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COMMERCIAL PLANT
CONCEPTUAL DESIGN AND COST ESTIMATE

CO_2 ACCEPTOR PROCESS
GASIFICATION
PILOT PLANT

FINAL REPORT, VOLUME 10
BOOK 1 OF 3, NORTH DAKOTA LIGNITE GASIFICATION, ECONOMICS AND DESCRIPTION
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FOREWORD

Volume 10 is the tenth report submitted in partial fulfillment of the DOE Contract (EX 76-C-01-1734) for development of a process known as the Conoco Coal Development Company CO_2 Acceptor Process for the production of pipeline quality gas from lignite. Previous reports are enumerated in the bibliography and include a feasibility study (1); description of bench scale research (2, 3, 4, 5); an economic evaluation (6); a pilot plant construction report (7); and several pilot plant operation reports (8, 9).

The work was done by Stearns-Roger Engineering Co. under a subcontract with Conoco Coal Development Company (CCDC). Principal investigators include G. P. Curran, G. J. Leaman, Jr., V. H. Melquist, C. A. Schulz and I. L. Zuber of CCDC.

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BOOK 1 — NORTH DAKOTA LIGNITE GASIFICATION - GENERAL ECONOMICS AND DESCRIPTION — PAGE NO.

ABSTRACT

This report presents the results of work to design and estimate the capital cost of a Carbon Dioxide (CO_2) Acceptor Coal Gasification plant. This plant is designed to produce 250 billion Btu/day of synthetic natural gas. Plant processes were, in part, optimized and the design is tailored to a specific lignite and plant location in North Dakota. This design was modified for gasification of Texas lignite. Plant operating costs and gas costs were calculated for the North Dakota lignite case and the Texas lignite case. These costs are compared with published data for a Lurgi Coal Gasification plant.

INTRODUCTION

This report presents the design, investment and operating costs for a commercially sized lignite gasification plant. Design is based upon Conoco Coal Development Company's Carbon Dioxide (CO_2) Acceptor Process. Plant production is approximately 250 billion Btu per day of pipeline gas. The assumed plant site is near Beulah, North Dakota. In addition, design of this plant has been modified to represent a plant producing synthesis gas (CO and H_2) from Texas lignite. Investment and operating costs are presented for the Texas lignite plant.

The design is based upon work by Conoco Coal Development Company (CCDC) in a developmental program sponsored by the Energy Research and Development Administration (ERDA), its predecessor, Office of Coal Reseach (OCR), and the American Gas Association (AGA). The program was initiated by a feasibility study in 1965 which was followed by bench-scale work at CCDC's Research facilities in Library, Pennsylvania. A large (40 ton per day lignite) pilot plant was then engineered and constructed in Rapid City, South Dakota. After plant acceptance in October of 1971, operations have continued with completion of the program scheduled by the end of 1977.

The CO_2 Acceptor Process embodies gasification of lignite coal with steam in a fluidized bed at temperatures of 1500-1600°F and pressures of 150 psig to produce synthesis gas. The primary source of heat for the endothermic steam-gasification reaction is provided by combining CaO with CO_2, to form $CaCO_3$, an exothermic reaction in the same gasification vessel. The CaO is regenerated in a second fluidized bed vessel by combustion of char with air.

Since the CaO (limestone) becomes deactivated after a number of cycles, a portion of it is rejected and must be replaced with fresh material or reconstituted reject stone. The choice is one of economics, and the major case of this study uses reconstitution.

The heat and material balances used in final design were based on recent data from the pilot plant. After considerable study, a four-train system was selected for the gasification section, Area 200. Variables considered in choosing four trains included practical vessel sizing, solids feeding, acceptor distribution, gas distribution, plant startup, and plant operating factor. In general, the four-train system was also used in the downstream processing sections.

In the interest of providing a comparison with the first generation gasification processes, an economic evaluation of a Lurgi plant is included. The Lurgi data is based on the American Natural Gas Company study, modified for a specific site and coal.

Essentially, the basis of economic data and evaluation was the C F Braun & Company Guidelines Report (10), with appropriate modification to suit the specific site and the specific coal.

SUMMARY

INTRODUCTION

This section of the report summarizes the results of the design and cost estimates for the CO_2 Acceptor and the Lurgi Plants. Both plants are based on producing 250 billion Btu/day of pipeline gas from Renner's Cove lignite in North Dakota. The design of the CO_2 Acceptor plant was optimized by preliminary studies. Design, capital and operating cost details are given in other sections of this report. The Lurgi design, capital need and operating costs were developed based on American Natural Gas (ANG) estimates reported in Docket No. CP-75-278. Design and cost information for the Lurgi plant is contained in Appendix 1 of this report.

A study was made to prove reconstitution of acceptor is more economic than making up with fresh stone. Details of this study are reported in Appendix 2.

A study was also made to estimate capital, operating, and gas cost for a CO_2 Acceptor plant having 1/4 the capacity of the North Dakota Lignite plant. Details of this study are reported in Appendix 3.

NORTH DAKOTA LIGNITE PIPELINE GAS PLANTS

The major feed, product and by-product streams, capital and operating costs, and gas costs of the full size CO_2 Acceptor and Lurgi plants are summarized below: (Cost basis is mid-1977.)

Major Feed Streams	CO_2 Acceptor	Lurgi
Coal Feed, dry basis, tpsd	16,693.8	20,977.0
Water in Coal Feed, tpsd	10,059	12,650.0
Water Supply, net makeup, gpm	3,709	7,650
Products-Stream Day Basis		
Pipeline Gas, MMscfd	262.63	255
Sulfur, ltpd	124.5	161.6
Ammonia, tpd	101.3	135
Tar Oil, gal/day		134,537
Naphtha, gal/day		45,223
Crude Phenols, gal/day		27,493
Coal Fines, tpd		2,886

SUMMARY - continued

INTRODUCTION - continued

Product Gas Composition-Mol Percent	CO_2 Acceptor	Lurgi
Methane	92.63	95.95
Hydrogen	4.65	3.00
Carbon Dioxide	0.50	0.40
Carbon Monoxide	0.01	0.05
Nitrogen	2.21	0.60
Total Mols/Hr	28,873.6	28,034.3
Total Lbs/Hr	455,496	442,942
Molecular Weight	15.8	15.8
Heating Value, Btu/scf (hhv)	952.07	979

CAPITAL AND OPERATING COSTS
(In Millions of Dollars)

	CO_2 Acceptor	Lurgi
Total Plant Investment (excluding coal mine	$717.14	$ 814.20
Initial Catalyst and Chemicals	4.80	7.81
Paid-Up Royalties	1.21	0.97
Allowance for Funds Used During Construction	121.02	137.40
Startup Costs	24.60	26.68
Working Capital (Private Financing)	21.61	24.07
TOTAL CAPITAL REQUIREMENT	$890.38	$1,011.13
TOTAL GROSS ANNUAL OPERATING COSTS	$123.01	$ 133.41
BY-PRODUCT CREDITS	5.68	35.73
TOTAL NET ANNUAL OPERATING COSTS	$117.33	$ 97.68
Gas Costs - Utility Financing		
Average Gas Cost, $/MMBtu	$ 2.73	$ 2.67
First Year Gas Cost, $/MMBtu	3.43	3.46
Gas Cost - Private Financing		
Constant Gas Cost, R/MMBtu	$ 3.70	$ 3.77

SUMMARY BALANCES

Summary energy balances for the CO_2 Acceptor and Lurgi Plants are shown on Table 3-1. A Summary Material Balance for the CO_2 Acceptor Process is given in Table 3-2. A Summary Material Balance in detail as presented for the CO_2 Acceptor plant is not available for the Lurgi plant.

TABLE 3-1

SUMMARY ENERGY BALANCE

	CO_2 ACCEPTOR		LURGI	
	MMBTU/HR	PERCENT OF TOTAL	MMBTU/HR	PERCENT OF TOTAL
Energy Input				
Coal To Gasifier, hhv	14,552	94.3	15,057	82.4
Coal to Wet Bottom Furnace, hhv	485	3.1		
Coal to Acceptor Reconstitution, hhv	405	2.6		
Coal to Steam Plant, hhv			2,680	14.8
Electric Power Consumption (1.25 MMkwh/day) (Coal Equiv. @ 10,000 BTU/Kwh)			519	2.8
Total Input	15,442	100.0	18,256	100.0
Surplus Fines			1,666	
Energy Distribution				
Product Gas, hhv	10,419	67.5	10,419	57.1
By-Products:				
Tar Oil			746	4.1
Naphtha			232	1.3
Crude Phenols			138	0.7
Ammonia	83	.5	110	0.6
Sulfur	46	.3	60	0.3
Subtotal, Product and By-Products	10,548	68.3	11,705	64.1
Consumption and Losses	4,894	31.7	6,551	35.9
Total Energy Distribution	15,442	100.0	18,256	100.0
Surplus Fines			1,666	
Cold Gas Efficiency, %		67.5		57.1
Plant Thermal Efficiency, %		68.3		64.1

TABLE 3-2

SUMMARY MATERIAL BALANCE

CO_2 ACCEPTOR PROCESS

INLET STREAMS	LB/HR	PERCENT OF TOTAL
Coal to Process, dry	1,391,151	13.99
Water in Process Coal	838,258	8.43
Acceptor Makeup (limestone)	10,089	0.10
Raw Water	1,854,370	18.64
Air to Regenerator (wet)	2,796,094	28.11
Air to CO Oxidizer (wet)	284,260	2.86
Air to Wet Bottom Furnace (wet)	458,132	4.61
Air to Reconstitution (wet)	418,996	4.21
Air to SO_2 Removal (wet)	1,818,087	18.28
Air to Sulfur Recovery (wet)	39,481	0.40
Rain Water	12,020	0.12
Process Sewer Water	24,990	0.25
TOTAL	9,945,928	100.00

OUTLET STREAMS	LB/HR	PERCENT OF TOTAL
Product Gas	455,496	4.58
Ammonia	8,444	0.09
Sulfur	11,617	0.12
Acceptor Reconstitution Vents	777,897	7.82
Flue Gas from SO_2 Scrubber	7,544,139	75.85
Waste Solids (dry)	177,645	1.79
Evaporation (Tailings Pond)	64,831	0.65
Revegetation Water	247,011	2.48
Cooling Tower Losses	609,962	6.13
Steam and Water Losses	40,070	0.40
Dust Suppression Water	3,000	0.03
Miscellaneous Vents and Losses	5,816	0.06
TOTAL	9,945,928	100.00

TEXAS LIGNITE SYNTHESIS GAS PLANT

The design of the North Dakota Lignite Pipeline Gas plant was modified to represent a Texas Lignite Plant producing synthesis gas. Major feed, product and by-product streams, capital and operating costs and gas cost are summarized below.

Major Feed Streams

Coal Feed, dry basis, tpsd	17,169.0
Water in Coal Feed, tpsd	10,345.4
Water Supply, net make up, gpm	4,546

Products - Stream Day Basis

Synthesis gas, MMscfd	774.87
Sulfur, ltpd	124.5
Ammonia, tpd	101.3

Product Gas Composition - Mol Percent

Methane	11.44
Hydrogen	65.53
Carbon Monoxide	15.74
Carbon Dioxide	4.66
Nitrogen	0.75
Water	1.88
Total Mols/Hr	85,187.8
Total Lbs/Hr	886,014
Molecular Weight	10.17
Heating Value, Btu/scf (hhr)	379.1

A summary of Capital and Operating Costs and a Summary Energy Balance are shown on the following tables. A complete discussion of the Texas Lignite plant appears in its own section of this report.

TABLE 3-3

CAPITAL AND OPERATING COSTS
(In Millions of Dollars)
TEXAS LIGNITE SYNTHESIS GAS PLANT

Total Plant Investment (excluding coal mine)	$ 624.57
Initial Catalyst and Chemicals	2.28
Paid-Up Royalties	1.18
Allowance for Funds Used During Construction	105.40
Startup Costs	22.83
Working Capital (Private Financing)	19.47
TOTAL CAPITAL REQUIREMENT	$ 775.73
TOTAL GROSS ANNUAL OPERATING COSTS	$ 114.13
BY-PRODUCT CREDITS	5.68
TOTAL NET ANNUAL OPERATING COSTS	$ 108.45
GAS COST (PRIVATE FINANCING) $/MMBtu	$ 2.81

TABLE 3-4

SUMMARY ENERGY BALANCE

TEXAS LIGNITE SYNTHESIS GAS PLANT

	MMBTU/HR	PERCENT OF TOTAL
Energy Input		
Coal to Gasifier, hhv	14,552	91.6
Coal to Wet Bottom Furnace, hhv	925	5.8
Coal to Acceptor Reconstitution	405	2.6
TOTAL	15,882	100.0
Energy Distribution		
Product Gas, hhv	12,240	77.1
By-Products, hhv		
Ammonia	79	0.5
Sulfur	45	0.3
Subtotal, Product and By-Products	12,364	77.9
Consumption and Losses	3,518	22.1
TOTAL	15,882	100.0
Cold Gas Efficiency, %		77.1
Plant Thermal Efficiency, %		77.9

PROJECT SCOPE

The purpose of this project is to perform preliminary designs and cost estimates of commercially sized Carbon Dioxide (CO_2) Acceptor Process Coal Gasification plants. These designs are to have incorporated into them new cost reducing and efficiency producing concepts. The designs are not necessarily based upon generalized Coal Gasification Plant design bases. Instead, the bases are more specific and include an actual coal and a specific plant location. As a result, the design can be tailored for lowest cost and maximum efficiency.

In order to achieve the results of low cost and high efficiency, the work progressed in two phases. Phase I was comprised of selected studies to evaluate low cost or high efficiency producing alternatives. The studies considered such alternatives as economics of different number of gasification reactors, energy recovery systems, gas treating processes, and others. The alternatives selected for study were based upon the judgment of the design engineers and the time available for study. There may be other viable alternatives which were not considered.

Phase II of the work consisted of incorporating the selected processes in a final design and making a cost estimate. Results of this work are presented in this report. Both phases involved the interchange of ideas between CCDC and Stearns-Roger.

Engineering work was performed in only enough detail to enable a selection among alternatives or to provide data for the cost estimate. Cost estimating, in general, was made by computer methods and in-house estimating procedures. However, Coal Preparation and Acceptor Reconstitution Areas, and the Gasification structure, are considered so unique that generalized estimating methods do not apply. Therefore, sufficient layouts and physical designs were made to enable partial material takeoffs. Further, the Gasification structure is modeled, thus improving arrangement perspective and aiding in the material takeoffs. None of the design is considered accurate enough for construction. Further detailed engineering is necessary to provide that basis. The work, however, is adequate to obtain a representative plant cost estimate.

During the work, some of the process units in which one receives feed or recycle from another were designed concurrently. Such procedure is acceptable but iterative adjustment is then necessary to eliminate inconsistencies. The tight schedule required that equipment service conditions be established for estimate prior to any iterative adjustment. Therefore, some inconsistencies may exist between various project documents such as flow diagrams and equipment data lists. These inconsistencies, however, should not affect the accuracy of the cost estimate.

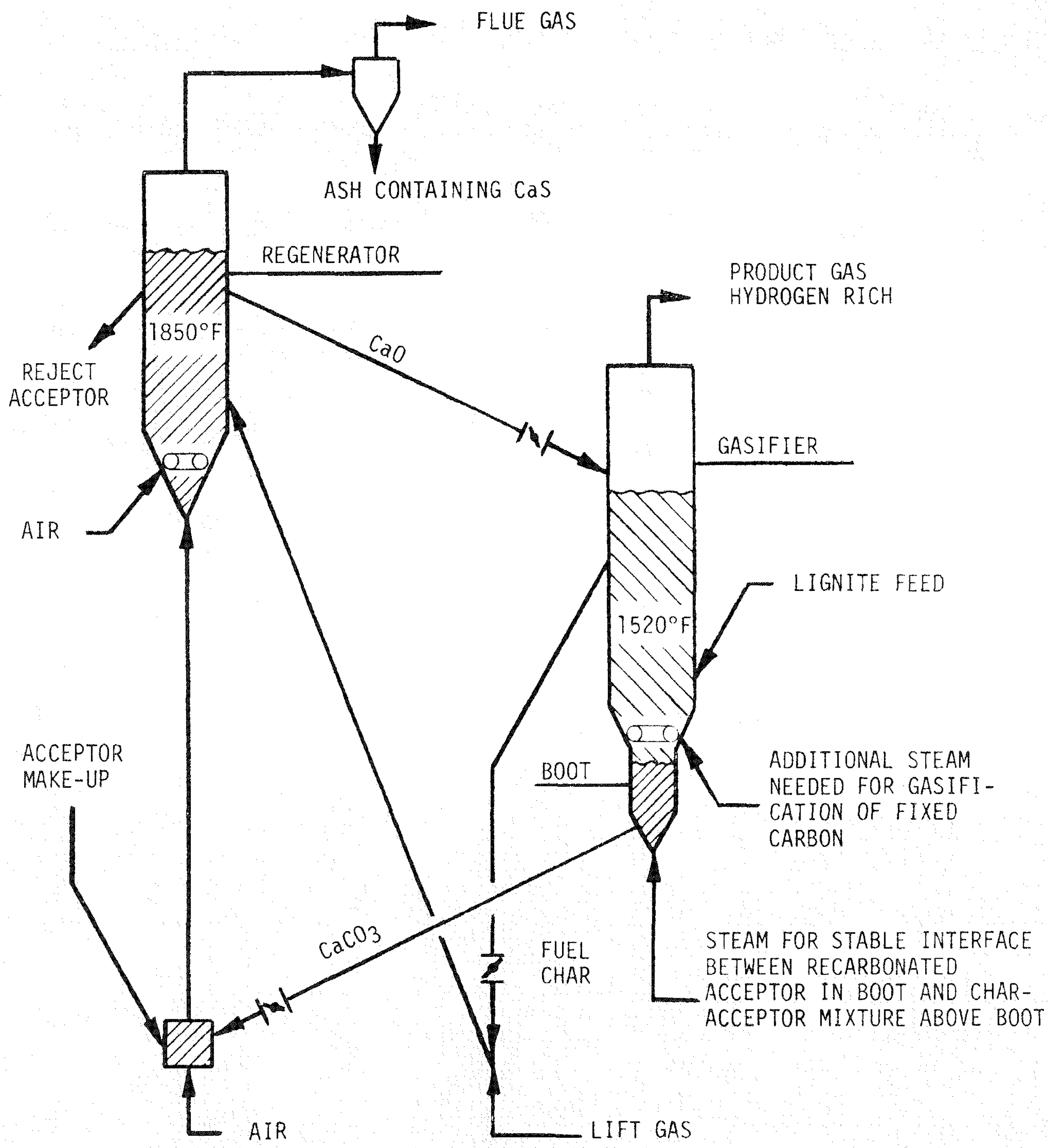
Two CO_2 acceptor plants were designed and estimated. The first plant designed was a plant to convert Renner's Cove lignite into pipeline quality gas. Design of the second involved modifying the first to produce a synthesis gas. The second design was performed by deleting appropriate processes from the first plant and then adjusting the utility and waste handling facilities as necessary. Coal feed and many of the front-end units were not changed. The second plant is representative of a plant processing Texas lignite to produce synthesis gas for use as a chemical plant feed stock or medium Btu clean fuel gas.

