u>	<u>Battelle With Coal Pre- treatment</u>	<u>IGT HYGAS Steam Oxygen</u>	<u>IGT HYGAS Steam-Iron Export Power</u>	<u>Conoco CO₂ Acceptor</u>	<u>BCR BI-GAS</u>	<u>PERC Synthane Slurry, Feed Export Char</u>	<u>Lurgi Dual Feed</u>
First Year Gas Cost Utility Financing Method	6.44	7.06	3.55	5.35	4.25	4.50	4.81	4.34
Average Gas Cost by the Utility Financing Method	5.20	5.67	2.71	4.10	3.38	3.50	3.70	3.30
Constant Gas Cost by the Private Investor's Financing Method	6.94	7.63	3.91	5.87	4.61	4.93	5.28	4.78

COMPARISON OF GAS COSTS-EXCLUDING COAL COSTS

BATTELLE PROCESS VS. OTHER PROCESSES

GAS COST EXCLUDING COAL COST, \$/MM BTU

TABLE 8-5

	<u>Battelle With Coal Pre- treatment</u>	<u>Battelle With Coal Pre- treatment</u>	<u>IGT HYGAS Steam Oxygen</u>	<u>IGT HYGAS Steam-Iron Export Power</u>	<u>Conoco CO₂ Acceptor</u>	<u>BCR BI-GAS</u>	<u>PERC Synthane Slurry, Feed Export Char</u>	<u>Lurgi Dual Feed</u>
Average Gas Cost by the Utility Financing Method	3.49	3.77	2.09	3.41	2.72	2.82	2.86	2.40

COMPARISON OF FUEL EFFICIENCIES

BATTELLE vs. OTHER PROCESSES

TABLE 8-6

	<u>COLD GAS EFFICIENCY</u>	<u>PLANT THERMAL EFFICIENCY</u>
HYGAS, Steam-Oxygen	69.7	77.6
HYGAS, Steam-Iron	62.2	70.0
Conoco CO ₂ Acceptor	65.2	67.0
BCR BI-GAS	62.9	63.6
Synthane, Dry Feed, Export Power	47.9	55.0
Lurgi Dual Feed	48.5	58.9
Lurgi Single Feed	59.0	71.7
BATTELLE	59.2	62.6
BATTELLE with pre-treatment	53.1	60.9

IX. EVALUATION OF SUGGESTED PROCESS IMPROVEMENTS

~~In the course of SD's evaluation, various process modifications were suggested as possible~~
improvements over those in the original Battelle process scheme. Two of these modifications have been evaluated in this section. These are:

- 1) Use of a single acid wash removal unit located between the CO shift reaction and the methanation step, rather than dual acid-gas removal, both prior to CO shift and following methanation, as suggested by Battelle.
- 2) Use of a hot gas recycle methanation scheme with indirect cooling as opposed to the cold gas recycle scheme for temperature control as suggested by Battelle. The former scheme permits a lower recycle to fresh feed ratio and therefore results in smaller equipment and lower capital cost.

For each of the above modifications, rough order of magnitude estimates have been prepared to determine whether significant improvements in plant economics could be achieved.

The following capital cost reductions were obtained, including allowance for contractors cost and fee plus contingency:

	<u>Saving</u>
(1) Use of Single vs. Dual Acid Wash	\$ 4,000,000
(2) Use of Hot vs. Cold Gas Methanation	<u>18,000,000</u>
TOTAL	\$ 22,000,000

The total reduction of \$22,000,000 represents a saving of less than 2% on the plant capital cost of \$1,260,000,000 and thus has a very minimal effect on the overall economics of the process.

In addition to these suggested modifications, SD has commented on the significant effect on capital cost of (1) the low operating pressure in the gasifier and combustor, (2) the low methane make and concentration in the gasifier, (3) the large volume of combustor off-gases with a high SO_2 content. However, any modification in process conditions in these areas would involve revisions to the basic Battelle gasification process. Such modifications would require both operating data and demonstration of performance at the revised conditions. It was felt, therefore, that such changes would involve major extrapolations of Battelle's current design basis, the validity of which could not be accurately predicted.

X. CONCLUSIONS

Based on the results of this evaluation, the following conclusions have been reached:

- 1) The Battelle Agglomerating Ash Burner process appears to be economically unattractive compared to alternative substitute natural gas processes. This conclusion is based on the significantly higher capital investment and gas cost for this process, even when allowance is made for the differences in cost for eastern vs. western coal.
- 2) The Battelle process shows thermal efficiencies significantly lower than the most efficient processes, i.e. HYGAS, CO₂ Acceptor and BI-GAS.
- 3) Some of the reasons for the higher capital cost and gas costs for the Battelle process are:
 - a. Low Gasifier and Combustor Operating Pressure
 - b. Low Methane Make and Concentration in the Gasifier Effluent
 - c. Large Volume of Combustion Gases Produced and SO₂ Content
 - d. Use of Dual CO₂ Removal Units.
 - e. Methanation Scheme

Items a,b, and c are basically due to Battelle's gasification design conditions. Items d and e can be modified by substitution of a single CO₂ removal unit for the dual units and revising the methanation scheme, respectively. However these improvements brought about a reduction of less than 2% in the overall plant cost. It appears therefore, that

modifications in the downstream gas processing operations will have only a minor effect on the overall economics, which are largely governed by the gasifier and combustor design and operating conditions.