The scope of work includes a comparison of the cost of pipeline gas produced in the CO_2 Acceptor Plant and that produced in a Lurgi plant. A Lurgi plant has been proposed for the same area by American Natural Gas and others. Reports on this plant have been published in the technical literature. The published data was used as the basis for the Lurgi plant gas cost. Lurgi plant data was adjusted as possible to provide a basis comparable to that for the CO_2 Acceptor plant.

Results of the work, detailed design bases, and plant and process descriptions are presented in this report.

PROCESS DESCRIPTION

The diagram below shows the "heart" of the CO_2 Acceptor process. There are two fluidized bed reactors, a gasifier and a regenerator. The gasifier operates at 1490° to 1550°F and 130 to 165 psia, and the regenerator operates at the same pressure but at a temperature range of 1830 to 1880°F.



PROCESS DESCRIPTION - continued

Raw, dry lignite (or subbituminous coal) sized to nominally 8 x 100 mesh is fed at the bottom of a fluidized bed of char in the gasifier. After the rapid hydrodevolatilization reactions occur, fixed carbon in the char is gasified by the endothermic reactions:

$$C + H_2O_{(g)} \longrightarrow CO + H_2,$$

$$2\ C + H_2O_{(g)} + H_2 \longrightarrow CH_4 + CO$$

Carbon dioxide is formed by the water gas shift reaction.

$$CO + H_2O_{(g)} \longrightarrow CO_2 + H_2$$

The overall heat of reaction is about 33,000 Btu/lb-atom (60^{o}F) of coal carbon gasified. This endothermic heat of reaction is supplied by the CO_2 Acceptor reaction:

$$CaO + CO_2 \longrightarrow CaCO_3 \quad \Delta H = -76{,}200\ \text{Btu/lb-mol}\ (60^oF)$$

The acceptor can be derived from either limestone or dolomite. In this study Minnekahta limestone is used. The calcined limestone leaving the top of the regenerator, sized nominally to 4 x 10 Tyler mesh, enters above the fluidized char bed, showers through the bed, and collects in the gasifier boot. Steam flow to the bottom of the boot is adjusted to strip cleanly the char from the acceptor so that a sharp, stable interface exists between the acceptor and the char-acceptor mixture above it. The remaining steam needed for gasification of fixed carbon enters through a distributor above the boot.

The recarbonated acceptor flows through a standleg and then is conveyed pneumatically by air to the bottom of the regenerator. Residual char from the gasifier also flows through a standleg and is conveyed by a stream of regenerator recycle gas to the regenerator where combustion of the char raises the temperature of the fluidized bed of acceptor to about 1850^{o}F. At this temperature the acceptor is calcined by reversal of the CO_2 Acceptor reaction. The calcined acceptor is returned by gravity through a standleg, thus completing the acceptor "loop".

Seals between the gasifier and regenerator are maintained by the four solids standlegs which are purged with recycle gas. Solids flow rates are controlled by butterfly valves or lift pot controls.

After combustion, the char ash (approximately 80% - 100 mesh) is elutriated from the acceptor bed and removed by external cyclones.

As the acceptor circulates between the process vessels, it loses activity toward the CO_2 Acceptor reaction. Inert CaO forms by crystallite size growth. To maintain activity some acceptor is withdrawn continuously from the regenerator and fresh stone makeup is added to the acceptor lift line. Makeup requirements are 2 mols of CaO per 100 mols of CaO flowing through the gravity return line.

The CO_2 Acceptor process consumes substantially all (99+%) of the carbon fed to the gasifier, and is totally self-sufficient with respect to gasifier steam generation and regenerator air compression requirements. The gasifier process steam is generated and superheated by heat exchange with the gasifier and regenerator offgases. The partially cooled regenerator gas is expanded through turbines which drive the air compressors. There is no fuel fired boiler.

Use of the acceptor has five principal advantages:

1. The acceptor is heated and calcined in a separate reactor where air supplies the oxygen for combustion without contaminating the product gas. This eliminates the need for an oxygen plant.

2. The required circulation rate is lower than in other heat carrier processes since most of the heat is supplied chemically. About 80% of the total heat supplied by the acceptor comes from the CO_2 Acceptor reaction.

3. The acceptor reacts with both H_2S and CO_2, the principal impurities in the gasifier product gas, thus minimizing the gas clean-up requirements.

4. The raw gasifier product gas is hydrogen-rich, compared with all other gasification processes which use steam. No water-gas shifting is required prior to methanation because the raw gas contains enough H_2 to methanate all of the CO and part of the CO_2.

5. The cost of environmental compliance is reduced because no tars, oils or phenols appear in the aqueous effluents.

Since the acceptor also is an efficient H_2S acceptor, the gasifier product gas contains very little sulfur. The distribution of coal sulfur is shown below.

	% of Sulfur in Dry Coal Fed to Gasifier
In Raw Gasifier Gas	8
In Regenerator Flue Gas	18 (as SO_2)
In Regenerator Overhead Ash	74

The sulfur in the regenerator ash is present as CaS. Sulfur is recovered by slurrying the ash with hot liquid water in the presence of CO_2 via the reaction

$$CaS + H_2O_{(1)} + CO_2 \longrightarrow CaCO_3 + H_2S$$

The innocuous $CaCO_3$ is dewatered and returned to the mine. The H_2S-rich stream is converted to elemental sulfur by conventional Claus technology. The Claus tail gas is incinerated and sent to the SO_2 Removal Unit. All sulfur containing waste gases are then scrubbed in this unit to comply with environmental standards.

The process design described in this report is based on use of a North Dakota lignite which has a relatively high sodium content. Specifically, the sodium content is 8 wt. % Na_2O in the SO_3-free ash. There are no large reserves of lignite in the Northern Great Plains area which have average sodium contents appreciably less than this amount. At gasifier conditions above about 1600^{o}F, the sodium combines with silica and alumina in the ash to form a thermally unstable liquid which, upon decomposition, cements together the char particles. The pilot plant and bench-scale experience indicates strongly that the CO_2 Acceptor process, with its inherently low gasifier bed temperature, is the only known fluidized bed process which can cope with these lignites.

STATE OF PROCESS DEVELOPMENT

INTRODUCTION

The CO_2 Acceptor pilot plant has been operated over the period from April 1972 through September 1977. Seventy-five separate runs, defined by startup from cold conditions, have been completed.

The plant was at system pressure (150 psig) for 16,000 hours and was at normal operating conditions of pressure and temperatures for 6,400 hours. Six thousand five hundred tons of dry coal were gasified. Process conditions are defined as: (1) the acceptor is circulating continuously, (2) both the gasifier and regenerator are at programmed temperatures, (3) combustion of char is supplying all the regenerator heat duty, and (4) the circulating acceptor is supplying all or part of the gasifier heat duty.

Most of the difference between being at system pressure and being at process conditions occurred during the first two years when problems involving proper startup procedure and severe corrosion of the fired heater coils occurred and were solved. The fired heaters are only a pilot plant necessity and will not be used in a large plant.

During the period the following coals were gasified successfully:

A. Three North Dakota lignites; Velva, Glenharold, and Husky.

B. One Texas lignite.

C. Three subbituminous coals; Sarpy Creek and Rosebud (Montana), and Wyodak (Wyoming).

Detailed operating data on the CO_2 Acceptor process were obtained with the Velva, Glenharold, Texas, and Wyodak coals.

Plant operations at steady conditions produced 12 heat and material balances. The length of runs covering these operations varied from 130 to 290 hours at process conditions of pressure and temperatures. Of particular importance are those runs where the entire gasifier heat duty was supplied by the recirculating acceptor (otherwise, small amounts of air were fed to the gasifier to make up the difference). These specific runs defined process conditions with Velva, Glenharold, and Wyodak coals and lasted a total of 1200 hours. In other runs which did not provide heat and material balances, the acceptor supplied the entire gasifier heat duty for an additional 1000 hours. An additional nine heat and material balances were produced from runs at non-process conditions in

INTRODUCTION - continued

which either the regenerator or gasifier was operated individually to obtain data on heat losses and on gasification kinetics.

DISCUSSION

The following listing covers major process and mechanical achievements:

Gasification Solids Flow and Pressure Balance

Solids flow through three standlegs and two pneumatic lift lines provides the energy transfer essential to the CO_2 Acceptor Process. Additionally, three fluid beds within the reactors must be controlled independently as follows:

(1) The gasifier char bed must be expanded sufficiently to allow the acceptor to shower through it.

(2) The acceptor bed in the gasifier boot must be fluidized with sufficient steam to strip char from it, but at a steam rate low enough to permit the acceptor to collect below a stable interface with the char bed above it.

(3) The regenerator bed must be fluidized to give adequate mixing to prevent hot spots with ash slagging, yet provide enough retention of char to complete combustion of the carbon.

Successful operation of the entire system had been accomplished by Run 8 in the pilot plant program. The key to good control lies in the knowledge of fluosolids mechanics gained from bench-scale work done by CCDC in Library, Pennsylvania. The pilot plant was designed on the basis of this knowledge.

Operations of the standlegs, pneumatic lift lines, butterfly flow control valves, acceptor showering, the gasifier boot, and the regenerator bed have proven to be reliable. No fundamental changes have been made to the as-built system.

Pilot Plant Reliability

An analysis of the causes of shutdowns in the 10 runs which provided the 12 heat and material balances referred to above can give a general view of reliability:

Gasification Solids Flow and Pressure Balance - continued

Cause of Shutdown

electric utility power failure (1)
weld failure (2)
plugged pressure tap (1)
ball valve failure (1)
control instrument failure (1)
regenerator cyclone erosion (1) (orderly shutdown)
voluntary (3)

During the calendar period of January through August, 1977, 9 runs were made and the pilot plant was onstream (at pressure) for 40% of the time. The acceptor supplied the entire gasifier heat during 70% of the hours in which it was possible to do so.

Coal Grinding and Drying

Conventional equipment is used. The hot gas-swept mill has proven to be satisfactory. Safety is assured by operating the system at a slight positive pressure and by maintaining the oxygen concentration at about 5 mol %.

Coal Feeding

A reliable lockhopper system was developed early in the pilot plant program.

Instrumentation

Standard hardware has been used with complete success. There are no exotic systems.

Reactor Refractories

Both the regenerator and gasifier were relined early in the pilot plant program to correct a design deficiency. Since then the gasifier refractory has been in use for 64 runs. In each run a complete temperature-pressure cycle occurred, starting and ending with an open, cold reactor with operation at full process pressure and temperature in between.

In the same period the regenerator, which operates at a much higher temperature than does the gasifier, has had three linings. The first lining had 23 temperature-pressure cycles as defined above, and the second lining had 30 cycles. The third, and current lining was installed in order to reduce the working diameter. It has undergone nine cycles and is in excellent condition.

Reactor Refractories - continued

There has been no appreciable interaction of either refractory system with the acceptor or with coal ash. The pilot plant experience clearly shows that a large plant, with infrequent cycling, will have no fundamental problems with refractories.

Environmental Study

An intensive sampling and analytical program has been completed which will lead to comprehensive definition of all process streams and effluent streams in the pilot plant.

Although final data workup is not yet complete, it appears that the envisioned environmental controls for the commercial process will be satisfactory.

Methanation

For production of SNG, the CO_2 Acceptor process, as with any other gasification process, is not wedded to any particular methanation process.

The packed tube methanation reactor, cooled by boiling Dowtherm, was assigned for study to the CO_2 Acceptor pilot plant by the Office of Coal Research. Accomplishments during operation of the methanation plant are listed below. All fresh feed gas to the methanator was made in the CO_2 Acceptor process gasifier.

A. Successful operation on the first startup.

B. Demonstration of rapid kinetics at the low pressure of 110 psig.

C. Production of SNG made from coal with an actual heating value greater than 900 Btu/scf.

D. Data obtained with a recycle/fresh feed ratio of less than unity.

E. Data obtained at 280 psig by use of the booster compressor.

CONCLUSION

The pilot plant experimental program will end on September 30, 1977. Operation of the pilot plant has accomplished the original goals which include demonstration of feasibility of the CO_2 Acceptor process on several different low-rank coals at full process temperatures and pressures. The next stage of development must be construction and operation of a demonstration plant.

DESIGN BASIS

INTRODUCTION

The design of the CO_2 Acceptor Process is based on a grass roots facility for producing 250 billion Btu/Day of synthetic natural gas. Its definition is based on a "point" design basis. This is a result of assuming that a demonstration plant has previously been operated and all required process and equipment data is available for a "point" design basis. Some of the equipment used in this study is larger than presently available. A basis was set that larger than presently available equipment could be used if vendors agreed that the larger equipment could or would be available in the near future.

The basis for design of several of the plant areas was set by preliminary process and equipment comparisons.

GASIFICATION

CCDC provided the overall gasification process including a heat and material balance around the gasifier-regenerator loop. This heat and material balance was designed to produce a product gas containing approximately 250 Billion Btu/day (HHV). Calculated heat losses are included in the gasifier and regenerator loop.

CCDC also provided design data for the gasifier and regenerator section, such as, bed heights, bed densities, fluidizing gas velocities, particle size distribution, etc. This data is included in the Area Description Section (Area 200) of this report.

Other data and information from the Rapid City pilot plant was used in this design. Start-up criteria, major controls and pressure balance determinations for the gasifier regenerator loop based on pilot plant experience were used where applicable.

COAL ANALYSIS

CCDC provided a coal analysis for the North Dakota Renner's Cove lignite used in this design. Data and analysis are:

Type of Coal	Renner's Cove Lignite
Coal Size (Supplied to Plant)	2" x 0, Max. 2 wt%-100 mesh
Bulk Density	50 lbs/ft^3 (assumed)

DESIGN BASIS - continued

COAL ANALYSIS - continued

Storage	10 days dead storage 4 days live storage

Ultimate Analysis, wt%, 1970 analysis

	Wt. Percent
Moisture	37.6

Dry Basis	Wt. Percent
Hydrogen	4.35
Carbon	65.24
Nitrogen	0.89
Oxygen	20.83
Sulfur	1.11
Ash	7.58
	100.00

Heating value of dry coal, Btu/lb (HHV)	11,100
Heating value of coal as received Btu/lb (HHV)	6,926

Ash Analysis	Wt. Percent
Na_2O	7.75
K_2O	0.56
CaO	32.20
MgO	10.40
Fe_2O_3	10.00
TiO_2	0.72
P_2O_5	0.39
SiO_2	23.50
Al_2O_3	13.60
Other	0.88

LIMESTONE

Limestone is required as acceptor make-up. Since the limestone is fed directly to and calcined in the Acceptor Reconstitution plant a sized material is not required. Reconstitution produces 4 x 10 Tyler mesh pellets as required in the gasification area. It is therefore assumed that quarry fines would be used as limestone make-up.

Limestone data and analysis are as follows:

DESIGN BASIS - continued

LIMESTONE - continued

Type	Minnekahta
Supplied to plant by	Rail hopper car
Storage	30 days at ave. make-up
Size	Quarry fines
Particle density	165 lb/ft^3
Bulk density	80 lb/ft^3
Moisture	1 wt. percent

Analysis	Wt. Percent
CaO	94.2
MgO	1.1
SiO_2	3.7
Al_2O_3	0.8
Fe_2O_3	0.4
Na_2O	0.05
K_2O	0.05
P_2O_5	0.04
TiO_2	0.20
SO_3	0.05
	100.59

SITE CONDITIONS

Since a selected plant site location was used specific data and information for the site such as, meteorological data, proximity to water, coal mine, access roads, and rail facilities is as used. Site conditions, water supply and analysis, access road tie-in, rail tie-in and power tie-in are described below.

General design conditions for the selected plant site are:

Elevation, feet	1950
Normal atmospheric pressure, psia	13.7
Summer dry bulb temperature, °F	95
Summer wet bulb temperature, °F	74
Winter dry bulb temperature, °F	-24
Design frost depth, feet	4.5
Wind loading at 30',lb/sq.ft.	25
Prevailing wind direction	NW
Earthquake zone	1

DESIGN BASIS - continued

WATER SUPPLY

Water supply is from Lake Sakakawea approximately 2.5 miles northwest of the plant site. A pumping station at the lake plus a water pipeline from the lake to the plant site are included in the design and cost estimate. Elevation difference between the lake and plant site is 200'. Pumping equipment is designed to provide this lift. It was assumed that water would be taken from the lake at no cost. Mean and maximum raw water analysis are as shown below.

Component	Mean PPM	Maximum PPM
Calcium (as $CaCO_3$)	50	65
Magnesium (as $CaCO_3$)	20	23
Sodium (as $CaCO_3$)	58	69
Potassium (as $CaCO_3$)	4	8
Bicarbonate (as $CaCO_3$)	183	201
Chloride (as $CaCO_3$)	9	11
Sulfate (as $CaCO_3$)	173	190
Silica (as SiO_2)	7.4	8.1
Suspended Solids (JTU)*	2.25	7
Conductivity, Micro Mho	650	720
Total Alkalinity ($CaCO_3$)	150	165
Total Hardness (as $CaCO_3$)	207	250
Total Dissolved Solids (TDS)	414	454
Total Organic Carbon (TOC)	3.7	4.0
pH	8.1	8.6 (Min. 7.5)
Temperature, °F	45	61 (Min. 32)

* Jackson Turbidity units

LAND

It is assumed that the 215 acres of land required for the plant facilities, offsites, and coal and solids storage is available. Cost for this land is not included in the capital cost estimate. Mine is also not included in the study. Rail and road are assumed to exist to the section boundaries in which the plant is located.

POWER

Normal electrical power for the plant and river water pumping facilities is supplied from in-plant generation facilities. Also, diesel driven electrical generators are provided for start-up and emergency shut-down. Therefore, no external electrical power tie-ins are required or included in the estimate.

GAS PIPELINE

The product gas is assumed to be delivered to a major gas pipeline. However, a pipeline from plant battery limits to the major gas pipeline is not included in the cost estimate. A gas metering station is also not included. A product gas pipeline to plant battery limits is provided in the design and cost estimate.

PLANT DESCRIPTION

INTRODUCTION

The CO_2 Acceptor plant designed for this study is a grass-roots facility, processing North Dakota Renner's Cove lignite and producing synthetic natural gas as its principal product. The North Dakota plant was modified to simulate a plant producing synthesis gas from Texas lignite. The North Dakota Lignite plant, as the base plant, is completely described in the following report sections. The changes made in the North Dakota plant, and the resultant Texas Lignite plant, are discussed in the Texas Lignite Plant Discussion section of this report.

The North Dakota Lignite plant is designed to be completely self-sufficient. It contains all facilities necessary to convert coal to pipeline quality gas and to deliver that gas to a pipeline. It includes all support systems such as utility systems, power generation, and by-products recovery. Offsites include offices, laboratory, cafeteria, and maintenance shops.

PLANT PROCESS

Coal sized to 2" x 0" at the mine is received at the plant battery limits. It is conveyed to size reduction and coal drying equipment. The sized and dried coal (2 percent moisture) is fed to the fluid bed gasifiers. These operate at about 10 atmospheres and 1520°F. Here the bulk of the coal is gasified with steam. A complete description of the gasification system is presented in the section, Process Description.

Gasifier product gas passes through waste heat recovery equipment and is cooled to 300°F. Ammonia and hydrogen sulfide are removed from this raw gas. Ammonia is recovered as anhydrous Ammonia in an Ammonia recovery unit. Elemental sulfur is recovered from the hydrogen sulfide. The raw gas is then compressed to 225 psig and methanated. The methanated gas is treated for removal of most of the residual CO_2, compressed to 1000 psig, and dehydrated for pipeline delivery.

Regenerator off gas passes through high efficiency cyclones for ash removal. Ash removed from the cyclones is cooled and slurried. Sulfur is removed and the slurry pumped to a tailings pond where the solids settle out. The water is reclaimed.

Regenerator off gas from the ash cyclones goes to waste heat recovery where high pressure steam is generated and superheated. The gas, now at 1585° is contacted with air to oxidize CO. The reaction increases the gas temperature to 1800°F. The oxidized regenerator gas passes through additional waste heat recovery where more 1500 psig

PLANT PROCESS - continued

steam is generated and superheated. Low pressure steam is also superheated for use in the gasifier. The regenerator off gas now at 1209^{o}F is expanded in a hot gas expander for power recovery. The expanders drive the air compressors which supply air to the regenerator for combustion.

The depressurized gas at 845^{o}F is used as coal drying gas. The coal dryer effluent gas, containing particulates and SO_2 goes to SO_2 Scrubbing for final clean-up prior to being released to the atmosphere.

Regenerator gas not expanded, is cooled, compressed, and used as recycle char lift gas and coal lock hopper pressuring gas.

ENERGY UTILIZATION

During the preliminary study phase. it was agreed that, if possible, all fines (mines 100 mesh) produced in coal preparation would be consumed in the plant, i. e., no fines exported. The gasifiers can tolerate up to 10 percent minus 100 mesh fines. Coal fines are burned in the wet bottom furnaces for generating hot flue gas and used to dry the high moisture content lignite. Coal fines are also combusted in the acceptor reconstitution pelletizers to provide the necessary heat to calcine reject acceptor from the regenerators. The preliminary studies indicated that all fines produced would be consumed in the plant.

The preliminary studies also indicated that plant energy requirements could be balanced by energy recovered from process waste heat (no excess steam) without removing additional energy from the expanded regenerator flue gas. The gas from the hot gas expander is at 845^{o}F, based on 1209^{o}F inlet temperature. By using the 845^{o}F gas in coal drying, the heat required from the wet bottom furnaces and therefore coal burned, is minimized. Even with the minimized amount of coal burned in the wet bottom furnaces all coal fines are consumed in the plant. Also, the capital cost of the wet bottom furnaces is reduced, additional heat exchange equipment is not required for removing heat from the 845^{o}F regenerator flue gas and additional capital is not required for turbogenerators supplying export power.

PLANT LOCATION

The design work performed for this study is based on a plant site located in central North Dakota, approximately 12 miles north of Beulah. The plant site is near large deposits of lignite coal and is approximately 2.5 miles from a reservoir on the Missouri River. (Garrison Resevoir, Lake Sakakawea). Maps showing the selected plant site location are included at the end of this section.

PLANT AREAS

This coal conversion complex is designed to convert lignite coal to synthetic natural gas. The complex is composed of a series of interconnected process plants, utility facilities to support the process plants, offsites facilities, and general plant facilities to handle administration and maintenance. The plant is divided in areas according to major functions. These areas are:

PLANT AREAS - continued

AREA NUMBER	
100	Coal Preparation
200	Gasification
300	Gasification Services
400	Raw Gas Cooling
500	Raw Gas Treating (H_2S Removal & Compression)
600	Methanation
700	Product Treating & Compression (CO_2 Removal, Product Compression & Dehydration)
800	Flue Gas System (Power Recovery)
900	Ash Handling
1000	Acceptor Reconstitution
1100	Sulfur Recovery
1200	Ammonia Recovery
1300	SO_2 Removal
1400	Raw Water Supply & Treating
1500	Waste Solids Disposal
1600	Steam System
1700	Electrical
1800	Miscellaneous Utilities
1900	Offsites
2000	General Facilities

PLANT AREAS - continued

Area 100 - Coal Preparation

This area includes coal unloading facilities, storage and reclaiming facilities, crushing and drying equipment, dried coal storage and coal feed to process facilities.

Area 200 - Gasification

This area includes the gasifier and regenerator vessels, the intervessel solids transfer equipment and piping, coal and acceptor feed lockhoppers, reject acceptor withdrawal lockhoppers and gasifier solids removal cyclones.

Area 300 - Gasification Services

This area includes recycle gas circuit, lock gas system; start-up char and dead-burned limestone unloading, storage and transfer facilities; and fresh limestone unloading, storage and transfer equipment.

Area 400 - Raw Gas Cooling

This area includes waste heat recovery equipment, and solids scrubbing and ammonia scrubbing facilities.

Area 500 - Raw Gas Treating

Included in this area are H_2S removal facilities (Stretford) and raw gas compression equipment.

Area 600 - Methanation

Included in this area are trace sulfur removal equipment (ZnO guard beds), Methanation reactors, heat recovery equipment and methanation unit start-up facilities.

Area 700 - Product Treating and Compression

This area includes CO_2 removal unit (Benfield) two stages of product compression and a glycol dehydration unit.

Area 800 - Flue Gas System

This area includes high efficiency ash removal cyclones, waste heat recovery equipment, carbon monoxide oxidation, tertiary solids removal and power recovery by hot gas expanders.

Area 900 - Ash Handling

This unit includes heat recovery and cooling of ash, ash slurrying and treating facilities and ash slurry pumping facilities.

Area 1000 - Acceptor Reconstitution

This area includes transfer and storage of reject acceptor, hydration equipment, inert separation facilities, pelletizing equipment and reconstituted acceptor storage and transfer to process facilities.

Area 1100 - Sulfur Recovery

This area includes a three stage Claus unit and a Claus tail gas incinerator.

Area 1200 - Ammonia Recovery

This area includes ammoniacal water holding tanks, ammonia recovery unit (USS Phosam-W Process) and recovered water cooling and pumping facilities.

Area 1300 - SO_2 Removal

This area includes SO_2 and particulate scrubbing facilities for the various SO_2 and particulate containing gas stream in the plant.

Area 1400 - Raw Water Supply and Treating

This area includes fresh water pumping facilities and a pipeline from Lake Sakakewea to the plant site. Also included in this area are water clarification and BFW and potable water treating facilities. Firewater storage, pumping, and sewage treatment facilities are included in this area

Area 1500 - Waste Solids Disposal

This area includes facilities for accumulating and pumping waste streams from the plant to a tailings pond. Storage and pumping facilities for recycling tailings pond water back to plant site are included. Process sewage and storm water collection and treating facilities are also included in Area 1500.

Area 1600 - Steam System

This area includes BFW deareation and pumping equipment, start-up steam boilers, and condensate collection and transfer facilities.

Area 1700 - Electrical Power Generation

This area includes plant electrical power generation equipment (turbo-generators) and diesel-driven start-up and shut-down generators.

Area 1800 - Miscellaneous Utilities

This area includes the plant cooling water system, emergency instrument air receivers and dryers, dryers for CO_2 purge gas and nitrogen generator package.

Area 1900 - Offsites

This area includes the plant flare system and storage facilities that are not included within a process area.

Area 2000 - General Facilities

This area includes site preparation, roads, rail spur, interconnecting piping, plant sewers, fencing, paving, and buildings not directly associated with a specific area.

TRAIN ARRANGEMENTS

The CO_2 Acceptor Plant is based on multiple trains of equipment. The number of trains of equipment vary from area to area as well as within certain areas.

Plant train diagrams were developed to show the number of trains for the individual process areas and/or the number of major pieces of equipment that make up the plant complex. Major process piping header arrangements are also shown on these drawings. The four gasifier-regenerator trains were selected based on economic and size considerations. The number of trains for other process areas are based on equipment size limitations; or operating reliability for maintaining an overall plant service factor of 90 percent; or to fit with the four train gasifier-regenerator configuration. The Plant Train Diagrams, Dwg. No's. 00-1-03 and 00-1-04 are included at the end of this section.

PLANT DESCRIPTION - continued

OVERALL BLOCK FLOW DIAGRAM

An overall block flow diagram for the plant Dwg. 00-1-01, is included at the end of this section. This drawing shows all of the plant areas and the process sequence. Drawing 00-1-02 also included at the end of this section shows the primary flows between process areas consistent with Dwg. 00-1-01. Only major or primary streams between process areas are shown. For detailed material balance in an area, refer to individual area process flow diagrams in Book 2 of this report.

PLANT WATER BALANCE

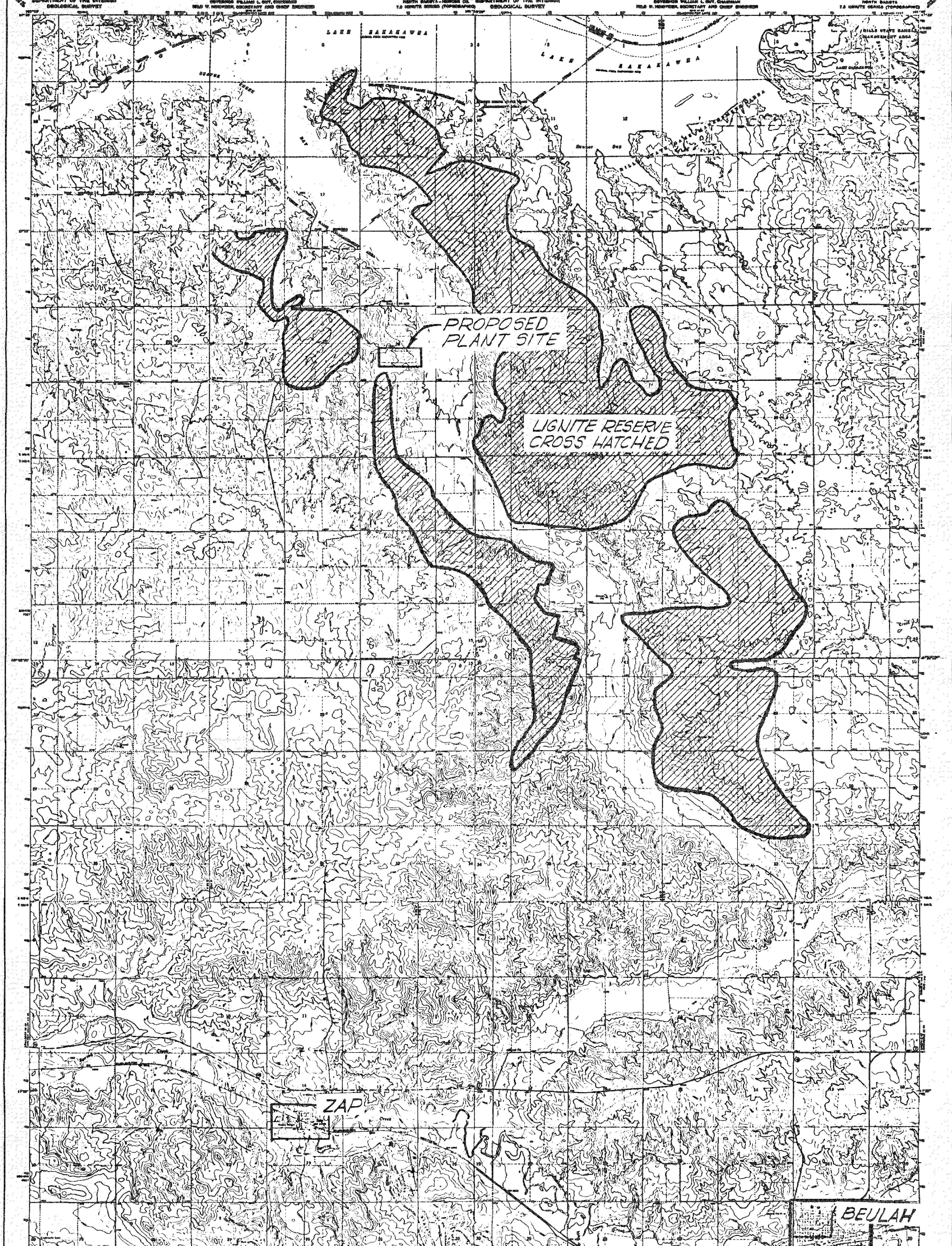
A plant water balance, Dwg. 00-1-05, is included at the end of this section. The balance was developed to establish raw water make-up and treating requirements. All water streams in the plant, including water vapor in process gas streams, are shown - approximately 3700 gpm (5383 acre-ft/year) of raw water are required. An average flow of about 500 gpm is shown for revegetating stripped mine areas. If this revegetating water is not required from the gasification plant, reclaiming of process water by evaporation, or other water reclamation processes could be used to reduce the plant raw water make-up requirement. Elimination of revegetation water would reduce raw water make-up to plant from 3700 gpm to approximately 3200 gpm.

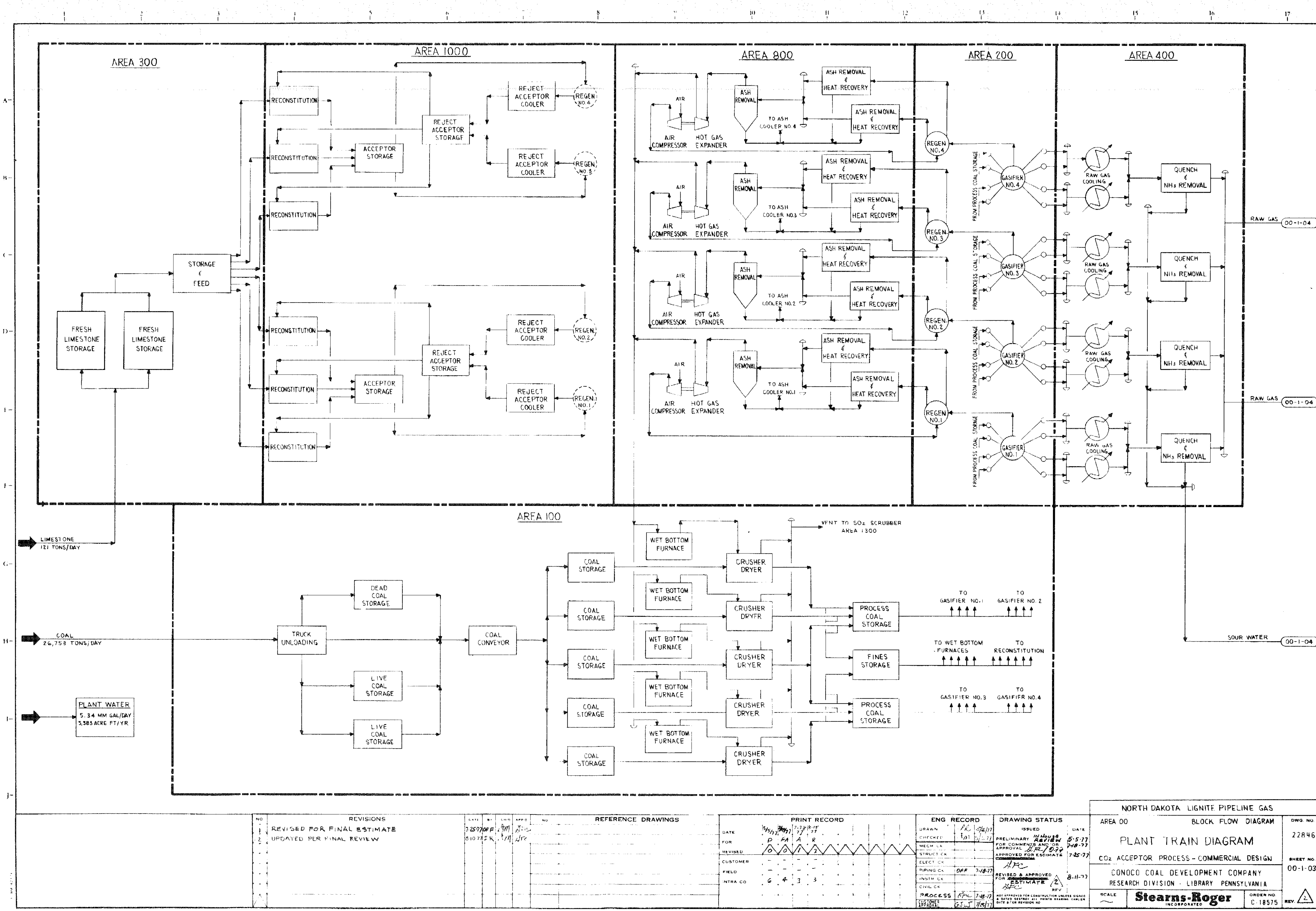
PLOT PLAN

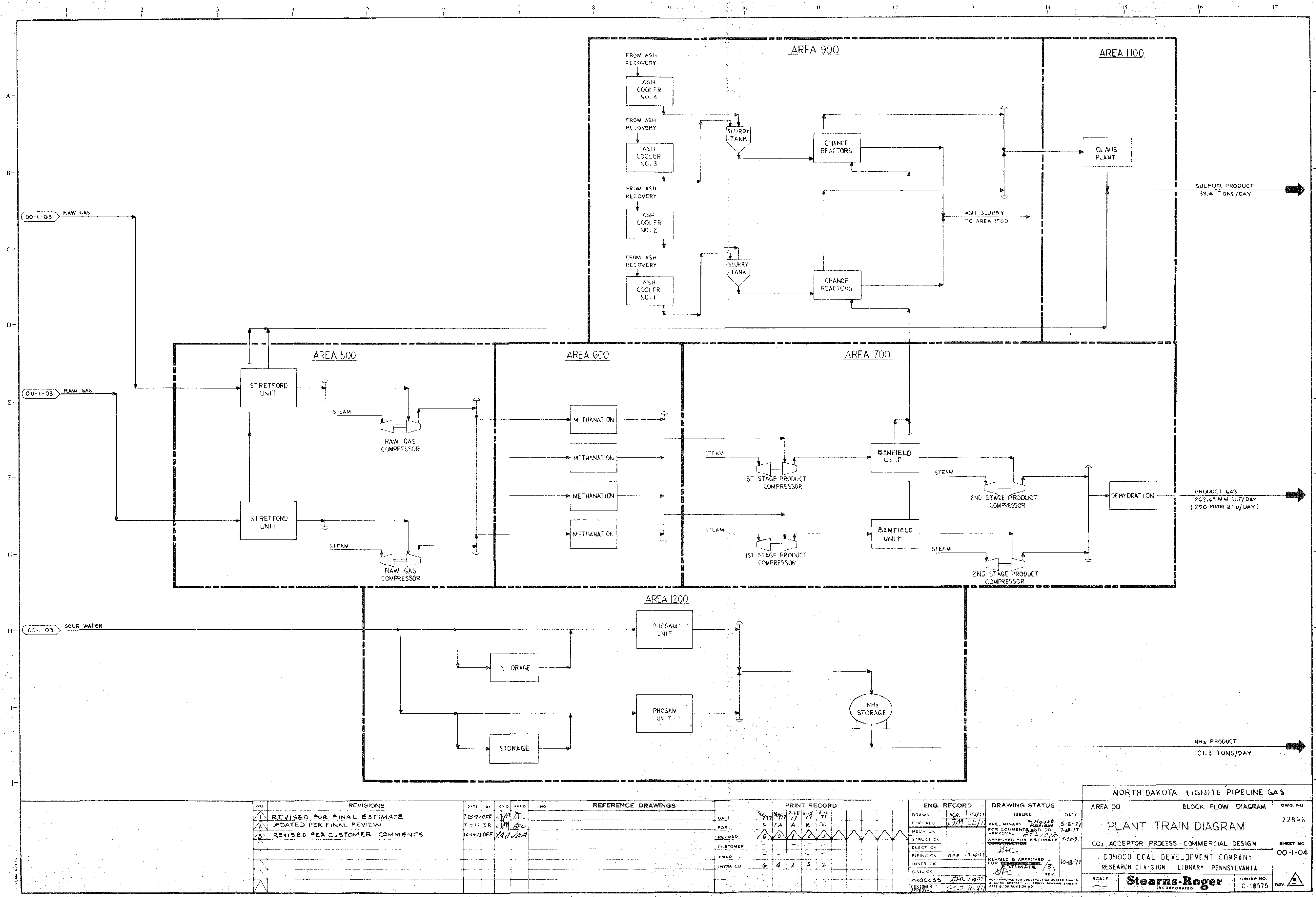
A preliminary plot plan was developed for the plant and is shown on Dwg. 00-2-01 included at the end of this section. The plot plan was developed to determine property area, site preparation quantities, and size of distribution and service systems. Areas arrangements are arbitrary and are based primarily on proximity to the area where major feed streams originate. Individual area plot sizes were estimated from major equipment sizes. Arrangement drawings, other than Areas 100, 200, 1000 and equipment in 400 and 800 that are located in the main gasifier structure, were not developed.

MINE

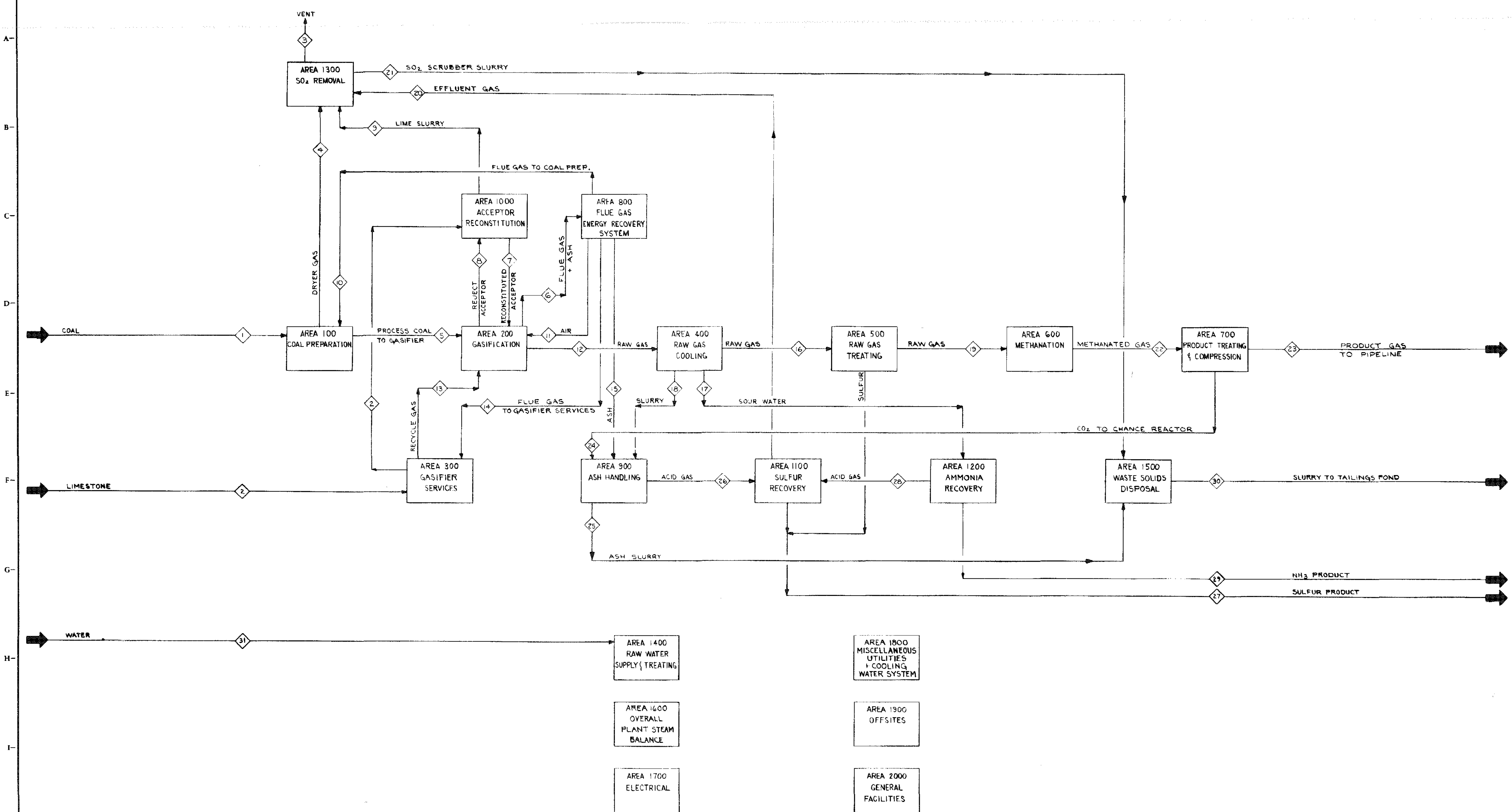
Mine costs are not included in the capital cost estimate. The basis for this study is that coal will be received from an independently developed and operated mine.


PROPOSED
PLANT SITE
LIGNITE RESERVE
CROSS HATCHED
ZAP
BEULAH


AREA 300
AREA 1000
AREA 800
AREA 200
AREA 400
AREA 100
FRESH LIMESTONE STORAGE
FRESH LIMESTONE STORAGE
STORAGE & FEED
RECONSTITUTION
ACCEPTOR STORAGE
REJECT ACCEPTOR STORAGE
REJECT ACCEPTOR COOLER
REGEN NO. 4
REGEN NO. 3
REGEN NO. 2
REGEN NO. 1
AIR
AIR COMPRESSOR
HOT GAS EXPANDER
ASH REMOVAL
TO ASH COOLER NO. 4
TO ASH COOLER NO. 3
TO ASH COOLER NO. 2
TO ASH COOLER NO. 1
ASH REMOVAL & HEAT RECOVERY
FROM PROCESS COAL STORAGE
GASIFIER NO. 4
GASIFIER NO. 3
GASIFIER NO. 2
GASIFIER NO. 1
RAW GAS COOLING
QUENCH & NH3 REMOVAL
RAW GAS
00-1-04
SOUR WATER
LIMESTONE
121 TONS/DAY
COAL
26,753 TONS/DAY
PLANT WATER
5.34 MM GAL/DAY
5,383 ACRE FT/YR
TRUCK UNLOADING
DEAD COAL STORAGE
LIVE COAL STORAGE
COAL CONVEYOR
COAL STORAGE
WET BOTTOM FURNACE
CRUSHER DRYER
VENT TO SO2 SCRUBBER AREA 1300
PROCESS COAL STORAGE
FINES STORAGE
TO GASIFIER NO. 1
TO GASIFIER NO. 2
TO WET BOTTOM FURNACES
TO RECONSTITUTION
TO GASIFIER NO. 3
TO GASIFIER NO. 4
REVISIONS
1 REVISED FOR FINAL ESTIMATE
2 UPDATED PER FINAL REVIEW
REFERENCE DRAWINGS
PRINT RECORD
ENG RECORD
DRAWING STATUS
NORTH DAKOTA LIGNITE PIPELINE GAS
AREA 00
BLOCK FLOW DIAGRAM
PLANT TRAIN DIAGRAM
CO2 ACCEPTOR PROCESS - COMMERCIAL DESIGN
CONOCO COAL DEVELOPMENT COMPANY
RESEARCH DIVISION - LIBRARY PENNSYLVANIA
Stearns-Roger INCORPORATED
DWG. NO. 22846
SHEET NO. 00-1-03
ORDER NO. C-18575
REV. 2



AREA 900
AREA 1100
FROM ASH RECOVERY
ASH COOLER NO. 4
ASH COOLER NO. 3
ASH COOLER NO. 2
ASH COOLER NO. 1
SLURRY TANK
CHANCE REACTORS
CLAUS PLANT
ASH SLURRY TO AREA 1500
SULFUR PRODUCT
139.4 TONS/DAY
00-1-03
RAW GAS
AREA 500
AREA 600
AREA 700
STRETFORD UNIT
STEAM
RAW GAS COMPRESSOR
METHANATION
1ST STAGE PRODUCT COMPRESSOR
BENFIELD UNIT
2ND STAGE PRODUCT COMPRESSOR
DEHYDRATION
PRODUCT GAS
262.63 MM SCF/DAY
(250 MMM BTU/DAY)
SOUR WATER
AREA 1200
PHOSAM UNIT
STORAGE
NH_3 STORAGE
NH_3 PRODUCT
101.3 TONS/DAY
REVISIONS
REVISED FOR FINAL ESTIMATE
UPDATED PER FINAL REVIEW
REVISED PER CUSTOMER COMMENTS
REFERENCE DRAWINGS
PRINT RECORD
ENG. RECORD
DRAWN
CHECKED
MECH CK
STRUCT CK
ELECT CK
PIPING CK
INSTR CK
CIVIL CK
PROCESS
CUSTOMER APPROVAL
DRAWING STATUS
ISSUED
DATE
PRELIMINARY IN HOUSE REVIEW 5-5-77
FOR COMMENTS AND OR APPROVAL 7-18-77
APPROVED FOR ESTIMATE 7-25-77
REVISED & APPROVED FOR ESTIMATE 10-13-77
NOT APPROVED FOR CONSTRUCTION UNLESS SIGNED & DATED DESTROY ALL PRINTS BEARING EARLIER DATE &/OR REVISION NO
NORTH DAKOTA LIGNITE PIPELINE GAS
AREA 00
BLOCK FLOW DIAGRAM
PLANT TRAIN DIAGRAM
CO_2 ACCEPTOR PROCESS - COMMERCIAL DESIGN
CONOCO COAL DEVELOPMENT COMPANY
RESEARCH DIVISION - LIBRARY PENNSYLVANIA
Stearns-Roger INCORPORATED
DWG. NO. 22846
SHEET NO. 00-1-04
ORDER NO. C-18575
REV. 3



NOTES:

1. NO CLOSED MATERIAL BALANCE CAN BE MADE AROUND ANY ONE PLANT AREA BLOCK ON THIS P.F.D. SINCE ONLY PRIMARY FLOWS ARE SHOWN.
2. REFER TO DRAWING 00-1-05 FOR PLANT WATER BALANCE FLOWS.

FORM 92176

NO.	REVISIONS	DATE	BY	CH'D	APP'D
1	REVISED FOR FINAL ESTIMATE	7-25-77	OFF	LAG	JFC
2	UPDATED PER FINAL REVIEW	8-11-77	JLH	JH7	JFC

NO.	REFERENCE DRAWINGS

PRINT RECORD				
DATE	7/5/77	7/18/77	7-28-77	8-15-77
FOR	P	FA	A	R
REVISED	0	0	1	2
CUSTOMER	—	—	—	—
FIELD	—	—	—	—
INTRA CO.	4	4	3	3

ENG. RECORD		
DRAWN	LR	1/6/77
CHECKED	LAG	7/15/77
MECH. CK.		
STRUCT. CK.		
ELECT. CK.		
PIPING CK.	OFF	7-18-77
INSTR. CK.		
CIVIL CK.		
PROCESS	JFC	7-18-77
CUSTOMER APPROVAL	GJLT	7/27/77

DRAWING STATUS	
ISSUED	DATE
PRELIMINARY IN HOUSE REVIEW	7-15-77
FOR COMMENTS AND/OR APPROVAL JFC / 020	7-18-77
APPROVED FOR ~~CONSTRUCTION~~ ESTIMATE JFC	7-25-77
REVISED & APPROVED FOR ~~CONSTRUCTION~~ ESTIMATE JFC REV. 2	8-11-77

NOT APPROVED FOR CONSTRUCTION UNLESS SIGNED & DATED. DESTROY ALL PRINTS BEARING EARLIER DATE &/OR REVISION NO.

NORTH DAKOTA LIGNITE PIPELINE GAS

AREA 00 BLOCK FLOW DIAGRAM

OVERALL PLANT

PRIMARY FLOWS

CO_2 ACCEPTOR PROCESS COMMERCIAL DESIGN

CONOCO COAL DEVELOPMENT COMPANY

RESEARCH DIVISION LIBRARY PENNSYLVANIA

SCALE ~ | Stearns-Roger INCORPORATED | ORDER NO. C-18575 | DWG. NO. 22846 | SHEET NO. 00-1-01 | REV. 2

100 % PLANT FLOWS

STREAM IDENTIFICATION		COAL	LIMESTONE	VENT	DRYER GAS	PROCESS COAL TO GASIFIER	FLUE GAS + ASH	RECONSTITUTED ACCEPTOR	REJECT ACCEPTOR	LIME SLURRY	FLUE GAS TO COAL PREP.	AIR TO REGENERATOR	RAW GAS	RECYCLE GAS	FLUE GAS TO GASIFIER SERVICES	ASH
STREAM NO.		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
CH_4		—	—	—	—	—	—	—	—	—	—	—	9,762.8	—	—	—
CO		—	—	—	—	—	2,972.0	—	—	—	—	—	13,413.9	—	—	—
CO_2		—	—	45,436.9	43,497.0	1,189.5	41,501.7	—	—	—	40,750.7	—	5,226.9	1,963.7	2,391.1	—
H_2		—	—	—	—	—	69.4	—	—	—	—	—	55,838.8	—	—	—
N_2		—	—	146,393.7	92,891.4	2,343.9	79,902.9	329.9	—	—	80,316.7	76,267.0	638.0	3,862.2	4,702.7	—
NH_3		—	—	1.3	—	—	—	—	—	—	—	—	500.2	—	—	—
H_2S		—	—	—	—	—	5.1	—	—	—	—	—	34.6	—	—	—
SO_2		—	—	46.3	88.8	4.1	139.6	—	—	—	138.6	—	—	6.6	8.0	—
COS		—	—	—	—	—	6.5	—	—	—	—	—	—	—	—	—
O_2		—	—	14,071.2	910.3	13.9	—	87.7	—	—	479.3	20,278.6	—	23.2	28.2	—
H_2O (V)		—	—	54,870.9	47,963.7	48.8	1,688.8	2.4	—	—	1,671.1	564.9	23,055.6	77.8	94.8	—
S		—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
TOTAL MOLES/HOUR		—	—	260,820.3	185,351.2	3,600.2	126,286.0	420.0	—	—	123,356.4	97,110.5	108,470.8	5,933.5	7,224.8	—
MOLECULAR WEIGHT		—	—	28.92	29.22	33.23	33.17	28.79	—	—	33.23	28.79	12.15	33.23	33.23	—
TOTAL LBS/HR (VAPOR)		—	—	7,542,815	5,416,049	119,615	4,188,688	12,092	—	—	4,098,236	2,796,094	1,317,904	197,208	240,125	—
MAF COAL	LBS/HR	1,285,702	—	—	—	1,211,626	—	—	—	—	—	—	—	—	—	—
MAF CHAR	LBS/HR	—	—	—	—	—	5,146	—	—	—	—	—	—	—	—	—
H_2O (L)	LBS/HR	838,258	—	—	—	26,220	—	—	—	31,444	—	—	—	—	—	—
ASH	LBS/HR	105,449				99,374										
MgO	LBS/HR	—	—	—	—	—	10,335	—	—	—	—	—	—	—	—	10,335
$MgCO_3$	LBS/HR	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
CaO	LBS/HR	—	—	—	—	—	13,025	113,813	113,813	—	—	—	—	—	—	13,025
$CaCO_3$	LBS/HR	—	9,943	—	—	—	—	—	—	—	—	—	—	—	—	—
$Ca(OH)_2$	LBS/HR	—	—	—	—	—	—	—	—	5,558	—	—	—	—	—	—
CaS	LBS/HR	—	—	—	—	—	24,406	—	—	—	—	—	1,450	—	—	24406
OTHERS (3)	LBS/HR	—	146	1,324	—	4,064	57,037	20,084	20,084	1,462	—	—	—	—	—	62,183
TOTAL LBS/HR	(SOLIDS)	2,229,409	10,089	1,324	—	1,341,284	104,803	133,897	133,897	38,464	—	—	1,450	—	—	109,949
TOTAL LBS/HR		2,229,409	10,089	7,544,139	5,416	1,460,899	4,298,637	145,989	133,897	38,464	4,098,236	2,796,094	1,319,354	197,208	240,125	109,949
OPER. TEMP., °F		AMB.	80	198	200	180	1,848	350	900	120	845	415	1,520	245	1,209	1,845
OPER. PRESS., PSIG		—	—	+1" W.C.	—	8.0	108.4	4.0	108.8	—	10	146.3	116.6	146.3	98	—
TONS/DAY		26,753	121.1	—	—	—	—	—	—	—	—	—	—	—	—	—
MM SCFD		—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
MM GAL/DAY		—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
ACRE FT./YR.		—	—	—	—	—	—	—	—	—	—	—	—	—	—	—

STREAM IDENTIFICATION		RAW GAS	SOUR WATER	SLURRY TO CHANCE REACTORS	RAW GAS	EFFLUENT GAS	SO_2 SCRUBBER SLURRY	METHANATED GAS	PRODUCT GAS	CO_2 TO CHANCE REACTORS	ASH SLURRY	ACID GAS	SULFUR PRODUCT	ACID GAS	NH_3 PRODUCT	SLURRY TO TAILINGS POND	MAKE-UP WATER
STREAM NO.		16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
CH_4		9,762.6	0.2	—	9,758.2	—	—	26,775.6	26,743.1	9.7	—	—	—	0.2	—	—	—
CO		13,411.5	2.4	—	13,407.2	—	—	3.4	3.4	—	—	—	—	2.4	—	—	—
CO_2		4,938.1	286.7	—	4,894.4	455.1	—	1,321.2	144.4	920.0	—	81.9	—	286.7	—	—	—
H_2		55,835.5	3.3	—	55,822.4	—	—	1,342.6	1,341.8	0.6	—	—	—	3.3	—	—	—
N_2		638.0	—	—	638.0	1,081.0	—	638.0	637.8	0.2	—	—	—	—	—	—	—
NH_3		(TRACE)	476.5	—	—	—	—	—	—	—	—	—	—	1.0	495.7	—	—
H_2S		34.6	—	—	(4 PPM)	—	—	—	—	—	—	336.1	—	—	—	—	—
SO_2		—	—	—	—	8.4	—	—	—	—	—	—	—	—	—	—	—
COS		—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
O_2		—	—	—	—	24.8	—	—	—	—	—	—	—	—	—	—	—
H_2O (V)		891.0	55,842.5	—	769.1	538.6	—	776.2	3.1	66.7	—	37.7	—	51.2	—	—	—
S		—	—	—	—	—	—	—	—	—	—	—	362.3	—	—	—	—
TOTAL MOLES/HOUR		85,511.3	56,611.6	—	85,289.3	2,107.9	—	30,837.0	28,873.6	997.2	—	455.7	362.3	344.8	495.7	—	—
MOLECULAR WEIGHT		10.49	18.14	—	10.46	29.10	—	16.91	15.80	42.0	—	34.54	32.06	39.54	17.034	—	—
TOTAL LBS/HR (VAPOR)		897,275	1,026,872	—	891,760	61,348	—	521,423	455,496	41,883	—	15,738	11,617	13,634	8,444	—	—
MAF COAL	LBS/HR	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
MAF CHAR	LBS/HR	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
H_2O (L)	LBS/HR	—	—	133,270	—	—	164,057	—	—	—	243,274	—	—	—	—	819,785	1,854,370
ASH	LBS/HR																
MgO	LBS/HR	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
$MgCO_3$	LBS/HR	—	—	—	—	—	—	—	—	—	21,613	—	—	—	—	21,613	—
CaO	LBS/HR	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
$CaCO_3$	LBS/HR	—	—	—	—	—	—	—	—	—	57,107	—	—	—	—	57,107	—
$Ca(OH)_2$	LBS/HR	—	—	—	—	—	1,394	—	—	—	—	—	—	—	—	3,194	—
CaS	LBS/HR	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	—
OTHERS (3)	LBS/HR	—	—	1,484	—	—	14,004	—	—	—	64,237	—	—	—	—	96,335	—
TOTAL LBS/HR	(SOLIDS)	—	—	134,754	—	—	179,455	—	—	—	386,231	—	—	—	—	998,034	—
TOTAL LBS/HR		897,275	1,026,872	134,754	891,760	61,348	179,455	521,423	455,496	41,883	386,231	15,738	11,617	13,634	8,444	998,034	1,854,370
OPER. TEMP., °F		105	109	239	300	300	192	150	105	100	200	150	—	150	120	146	AMB
OPER. PRESS., PSIG		107.8	85	15	219	0	—	105.7	1,000	10.2	—	10.0	—	10	270	—	—
TONS/DAY		—	—	—	—	—	—	—	—	—	—	—	139.4	—	101.3	—	—
MM SCFD		—	—	—	—	—	—	—	262.63	—	—	—	—	—	—	—	—
MM GAL/DAY		—	—	—	—	—	—	—	—	—	—	—	—	—	—	2.814	5.341
ACRE FT./YR.		—	—	—	—	—	—	—	—	—	—	—	—	—	—	—	5,383

NOTES:

1. NO CLOSED MATERIAL BALANCE CAN BE MADE AROUND ANY ONE PLANT AREA BLOCK ON THIS P.F.D. SINCE ONLY PRIMARY FLOWS ARE SHOWN.
2. REFER TO DRAWING 00-1-05 FOR PLANT WATER WATER BALANCE FLOWS.
(3) "OTHERS" IN TABULATIONS MAY INCLUDE BOTH INERT AND REACTIVE COMPONENTS NOT SHOWN. SEE AREA P.F.D'S. FOR FURTHER IDENTIFICATION OF "OTHERS".

NO	REVISIONS	DATE	BY	CHK'D	APP'D
1	REVISED FOR FINAL ESTIMATE	7-25-77	OFF		
2	UPDATED PER FINAL REVIEW	8-11-77	JLH		
3	REVISED PER CUSTOMER COMMENTS	9-23-77	OFF		

REFERENCE DRAWINGS

PRINT RECORD

DATE	7/15/77	7/18/77	7-28-77	8-15-77	9-27-77
FOR	P	FA	A	R	R
REVISED	0	0	1	2	3
CUSTOMER	—	—	—	—	—
FIELD	—	—	—	—	—
INTRA CO.	4	4	3	3	2

ENG. RECORD

DRAWN	gc	1/6/77
CHECKED		9/15/77
MECH CK		
STRUCT CK		
ELECT CK		
PIPING CK	OFF	7-18-77
INSTR CK		
CIVIL CK		
PROCESS		7-18-77
CUSTOMER APPROVAL	GJLT	7/28/77

DRAWING STATUS

ISSUED	DATE
PRELIMINARY IN HOUSE REVIEW	7-15-77
FOR COMMENTS AND/OR APPROVAL	7-18-77
APPROVED FOR ESTIMATE	7-25-77
REVISED & APPROVED FOR ESTIMATE 3	9-23-77

NOT APPROVED FOR CONSTRUCTION UNLESS SIGNED & DATED. DESTROY ALL PRINTS BEARING EARLIER DATE &/OR REVISION NO.

NORTH DAKOTA LIGNITE PIPELINE GAS

AREA 00 BLOCK FLOW DIAGRAM

OVERALL PLANT

PRIMARY FLOWS

CO_2 ACCEPTOR PROCESS COMMERCIAL DESIGN

CONOCO COAL DEVELOPMENT COMPANY

RESEARCH DIVISION LIBRARY PENNSYLVANIA

SCALE

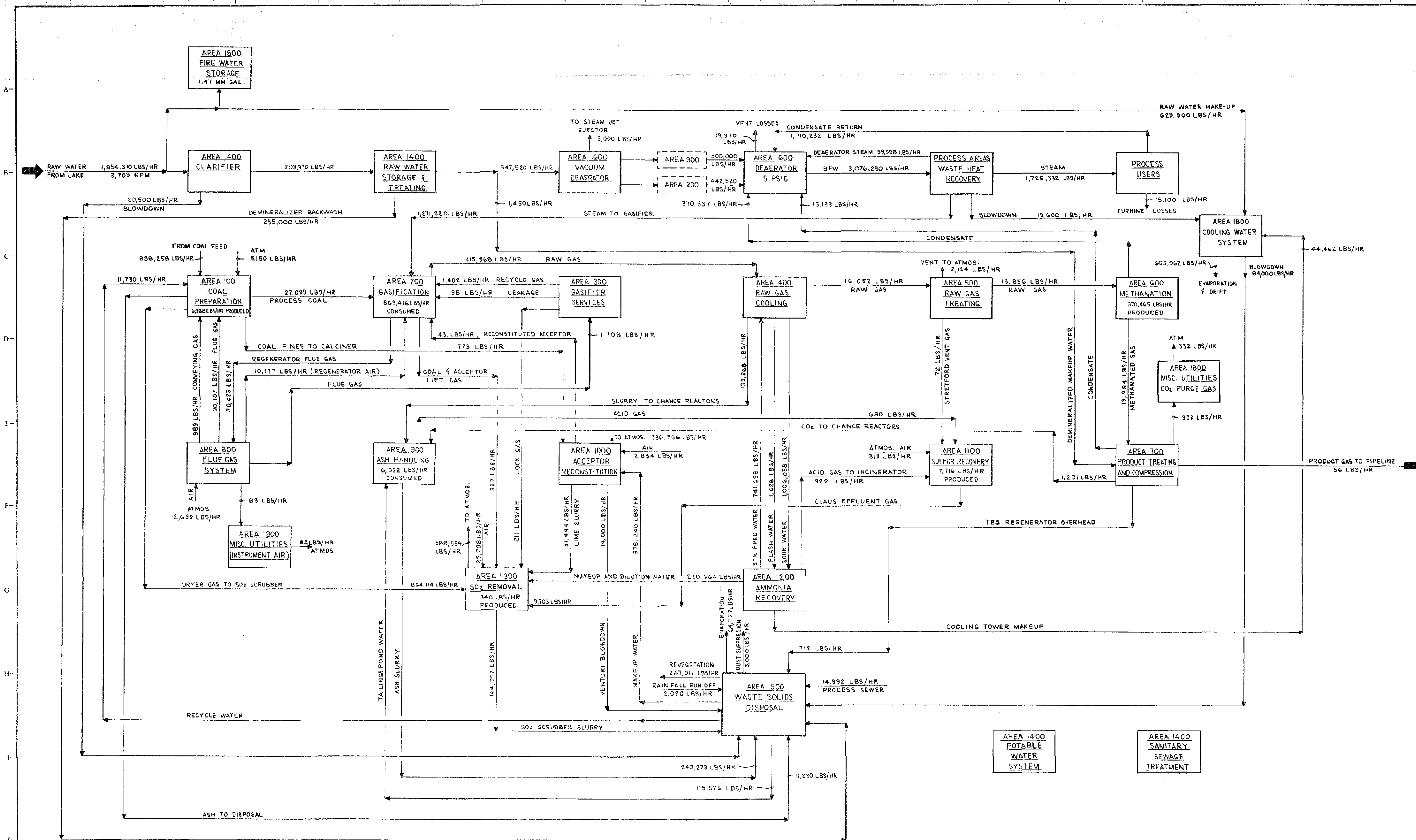
Stearns-Roger INCORPORATED

DWG NO. 22846

SHEET NO. 00-1-02

ORDER NO. C-18575

REV 3



NO.	REVISIONS	DATE	BY	CHK'D	APP'D
1	UPDATED DESIGN	7-20-77	S.R.		
2	UPDATED PER FINAL REVIEW	8-11-77	JLR		

REFERENCE DRAWINGS

PRINT RECORD

ENG. RECORD

DRAWING STATUS

NORTH DAKOTA LIGNITE PIPELINE GAS

AREA 00 BLOCK FLOW DIAGRAM

WATER BLOCK FLOW DIAGRAM

CO_2 ACCEPTOR PROCESS - COMMERCIAL DESIGN

CONOCO COAL DEVELOPMENT COMPANY
RESEARCH DIVISION LIBRARY PENNSYLVANIA

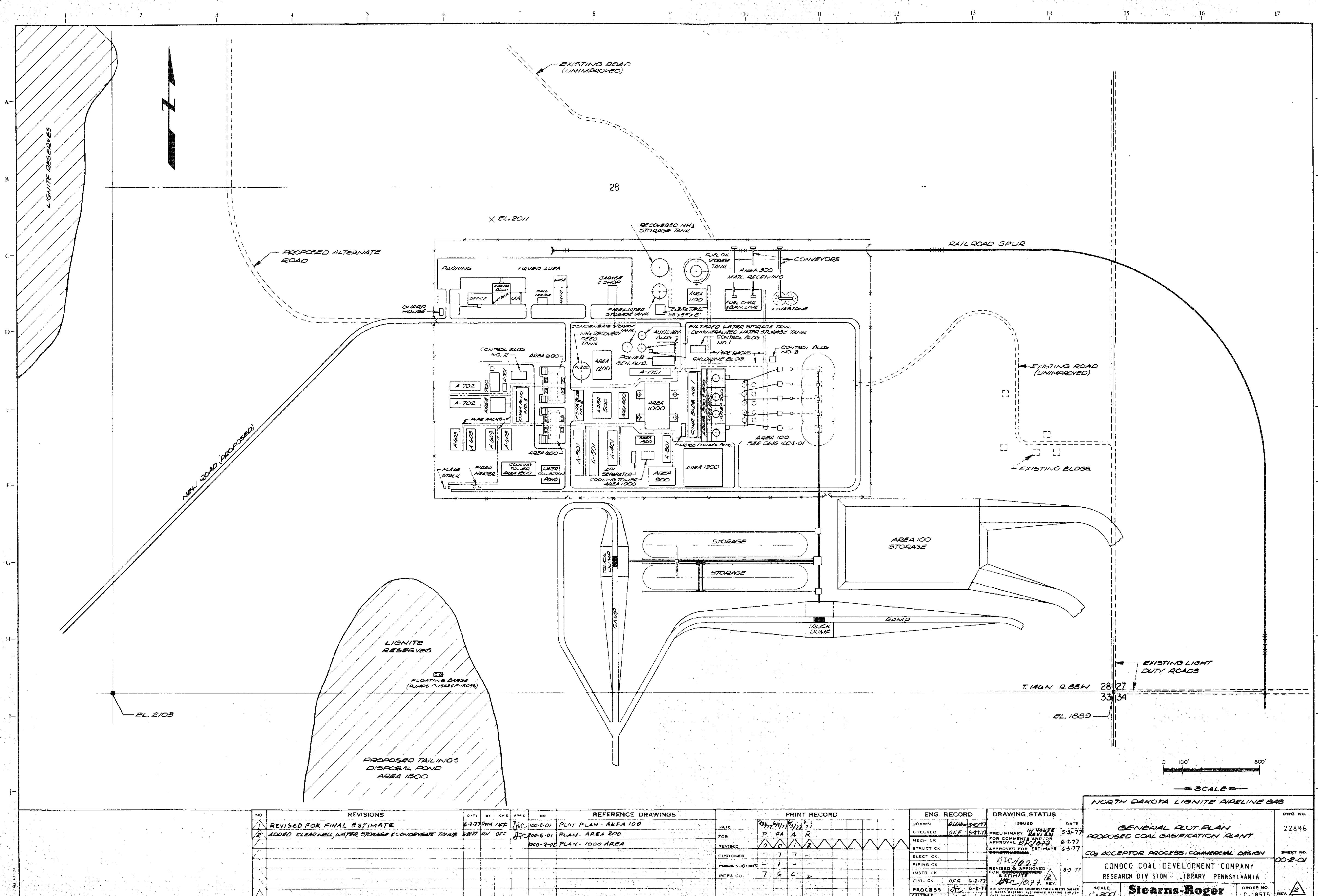
Stearns-Roger INCORPORATED

DWG. NO. 22846

SHEET NO. 00-1-05

ORDER NO. C 18575

REV. 2


EXISTING ROAD (UNIMPROVED)
LIGNITE RESERVES
PROPOSED ALTERNATE ROAD
X EL. 2011
28
RECOVERED NH3 STORAGE TANK
RAILROAD SPUR
FUEL OIL STORAGE TANK
CONVEYORS
AREA 300 MATL. RECEIVING
PARKING
PAVED AREA
GARAGE & SHOP
AREA 1100
FUEL CHAR & BURN LIME
LIMESTONE
GUARD HOUSE
FIREWATER STORAGE TANK
CLEARWELL 55'x55'x8'
CONDENSATE STORAGE TANK
NH3 RECOVERY FEED TANK
AUXILIARY BLDG.
FILTERED WATER STORAGE TANK
DEMINERALIZED WATER STORAGE TANK
CONTROL BLDG. NO. 1
CONTROL BLDG. NO. 3
CONTROL BLDG. NO. 2
AREA 600
POWER GEN. BLDG.
AREA 1200
A-1701
PIPE RACKS
CHLORINE BLDG.
AREA 1000
AREA 500
AREA 100 SEE DWG 100-2-01
MOTOR CONTROL BLDG.
EXISTING ROAD (UNIMPROVED)
EXISTING BLDGS.
FLARE STACK
FIRED HEATER
COOLING TOWER AREA 1500
WATER COLLECTION POND
API SEPARATOR
COOLING TOWER AREA 1000
AREA 900
AREA 1300
NEW ROAD (PROPOSED)
STORAGE
TRUCK DUMP
AREA 100 STORAGE
RAMP
LIGNITE RESERVES
FLOATING BARGE (PUMPS P-1503 & P-1503S)
EXISTING LIGHT DUTY ROADS
T. 146N R. 88W
EL. 1889
EL. 2103
PROPOSED TAILINGS DISPOSAL POND AREA 1500
SCALE
NORTH DAKOTA LIGNITE PIPELINE GAS
GENERAL PLOT PLAN
PROPOSED COAL GASIFICATION PLANT
CO2 ACCEPTOR PROCESS - COMMERCIAL DESIGN
CONOCO COAL DEVELOPMENT COMPANY
RESEARCH DIVISION - LIBRARY PENNSYLVANIA
Stearns-Roger INCORPORATED
DWG. NO. 22846
SHEET NO. 00-2-01
ORDER NO. C-18575
SCALE 1" = 200'
REVISIONS
REVISED FOR FINAL ESTIMATE
ADDED CLEARWELL, WATER STORAGE & CONDENSATE TANKS
REFERENCE DRAWINGS
100-2-01 PLOT PLAN - AREA 100
200-6-01 PLAN - AREA 200
1000-2-02 PLAN - 1000 AREA
PRINT RECORD
ENG. RECORD
DRAWING STATUS
NOT APPROVED FOR CONSTRUCTION UNLESS SIGNED & DATED DESTROY ALL PRINTS BEARING EARLIER DATE &/OR REVISION NO.

PLANT INVESTMENT

The total plant investment for the CO_2 Acceptor Process Commercial Plant producing pipeline gas at Beulah, North Dakota, is estimated to be $717,140,000 in mid-1977 dollars. The breakdown of this cost into the various areas is found in Table (6-1).

This estimate is of a preliminary nature and was developed with the aid of the ICARUS COST System. Approximately eighty percent of the equipment pricing was developed by Stearns-Roger, either in conjunction with vendors or by in-house methods. Detailed take-offs were made for the costs and erection of the main gasification structure and all major piping included therein. There is about $80,000,000 included in the estimate for the proprietary-type units and packages, among which are the Stretford Units, Phosam-W Units, UOP Sulfur Removal System, various water-treatment packages and the flare.

The labor productivity for the North Dakota area was assumed to be 100% per the Braun Guidelines (10). It was assumed that an adequate local labor pool exists and that no camp would be required. Current applicable union wage rates were used for all crafts except finishers and teamsters; these unions had not reached a new agreement when the estimate was prepared and an average escalation was assumed. The average direct craft rate is $10.16 per hour. All labor is based on a forty-hour week with casual overtime only.

Subsistence and travel pay, in addition to normal fringe benefits, have been included where appropriate.

Indirect Field Costs were calculated at 125% of Direct Field Labor Cost. A sales and use tax of 3.0% was included and casual premium pay of 3% of Direct Field Labor was added. To this subtotal, engineering costs and contractor fees were added at 11%, based on the method used to calculate coal gasification plant investment by C F Braun and Company.

Contingency was added to the Total Field and Engineering Cost (including fees, sales tax and premium pay) at 20% for the process areas (100 through 1300) and at 25% for the utility and offsite areas (1400 through 2000). Twenty percent contingency was used in the process areas to provide for process uncertainties. Twenty-five percent was used in the utility and offsite areas because they are not as well defined as the process areas. The accuracy range of the estimate is ± 20%.

The cost of spare parts, catalysts and chemicals, license fees, start-up, interest during construction, working capital, client overhead and administrative costs are not included in the base estimate but they are included in "Plant Economics".

A preliminary cash flow curve and master-plan schedule are included based on an anticipated project duration of five years.

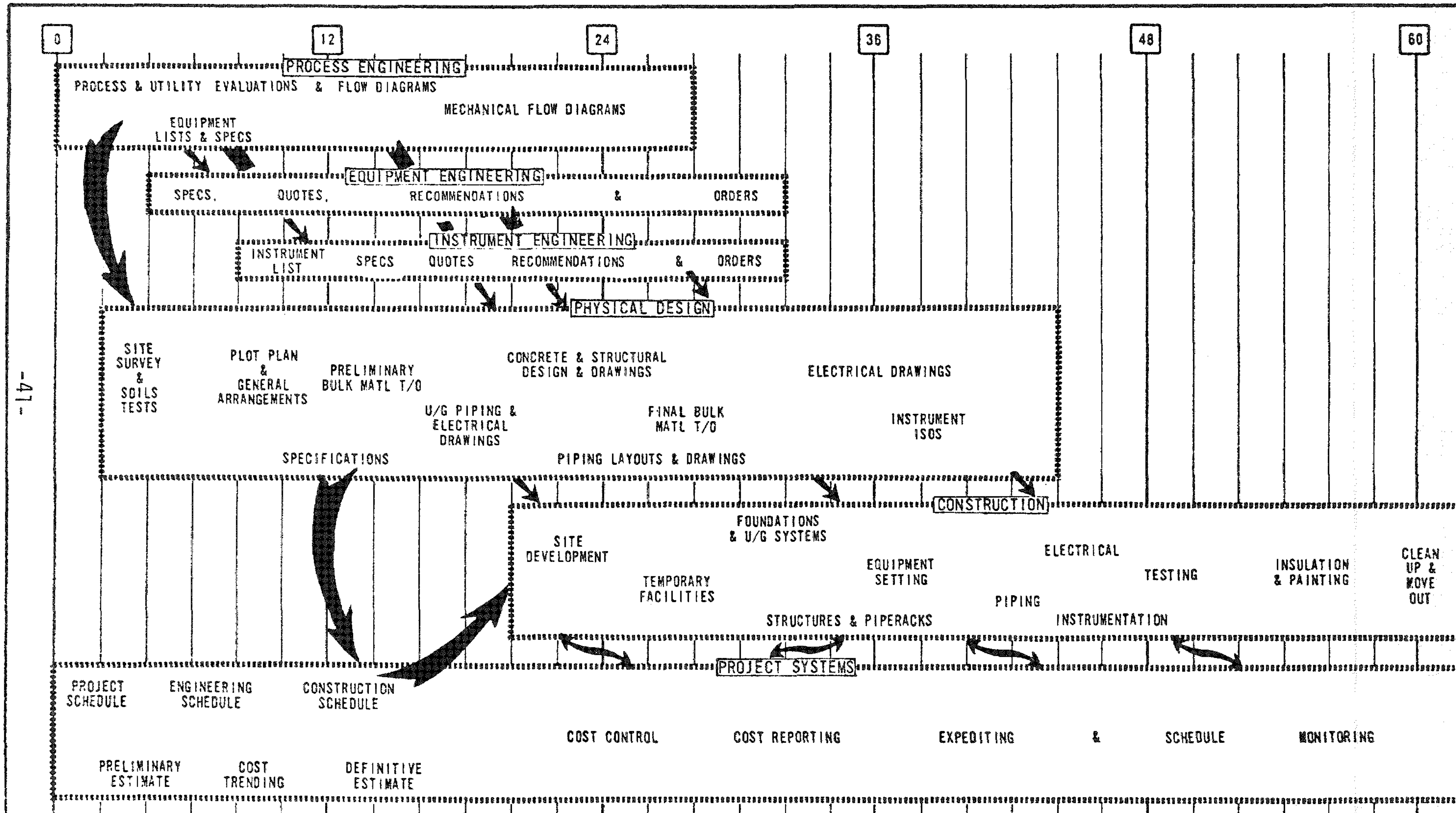
TABLE 6-1

PLANT INVESTMENT

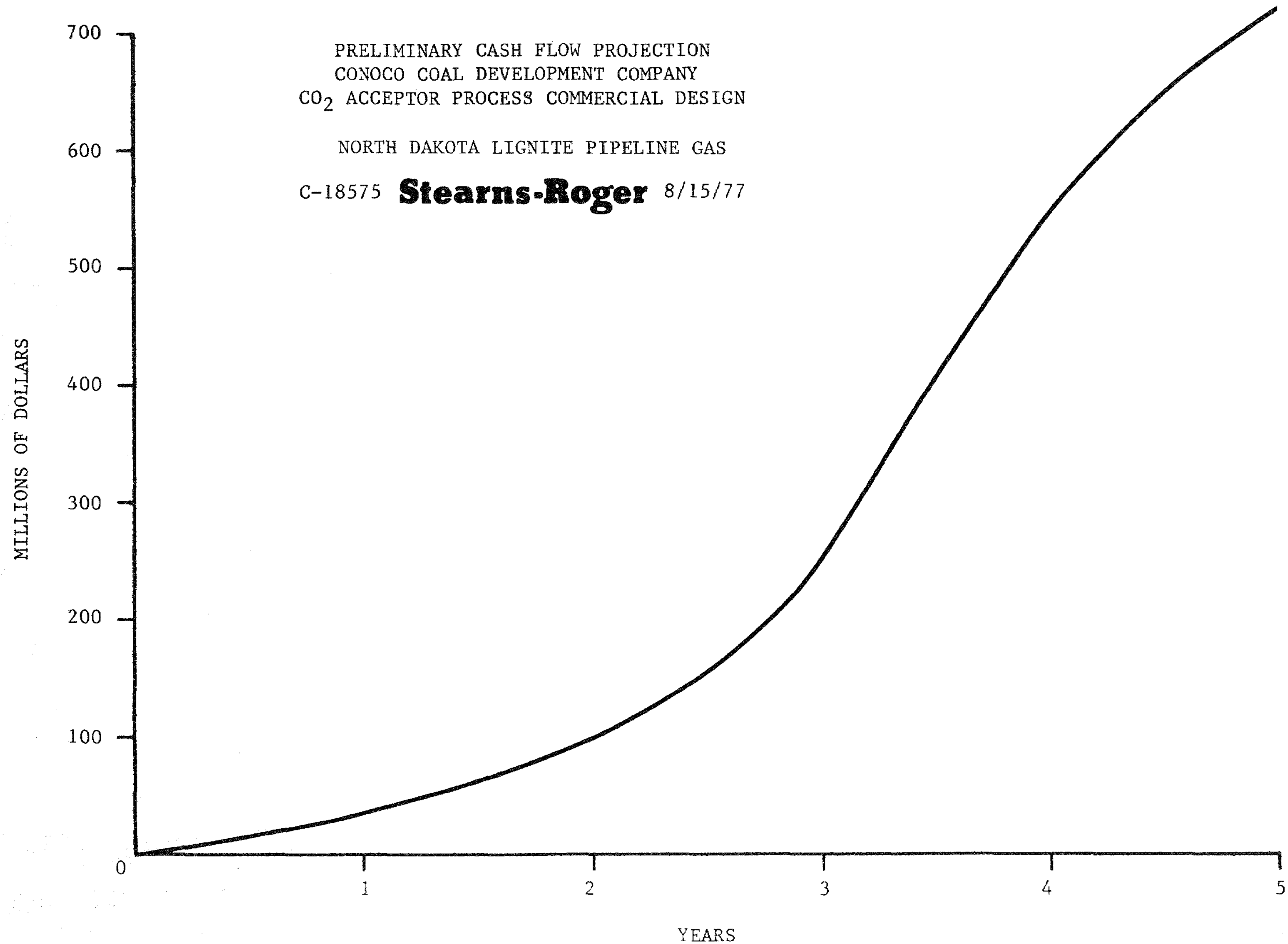
(In Millions of Dollars)

NORTH DAKOTA LIGNITE PIPELINE GAS PLANT

AREA NO.	ITEM	AREA COST
100	Coal Preparation	92.53
200	Gasification	139.26
300	Gasifier Services	15.62
400	Raw Gas Cooling	40.76
500	Raw Gas Treating	26.35
600	Methanation	47.95
700	Product Treating and Compression	25.60
800	Flue Gas System	114.68
900	Ash Handling	7.13
1000	Acceptor Reconstitution	41.05
1100	Sulfur Recovery	3.96
1200	Ammonia Recovery	18.31
1300	SO_2 Removal	44.18
1400	Raw Water Supply and Treating	9.21
1500	Waste Solids Disposal	1.11
1600	Steam System	12.36
1700	Electrical	18.41
1800	Miscellaneous Utilities	15.44
1900	Offsites	4.03
2000	General Facilities	39.20
	TOTAL PLANT INVESTMENT	717.14



CONOCO COAL DEVELOPMENT CO. N.D. LIGNITE PIPELINE GAS	**Stearns-Roger** AUGUST 15, 1977 PROPOSAL NO. C-18575	PRELIMINARY MASTER PLAN SCHEDULE CO_2 ACCEPTOR PROCESS COMMERCIAL DESIGN


PRELIMINARY CASH FLOW PROJECTION
CONOCO COAL DEVELOPMENT COMPANY
CO_2 ACCEPTOR PROCESS COMMERCIAL DESIGN
NORTH DAKOTA LIGNITE PIPELINE GAS
C-18575 Stearns-Roger 8/15/77
MILLIONS OF DOLLARS
700
600
500
400
300
200
100
0
1
2
3
4
5
YEARS

PLANT ECONOMICS

INTRODUCTION

The summary of capital and operating costs as well as gas cost for the CO_2 Acceptor plant is shown in table 7-1. In addition, there are tables in this section showing capital requirements, annual operating costs, and annual maintenance costs. Basis of economic data and evaluation was the C F Braun & Company Guidelines report (10).

When comparing various gasification processes, gas cost is an important consideration. Cost of gas depends upon many factors, e.g., investment amortization, taxes, operating costs, etc. This section of the report discusses those factors.

FINANCING METHOD

There are two potential methods of financing the gasification plants. These are the utility financing and private investor financing methods. Because the gas costs are of interest to various groups, the costs are presented using both methods.

The following parameters apply to the two financing methods:

	Utility	Private
Project life, years	20	20
Debt - equity ratio	75/25	100% Equity
Return on equity, %	15	12 DCF
Federal Income Tax, %	48	48

Depreciation for the utility method is 20-year straight line depreciation on plant investment, allowance for funds used during construction, and capitalized portion of start-up costs. Depreciation for the private investor method is 16-year sum of the years digits depreciation on total plant investment.

CAPITAL REQUIREMENT

Capital requirements and calculation methods are shown in the table 7-2. Cost of initial charge of catalyst and chemicals reported in the table was calculated based on the initial fill volumes required in the various process areas.

PLANT ECONOMICS - continued

ANNUAL OPERATING COSTS

Annual operating costs are shown in the table 7-3. Process operating manpower was estimated in accordance with complexity and number of units being operated. Maintenance labor is taken as 50% of total maintenance cost. On-stream factor is 0.9. All costs, wages, and product values are mid 1977 values.

MAINTENANCE COSTS

Annual maintenance costs shown in the table 7-4 are calculated as a percent of installed unit cost. Maintenance factors are as shown in the table.

CATALYST AND CHEMICALS

Annual catalyst replacement and chemical make-up costs are defined in annual Catalyst and Chemical Consumption Costs, Table 7-5.

GAS COST

A char showing the variation of gas cost with coal cost is included in this section.

TABLE 7-1

SUMMARY OF CAPITAL AND OPERATING COSTS

(In Millions of Dollars)

NORTH DAKOTA LIGNITE PIPELINE GAS PLANT

	UTILITY FINANCING		PRIVATE FINANCING
	AVERAGE GAS COST	FIRST YEAR GAS COST	CONSTANT GAS COST
Capital Costs			
Total Plant Investment	$717.14	$717.14	$717.14
Initial Charge of Catalysts and Chemicals	4.80	4.80	4.80
Allowance for Funds Used During Construction	121.02	121.02	121.02
Paid-Up Royalties	1.21	1.21	1.21
Start-Up Costs	24.60	24.60	24.60
Working Capital	18.29	20.68	21.61
TOTAL CAPITAL REQUIREMENT	887.06	889.45	890.38
Operating Costs			
Raw Materials	45.09	45.09	45.09
Catalysts and Chemicals	4.67	4.67	4.67
Labor			
Process Operating Labor	2.97	2.97	2.97
Maintenance Labor	16.20	16.20	16.20
Supervision	3.83	3.83	3.83
Administration and General Overhead	13.80	13.80	13.80
Supplies			
Operating	0.89	0.89	0.89
Maintenance	16.20	16.20	16.20
Local Taxes and Insurance	19.36	19.36	19.36
TOTAL GROSS OPERATING COSTS/YEAR	$123.01	$123.01	$123.01
TOTAL BY-PRODUCT CREDITS	5.68	5.68	5.68
TOTAL NET OPERATING COSTS/YEAR	$117.33	$117.33	$117.33
AVERAGE GAS COST, $/MM Btu	2.73		
FIRST YEAR GAS COST, $/MM Btu		3.43	
CONSTANT GAS COST, $/MM Btu			3.70

TABLE 7-2

CAPITAL REQUIREMENTS

(In Millions of Dollars)

NORTH DAKOTA LIGNITE PIPELINE GAS PLANT

		UTILITY FINANCING		PRIVATE FINANCING
		AVERAGE GAS COST	FIRST YEAR GAS COST	CONSTANT GAS COST
Total Plant Investment		$ 717.14	$ 717.14	$ 717.14
Initital Charge of Catalysts and Chemicals				
Catalysts	$3.05			
Chemicals	1.75			
		4.80	4.80	4.80
Paid-Up Royalties		1.21	1.21	1.21
Allowance for Funds Used During Construction (Total plant investment x average spending period in years x 9%)		121.02	121.02	121.02
Startup Costs (20% of total annual gross operating costs)		24.60	24.60	24.60
Working Capital (14-day inventory of raw materials + materials and supplies at 0.9% of total plant invest ment + net receivables at 1/24 annual gas and by-product revenue at calculated sales price)		18.29	20.68	21.61
TOTAL CAPITAL REQUIREMENT		$887.06	$889.45	$890.38

TABLE 7-5

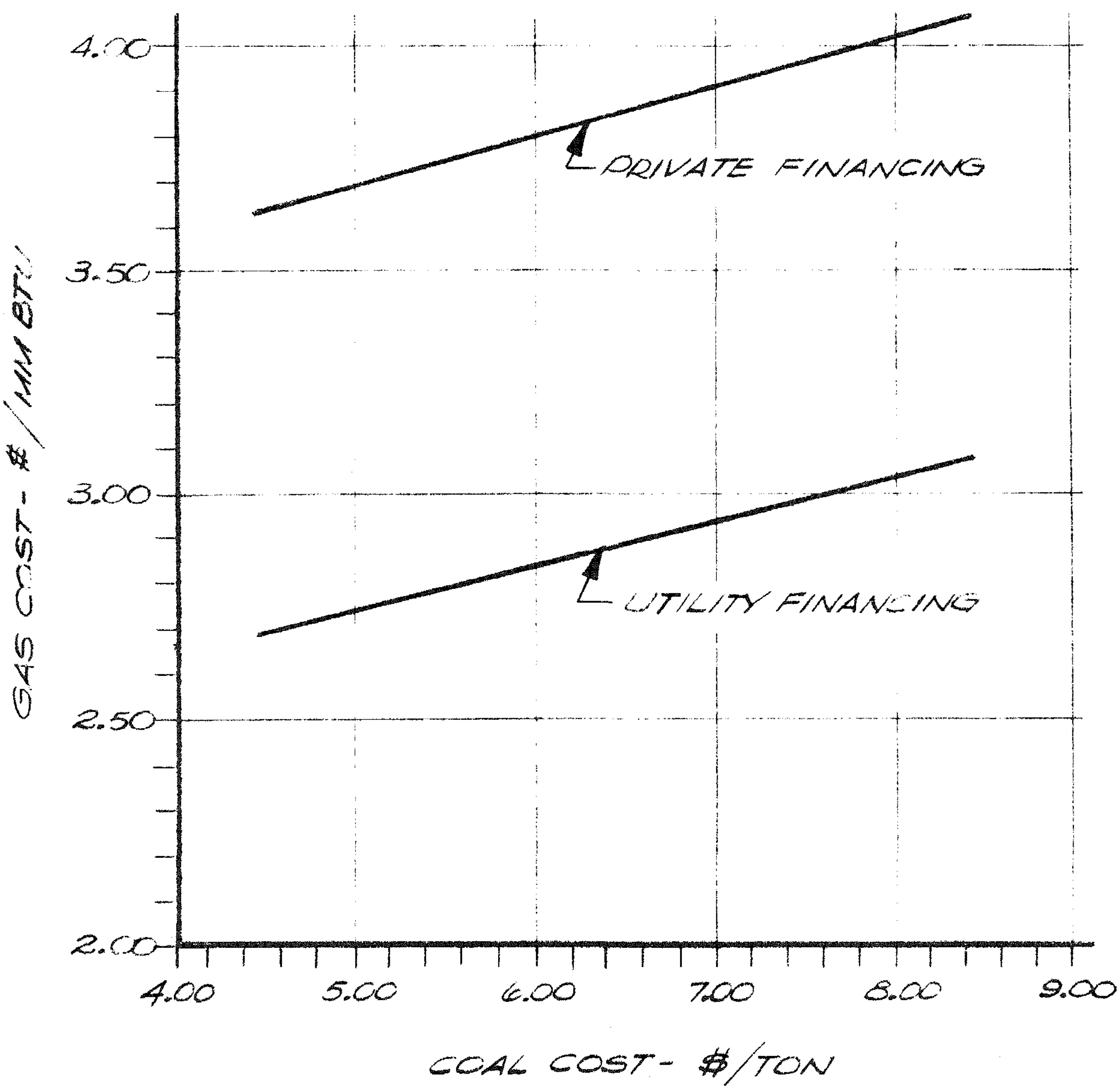
CHEMICALS AND CATALYSTS

NORTH DAKOTA LIGNITE PIPELINE GAS PLANT

Catalysts Area Number	Designation	
600	Methanation	$3,033,200
1100	Sulfur Recovery	3,760
	Subtotal Catalysts	$3,036,960

Chemicals Area Number		
500	Raw Gas Treating (Stretford)	$ 56,450
700	Benfield Solution	11,000
	TEG	31,000
1200	Ammonia Recovery (PHOSAM-W)	143,000
1400	Raw Water Supply and Treating	1,391,150
	Subtotal Chemicals	$1,632,600
	TOTAL CATALYSTS AND CHEMICALS	$4,669,560

GAS COST VS. COAL COST
NORTH DAKOTA LIGNITE PIPELINE GAS



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6. Interim Report No. 4, Pipeline Gas from Lignite Gasification - Current Commercial Economics, January, 1970, Research and Development Report No. 16. National Technical Information Service PB-235856/AS.

7. Interim Report No. 5, Phase III and Phase IV-a, Design and Construction of the Consolidation Synthetic Gas Pilot Plant, Rapid City, South Dakota, March, 1966 to January, 1972. Research and Development Report No. 16. May, 1974. National Technical Information Service PB-236504/AS. (Microfiche)

8. Interim Report No. 1, Appendices A & B, CSG Pilot Plant Operations, January, 1972 - June, 1973. Submitted to the Office of Coal Research/ American Gas Association, August, 1973.

9. Interim Report No. 2, CSG Pilot Plant Operations, July, 1973 - July, 1974, Submitted to United States Energy Research and Development Administration May 12, 1975.

10. FE-1235-1, Coal Gasification Commercial Concepts - Gas Cost Guidelines, January, 1976. Prepared for United States Energy Research and Development Administration and the American Gas Association by C F Braun & Company.

APPENDIX 1

LURGI PLANT

The design and cost estimate used in this study are based upon published data for a Lurgi plant designed to gasify North Dakota lignite. American Natural Gas Company, through their subsidiary, ANG Coal Gasification Company, has proposed to construct such a coal gasification plant in Mercer County, North Dakota. The data used in this study was obtained from their application to the Federal Power Commission for a certificate to build the plant.*

The gasification plant will utilize the Lurgi Process. The plant will gasify a North Dakota lignite mined from the Beulah-Hazen area. It will produce approximately 275 MM SCF per stream day of gas having a heating value of 979 Btu per standard cubic foot.

In the process, coal is first crushed to the desired size and then introduced into gasifier vessels. In the gasifiers, the coal is dehydrated, devolitized and gasified by reacting with oxygen and steam. The raw gas which is produced is cooled and scrubbed to remove tar, dust, heavy oils and gas liquor. Subsequent processing steps include by-product recovery, gas purification, CO shift and methanation.

The published information includes data for plant investment, operating costs, operating manpower, and utilities. These data and the design information were adjusted by appropriate factors to convert the ANG data to the CCDC basis: for example,

1. The instantaneous pipeline gas production from the ANG plant is equivalent to approximately 269 billion Btu per day where as the CCDC rate was set at 250 billion Btu per day. Material balances were adjusted accordingly and the ANG estimates for all areas directly related to SNG gas rate were adjusted by an appropriate factor.

2. Although the moisture and ash free analyses and heating value for the ANG and CCDC coals are very nearly the same, (HHV = 12,010 vs 12,057 respectively), there is a significant difference in ash content (11.59 for ANG vs 7.58 for CCDC). Estimates for areas affected by ash quantities were adjusted.

By making these and other adjustments based upon capacities and small differences in coal analyses, coal feed and product rates were established for a Lurgi plant processing Renner's Cove lignite to produce 250 billion Btu per day of pipeline gas. These rates are shown on a block flow diagram of the plant, drawing No. 00-1-08. A plant energy balance was also developed. This is shown in Table 9-6.

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* Application for Certificate of Public Convenience, Docket No. CP-75-278.

APPENDIX 1 - LURGI PLANT - continued

The cost estimate for the Lurgi Plant was developed by factoring the ANG estimate on an area-by-area basis, applying the proper capacity factor and cost exponent.

Costs presented in the ANG estimate for land, location bonus and construction housing were deleted so that plant cost estimate is comparable to the CO_2 Acceptor plant estimate. Exactly what is covered by costs listed in the ANG estimate as taxes, miscellaneous, and contingency are unknown. Therefore, these costs were subtracted from the ANG estimate and similar costs developed for the CO_2 Acceptor plant added. The ANG cost figure presented in the 1975 FPC application were given in September 1974 dollars. The estimates for this study were therefore escalated to a mid-1977 basis.

Tables are presented showing Plant Investment, Summary of Capital and Operating Cost, Capital Requirement, Annual Operating Costs, Annual Maintenance Costs, and the Summary Energy Balance. Maintenance costs were calculated in the same manner as for the CO_2 Acceptor plants. The Lurgi plant as designed for ANG requires an external source of electrical power. Power costs were calculated from industrial rates published for North Dakota.

TABLE 9-1

PLANT INVESTMENT

(In Millions of Dollars)

LURGI PIPELINE GAS PLANT

AREA NO.	ITEM	AREA COST
1100	Gasification	128.4
1200	Shift Conversion	18.5
1300	Gas Cooling	17.8
1400 1500	Rectisol	95.6
1600	Phenosolvan	49.9
1700	Methanation	50.5
1800	Gas-Liquid Separation	16.1
1900	Gas Compression	14.3
2000	Coal Preparation and Handling	57.5
3000	Oxygen	85.8
4000 4200	Sulfur Recovery	31.1
5000	Steam Generation and Distribution	114.7
5200	Stack Gas Scrubbing	25.4
5300	Power Distribution	11.5
5400	Raw Water Intake and Placement Water Treatment and Distribution	11.4
5500	Cooling Water	16.2
5600	Fire Water	1.1
5700	Miscellaneous Utilities	1.6
6000	Offsite Storage	1.8
7000	Interconnecting Pipe	13.1
8100	Liquid Waste Effluent	7.8
8200	Ash Disposal and Sanitary	2.4
8300	Flare	3.7
	Site Preparation, Buildings, & Vehicles	38.0
	TOTAL PLANT INVESTMENT	814.2

TABLE 9-2

SUMMARY OF CAPITAL AND OPERATING COSTS

(In Millions of Dollars)

LURGI PIPELINE GAS PLANT

	UTILITY FINANCING AVERAGE GAS COST	UTILITY FINANCING FIRST YEAR GAS COST	PRIVATE FINANCING CONSTANT GAS COST
Capital Costs			
Total Plant Investment	$814.20	$814.20	$814.20
Initial Charge of Catalysts and Chemicals	7.81	7.81	7.81
Allowance for Funds Used During Construction	137.40	137.40	137.40
Paid-Up Royalties	0.97	0.97	0.97
Start-Up Costs	26.68	26.68	26.68
Working Capital	20.27	23.01	24.07
TOTAL CAPITAL REQUIREMENT	1007.33	1010.07	1011.13
Operating Costs			
Raw Materials	55.22	55.22	55.22
Catalysts and Chemicals	3.91	3.91	3.91
Power Costs	5.29	5.29	5.29
Labor			
Process Operating Labor	3.33	3.33	3.33
Maintenance Labor	13.63	13.63	13.63
Supervision	3.28	3.28	3.28
Administration and General Overhead	12.14	12.14	12.14
Supplies			
Operating	1.00	1.00	1.00
Maintenance	13.63	13.63	13.63
Local Taxes and Insurance	21.98	21.98	21.98
TOTAL GROSS OPERATING COSTS/YEAR	$133.41	$133.41	$133.41
TOTAL BY-PRODUCT CREDITS	35.73	35.73	35.73
TOTAL NET OPERATING COSTS/YEAR	$ 97.68	$ 97.68	$ 97.68
AVERAGE GAS COST, $/MM Btu	2.67		
FIRST YEAR GAS COST, $/MM Btu		3.46	
CONSTANT GAS COST, $/MM Btu			3.77

TABLE 9-3

CAPITAL REQUIREMENTS

(In Millions of Dollars)

LURGI PIPELINE GAS PLANT

	UTILITY FINANCIAL		PRIVATE FINANCING
	AVERAGE GAS COST	FIRST YEAR GAS COST	CONSTANT GAS COST
Total Plant Investment	814.20	814.20	814.20
Initial Charge of Catalyst and Chemicals	7.81	7.81	7.81
Paid-Up Royalties	0.97	0.97	0.97
Allowance for Funds Used During Construction (Total plant invesment x average spending period in years x 9%)	137.40	137.40	137.40
Startup Costs (20% of total annual gross operating costs)	26.68	26.68	26.68
Working Capital (14-day inventory of raw materials + materials and supplies at 0.9% of total plant investment + net receivables at 1/24 annual gas and by-product revenue at calculated sales price)	20.27	23.01	24.07
TOTAL CAPITAL REQUIREMENT	1007.33	1010.07	1011.13

TABLE 9-4

ANNUAL OPERATING COSTS

(In Millions of Dollars)

LURGI PIPELINE GAS PLANT

Raw Materials		
Coal (33,617 tons/day @ \$5.00/ton)		\$ 55.22
Catalysts and Chemicals		\$ 3.91
Power Cost		
Energy (1.25 x 10^6 kwh/day @ \$0.01262/kwh)	\$ 5.18	
Demand (52,080 kw @ \$2.09/kw)	.11	
Total Power Cost		\$ 5.29
Labor Cost		
Process Operating Labor (55 men/shift @ \$7.30/hr)	\$ 3.33	
Maintenance (50% of total maintenance)	13.63	
Supervision (20% of direct labor)	3.28	
Total Labor Cost		\$ 20.24
Administrative and General Overhead		
(60% of labor)		\$ 12.14
Supplies		
Operating (30% of process operating labor)	\$ 1.00	
Maintenance (50% of total maintenance)	13.63	
Total Supplies		\$ 14.63
Local Taxes and Insurance		
(2.7% of total plant investment)		\$ 21.98
TOTAL GROSS ANNUAL OPERATING COSTS		\$133.41
By-Product Credits		
Sulfur (161.3 long tons/day @ \$25 long ton)	\$ 1.325	
Ammonia (135 tons/day @ \$140/ton)	6.209	
Naphtha (45,223 gal/day @ \$0.327/gal)	4.858	
Light Oil (134,537 gal/day @ \$0.414/gal)	18.297	
Phenols (27,493 gal/day @ \$0.164/gal)	1.481	
Coal Fines (2,886 tons/day @ \$3.75/ton)	3.555	
Total By-Product Credits		\$ 35.73
TOTAL NET ANNUAL OPERATING COSTS		\$ 97.68

TABLE 9-5

ANNUAL MAINTENANCE COSTS

(In Millions of Dollars)

LURGI PIPELINE GAS PLANT

AREA NO.	ITEM	AREA COST	MAINTENANCE FACTOR PERCENT	ANNUAL MAINTENANCE COST
1100	Gasification	128.4	6	7.70
1200	Shift Conversion	18.5	3	0.56
1300	Gas Cooling	17.8	5	0.89
1400 1500	Rectisol	95.6	3	2.87
1600	Phenosolvan	49.9	3	1.50
1700	Methanation	50.5	3	1.52
1800	Gas-Liquid Separation	16.1	5	0.81
1900	Gas Compression	14.3	3	0.43
2000	Coal Prep & Handling	57.5	6	3.45
3000	Oxygen	85.8	3	2.57
4000 4200	Sulfur Recovery	31.1	3	0.93
5000	Stm. Gen. & Distribution	114.7	1	1.15
5200	Stack Gas Scrubbing	25.4	6	1.52
5300	Power Distribution	11.5	1	0.12
5400	Raw Water Intake & Placement Water Treat & Distribution	11.4	1	0.11
5500	Cooling Water	16.2	1	0.16
5600	Fire Water	1.1	1	0.01
5700	Miscellaneous Utilities	1.6	1	0.02
6000	Offsite Strg.	1.8	1	0.02

TABLE 9-5 ANNUAL MAINTENANCE COSTS - continued

AREA NO.	ITEM	AREA COST	MAINTENANCE FACTOR PERCENT	ANNUAL MAINTENANCE COST
7000	Interconnect Pipe	13.1	1	0.13
8100	Liquid Waste Effluent	7.8	3	0.23
8200	Ash Disposal & Sanitary	2.4	6	0.14
8300	Flare	3.7	1	0.04
	Site Prep, Bldgs, Vehicles	38.0	1	0.38
		814.2		27.26

Annual Maintenance Costs

Maintenance Labor	\$13.63
Maintenance Supplies	\$13.63

$$\text{Average Maintenance Factor} = \frac{27.26}{814.2} = 0.0335$$

GAS COST VS. COAL COST
LURGI PIPELINE GAS

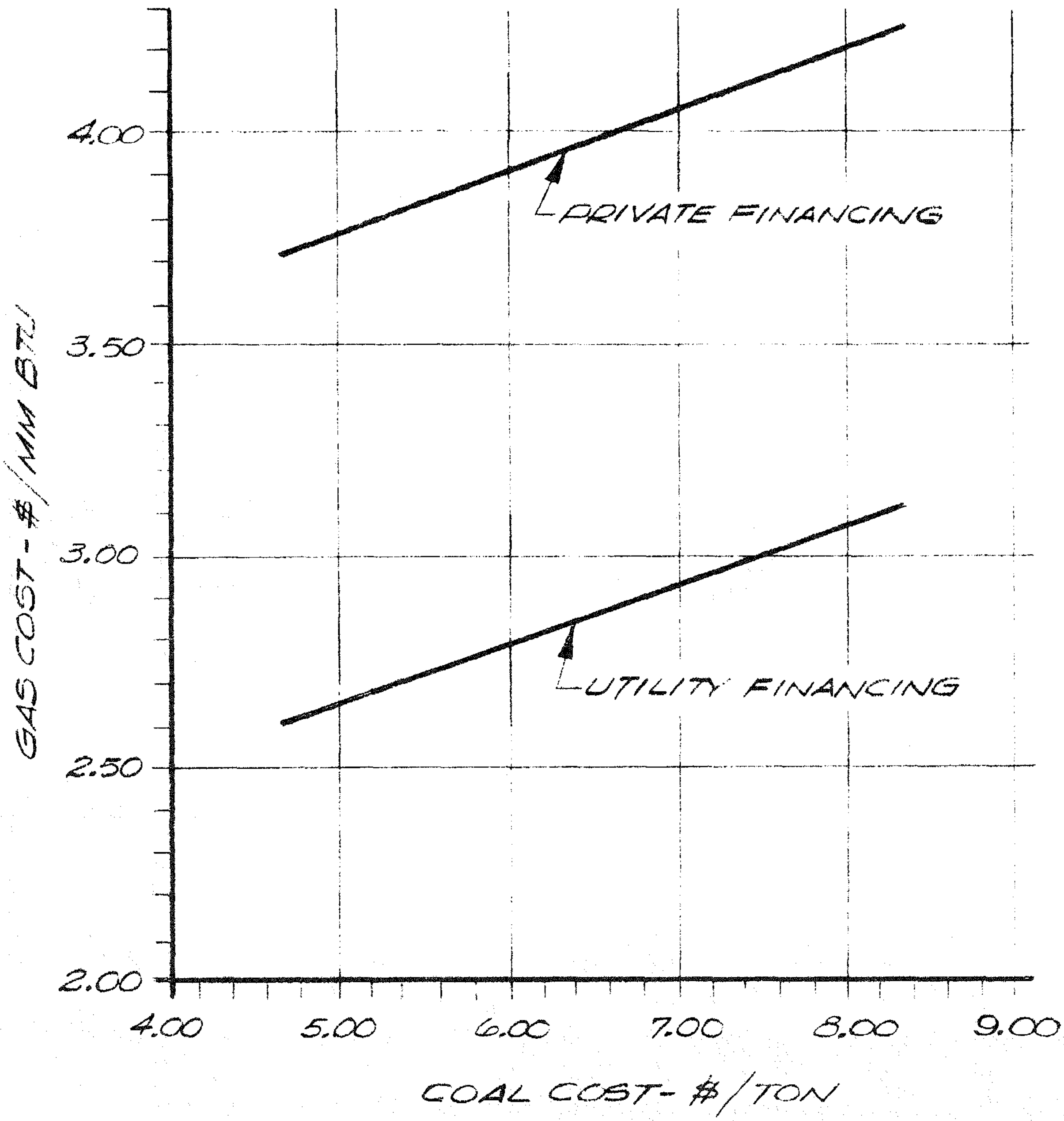
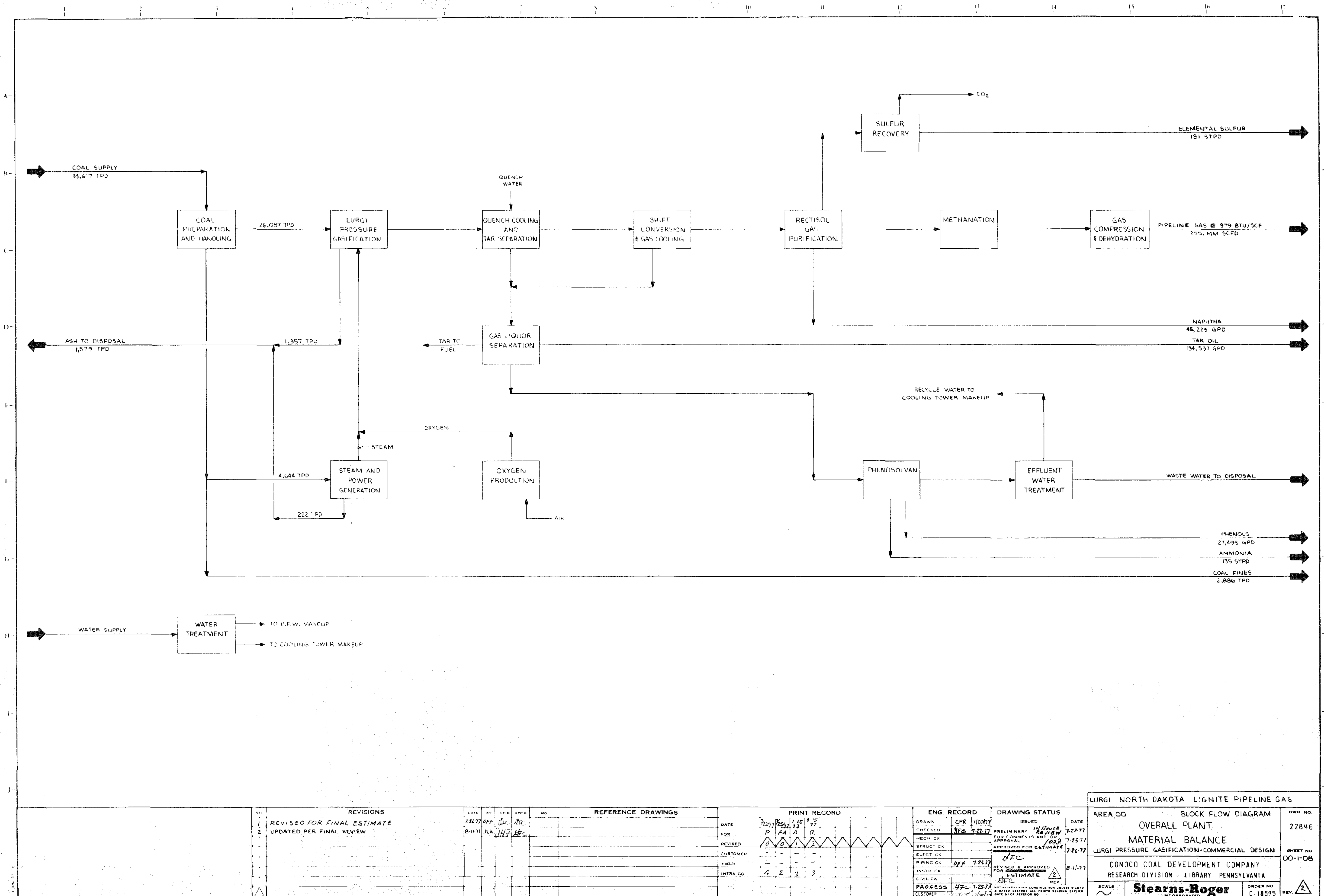


TABLE 9-6

SUMMARY ENERGY BALANCE

	MMBTU/HR	PERCENT OF TOTAL
Energy Input		
Coal to Gasifier, hhv	15,057	82.4
Coal to Steam Plant, hhv	2,680	14.8
Electrical Power Consumption (1.25 MM kwh/day) (Coal equivalent @ 10,000 BTU/kwh)	519	2.8
Total Input	18,256	100.0
Surplus Fines	1,666	
Energy Distribution		
Product Gas, hhv	10,419	57.1
By-Products:		
Tar oil	746	4.1
Naphtha	232	1.3
Crude Phenols	138	0.7
Ammonia	110	0.6
Sulfur	60	0.3
Subtotal, Products & By-Products	11,705	64.1
Consumption & Losses	6,551	35.9
Total Energy Distribution	18,256	100.0
Surplus Fines	1,666	
Cold Gas Efficiency, %		57.1
Plant Thermal Efficiency, %		64.1


COAL SUPPLY 33,617 TPD
COAL PREPARATION AND HANDLING
26,087 TPD
LURGI PRESSURE GASIFICATION
QUENCH WATER
QUENCH COOLING AND TAR SEPARATION
SHIFT CONVERSION & GAS COOLING
RECTISOL GAS PURIFICATION
CO_2
SULFUR RECOVERY
ELEMENTAL SULFUR 181 STPD
METHANATION
GAS COMPRESSION & DEHYDRATION
PIPELINE GAS @ 979 BTU/SCF 255. MM SCFD
NAPHTHA 45,225 GPD
TAR OIL 134,537 GPD
ASH TO DISPOSAL 1,579 TPD
1,357 TPD
TAR TO FUEL
GAS LIQUOR SEPARATION
RECYCLE WATER TO COOLING TOWER MAKEUP
OXYGEN
STEAM
4,644 TPD
STEAM AND POWER GENERATION
222 TPD
OXYGEN PRODUCTION
AIR
PHENOSOLVAN
EFFLUENT WATER TREATMENT
WASTE WATER TO DISPOSAL
PHENOLS 27,493 GPD
AMMONIA 135 STPD
COAL FINES 2,886 TPD
WATER SUPPLY
WATER TREATMENT
TO B.F.W. MAKEUP
TO COOLING TOWER MAKEUP
REVISIONS
1 REVISED FOR FINAL ESTIMATE 7-26-77 OFF
2 UPDATED PER FINAL REVIEW 8-11-77
REFERENCE DRAWINGS
PRINT RECORD
DATE
FOR
REVISED
CUSTOMER
FIELD
INTRA CO.
ENG. RECORD
DRAWN CPR 7/20/77
CHECKED 7-22-77
MECH CK
STRUCT CK
ELECT CK
PIPING CK OFF 7-26-77
INSTR CK
CIVIL CK
PROCESS 7-25-77
CUSTOMER APPROVAL GJJ 7/25/77
DRAWING STATUS
ISSUED DATE
PRELIMINARY IN HOUSE REVIEW 7-22-77
FOR COMMENTS AND/OR APPROVAL 7-25-77
APPROVED FOR ESTIMATE 7-26-77
REVISED & APPROVED FOR ESTIMATE 8-11-77
REV. 2
NOT APPROVED FOR CONSTRUCTION UNLESS SIGNED & DATED DESTROY ALL PRINTS BEARING EARLIER DATE &/OR REVISION NO
LURGI NORTH DAKOTA LIGNITE PIPELINE GAS
AREA OO BLOCK FLOW DIAGRAM
OVERALL PLANT
MATERIAL BALANCE
LURGI PRESSURE GASIFICATION-COMMERCIAL DESIGN
CONOCO COAL DEVELOPMENT COMPANY
RESEARCH DIVISION · LIBRARY PENNSYLVANIA
SCALE ~
Stearns-Roger INCORPORATED
ORDER NO. C-18575
DWG. NO. 22846
SHEET NO 00-1-08
REV. 2

APPENDIX 2

COST OF RECONSTITUTED ACCEPTOR

INTRODUCTION

Acceptor circulating between the process vessels slowly loses activity toward the CO_2 acceptor reaction

$$CaO + CO_2 \longrightarrow CaCO_3.$$

Activity is maintained by withdrawing acceptor from the system and replacing it with fresh stone. At the time of this writing, an Acceptor Reconstitution process is being developed. Upon complete development, the withdrawn acceptor can be processed to restore its activity and returned to the gasification system. Preliminary technical results from the developmental effort are favorable. Therefore a Reconstitution unit (area 1000) was designed, based upon the preliminary technical data, and included in this study. The cost of reconstitution has been calculated and is presented in this appendix. The cost calculations are based on the guidelines established for this study.

BASIS OF CALCULATIONS

1. Capital investment for a Reconstitution Unit installed on a flat and level site is $41,050,000.

2. Electrical power consumed in the unit is valued at $0.0045 per kwh.

3. Reconstituted acceptor cost is calculated for equivalent Minnekahta limestone. Equivalent Minnekahta limestone is the amount of limestone which, upon calcining, will provide the same amount of CaO that is produced in the Reconstitution Unit pelletizer.

4. Composition of Minnekahta limestone is as shown in the Design Basis section of the report.

5. Fuel coal cost is cost of wet coal to the Reconstitution Unit less the coal required to calcine Minnekahta limestone at 1848°F and to heat the air required for combustion of the coal. More coal is required for calcining, thus there is a credit to Reconstitution Unit operating costs.

DISCUSSION

The value of reconstituted acceptor is equivalent to that of Minnekahta limestone at $18.48 per ton. A lesser cost for the limestone would therefore make purchase of fresh limestone more economical than reconstitution of spent acceptor. Calculation of this cost is shown in the attached table. For this study, Minnekahta limestone was valued at $20.00 per ton delivered.

The value of reconstituted acceptor is approximate because information about factors affecting cost is incomplete at this date. The most important of these factors include the following:

APPENDIX 2 - continued

DISCUSSION - continued

1. Cost of site preparation for reconstitution was not considered.

2. Effect of additional plant fencing, roads, fire protection, longer interconnecting lines between other units, etc., when reconstitution is present were not considered.

3. Possible increased diameter of the regenerator when calcining fresh limestone was not considered.

4. Effect of additional air requirements in the regenerator were not considered. This also involves compressor costs and horsepower (utility) increases.

5. Disposal problems and cost for handling spent acceptor without reconstitution were not included.

ANNUAL OPERATING COSTS

RECONSTITUTION UNIT

(In Millions of Dollars)

Raw Materials		
Fuel Coal (-12,388 T/yr @ \$5.00/ton) (Credit)		\$-0.062
Electrical Power (35.625×10^6 Kwh/yr @ \$0.0045/kw)		0.160
Labor		
Process (3 1/3 men/shift x 8304 hr/yr x \$7.30/hr)	\$0.202	
Maintenance (50% of total maintenance)	1.231	
Subtotal	\$1.433	
Supervision (20% of Process & Maintenance)	0.287	
		1.720
Administration and General Overhead		
(60% of total labor)		1.032
Supplies		
Operating (30% of 0.202)	0.061	
Maintenance (½ of total maintenance)	1.232	
		1.293
Local Taxes and Insurance		
(2.7% of $\$41.05 \times 10^6$)		1.106
GROSS ANNUAL OPERATING COSTS		\$5.249

CAPITAL AND INVESTMENT

RECONSTITUTION UNIT

PLANT INVESTMENT	41.05
Startup Costs (20% of 5.249)	1.050
Working Capital	0.418

COST CALCULATION

Cost of Reconstituted Acceptor as equivalent Minnekahta limestone:

$$\text{Cost} = \frac{(5.249 + 0.247 \times 41.05 + 0.1337 \times 1.050 + 0.2305 \times 0.418) \times 10^6}{845{,}420}$$

$$= \frac{15.625 \times 10^6}{845{,}420} = \$18.48\text{/ton of limestone purchased}$$

APPENDIX 3

COST ESTIMATE FOR 62.5 BILLION BTU/DAY PIPELINE GAS PLANT

SUMMARY

INTRODUCTION

Presented within this section is a summary of the results of the order of magnitude estimate for the one-fourth size CO_2 Acceptor Plant. CCDC directed that this plant estimate be based on producing 62.5 billion Btu/day of pipeline gas in two 50% trains. At the request of CCDC, no process design work was done for the one-fourth size plant cost estimate. Capital costs were estimated by prorating equipment costs using exponential factors on the changed capacities and applying suitable multiplying factors to obtain installed costs. The process design and flow sequence of the smaller plant is identical to the conceptual design of the commercial CO_2 Acceptor plant using North Dakota Lignite as feed.

The major feed, product and by-product streams are shown below and are one-fourth of the amounts indicated on page 3, Book 1 of the commercial report. Composition of the product gas is not indicated as it is not known without heat and material balance calculations. However, an assumed heating value is presented. Capital, operating, and gas costs are also summarized below.

Major Feed Streams	
Coal Feed, dry basis, tpsd	4,173.4
Water in Coal Feed, tpsd	2,514.7
Water Supply, net makeup, gpm	927.2
Products--Stream Day Basis	
Pipeline Gas, MMscfd	65.66
Sulfur, ltpd	31.1
Ammonia, tpd	25.3
Product Gas	
Total Moles/Hr	7,218.4
Total Lbs/Hr	113,837.0

Molecular Weight	15.8
Heating Value, Btu/scf (hhv)	951.88

CAPITAL AND OPERATING COSTS
(In Millions of Dollars)

	1/4 CO_2 Acceptor Plant
Total Plant Investment (excluding coal mine)	$309.78
Initial Catalyst and Chemicals	1.28
Paid-Up Royalties	0.30
Allowance for Funds Used During Construction	52.27
Startup Costs	9.02
Working Capital (Private Financing)	8.57
Total Capital Requirement	$381.22
TOTAL GROSS ANNUAL OPERATING COSTS	$ 45.11
BY-PRODUCT CREDITS	1.41
TOTAL NET ANNUAL OPERATING COSTS	$ 43.70

Gas Costs - Utility Financing

Average Gas Cost, $/MMBtu	$ 4.35
First Year Gas Cost, $/MMBtu	5.56

Gas Cost - Private Financing

Constant Gas Cost, $/MMBtu	$ 6.03

APPENDIX 3 - continued

SCOPE OF WORK

The purpose of this work is to provide a cost estimate and new gas cost for a CO_2 Acceptor process Coal Gasification plant which is one-fourth the capacity of the commercial design completed by Stearns-Roger for CCDC in August, 1977. CCDC instructed that the one-fourth size plant estimate be based on producing 62.5 billion Btu/day of pipeline gas via two 50% trains. Basic process design and flow sequence is identical to the commercial plant conceptual design completed by Stearns-Roger. Stearns-Roger was allowed some flexibility in selecting the actual number of downstream gas treating equipment trains. Consideration in choosing the number of trains in each plant area included economy of equipment and package unit sizing, solids feeding, plant startup, and operating flexibility.

As instructed by CCDC and due to the short time allowed to obtain the one-fourth size plant cost estimate, no process design work was done to obtain the new cost of equipment. A couple of vendor contacts were made in the time available to obtain new budget costs on proprietary units in order to evaluate desirability of single or dual train with sizing economy considerations. Some in-house equipment costing was done by Stearns-Roger. Other plant area equipment items were recosted using exponential factors on capacity changes. Suitable multiplying factors were then applied to the equipment costs to obtain the installed costs.

Utilities areas were not balanced or reviewed in detail. Capacity changes were chosen based on engineering judgment to provide costs for the one-quarter capacity plant.

The Gasification Area accounts, such as piping and steel which were obtained by material takeoffs on the commercial design, were reviewed and engineering judgment applied to arrive at representative factors to apply to such accounts to obtain the new costs for these items. The same factors were applied to the accounts taken off for Areas 400 and 800.

COST ESTIMATE BASIS BY PLANT AREA

In general, a basis of two 50% trains was used for the one-fourth size plant producing 62.5 billion Btu/day. However, in some plant areas, economic incentives justified a departure from both the commercial plant operating philosophy and the two 50% train basis. The plant areas and the basis used for costing each are presented below. Deviations from the two 50% train basis is discussed. It should be noted that time was not available during this estimate to economically evaluate single versus dual trains for all plant process areas downstream of Gasification. Using a single train concept throughout all process areas downstream of Area 200 may prove to offer significant capital and operating cost savings and; consequently, provide a lower gas cost than reported herein for the 1/4 size CO_2 Acceptor plant. Reference should be made to the Plant Train Diagrams, Drawing Nos. 00-1-20 and 00-1-21 at the end of this section.

Area 100 - Coal Preparation

A single train is provided for coal unloading facilities. One dead coal storage pile and one live coal storage pile are utilized for the one-fourth size plant. The single train is 25% of the capacity of the commercial plant. Three 50% trains (each train 1/2 size of commercial plant train) of working coal storage and reclaiming facilities are provided. Also, three 50% trains (each train 1/2 size of commercial) of crushing and drying equipment have been included to provide two operating trains and a spare, thus keeping the same operating philosophy used in the commercial plant. The individual train costs for the Williams crusher-dryer units and the wet bottom furnaces were factored based on 50% of commercial size train capacity. A single train of equipment was provided for dried coal storage and coal feed to process facilities.

Area 200 - Gasification

Two 50% trains (each 1/2 capacity of commercial train) of gasification area equipment are provided in the one-fourth size plant cost estimate. Four feed lockhoppers are provided for each gasifier. Future study may show that three such systems may prove adequate for the smaller size (23' Ø versus 33' Ø) gasifiers in the one-fourth size plant. For the two gasifier-regenerator trains in the smaller scale plant, it was estimated that 35% of the total structural steel and civil cost of the commercial gasifier-regenerator structure is required. It should be noted that the plot size at grade for the two smaller gasifier-regenerator trains is larger than that of a single gasifier-regenerator train structure for the commercial plant. Also, the overall structure height does not vary significantly from the commercial plant. Instrumentation material cost is based on 50% of the total instrumentation cost used for Area 200 in the commercial plant design. After reviewing all piping from Areas 200, 400 and 800 in the gasifier-regenerator structure, a weight averaged factor of 25% was applied to the total gasifier structure piping cost in the commercial plant to arrive at the one-fourth size piping cost for Area 200. All account cost factors

Area 200 - Gasification - continued

excluding the equipment account were identical for the 1/4-size plant areas 200, 400 and 800.

Area 300 - Gasification Services

The recycle gas circuit and lock gas system consist of two 50% trains (each train 1/2 capacity of each commercial train). The startup char and dead-burned limestone unloading, storage and transfer facilities are based on one train at 1/4 the size of the commercial plant. One fresh limestone storage system is provided for the one-fourth size plant. It is over-sized by 20% to allow for additional capacity in the limestone surge bin should maintenance be required in Reconstitution, Area 1000.

Area 400 - Raw Gas Cooling

There is one waste heat recovery line per raw gas cooling train. The size and cost of each waste heat recovery line is identical to the cost of one parallel line of Area 400 waste heat recovery equipment in the commercial plant. Two trains of solids scrubbing and ammonia scrubbing facilities (each train 1/2 size of commercial train) are included in the present estimate.

Area 500 - Raw Gas Treating

Two 50% trains of raw gas compression equipment and H_2S removal facilities (Stretford) are incorporated into the one-fourth size plant cost estimate. Each raw gas treating train is 1/4 the capacity of the commercially sized trains. A new installed cost from Peabody Holmes for the smaller Stretford units was obtained and used in the estimate. The raw gas compressor package was reestimated in-house by Stearns-Roger based on the new flow requirements.

Area-600 - Methanation

For the one-quarter size plant estimate employing two 50% trains in methanation, two alternatives were considered. The first alternative was to provide two trains identical to the commercial case, but reduced in size to handle only 50% of the commercial single train flow. For this case, each methanation train would consist of four primary reactors, one recycle gas compressor and one cleanup reactor. A second alternative considered was based on using only two primary reactors, one recycle gas compressor and one cleanup reactor per train of methanation.

The second alternate described was used for this estimate because of the following advantages:

1. The primary reactors are the same size as the commercial reactors, therefore, no factoring of equipment cost is required for the reactors.
2. Economy of reactor cost exists in using two large size reactors per train as opposed to four smaller size reactors per train.
3. Economy of heat exchanger cost exists in that there would be fewer exchanger shells with slightly more surface area per shell.

Area 600 - Methanation - continued

The recycle gas compressor horsepower is approximately 20% more for the second methanation alternative selected. However, the increased cost of the compressor is more than offset by the savings achieved in reactor and heat exchanger costs. Since the commercial plant methanation startup system was designed to startup only one train at a time, the same cost of the startup system was used in the one-quarter size plant estimate.

Area 700 - Product Treating and Compression

This area includes two 50% trains of equipment for CO_2 removal (Benfield) and product gas compression. Each train is 25% of the commercial train size. The product gas compressors were reestimtaed in-house by Stearns-Roger. Benfield costs were factored from the commercial size unit costs. A single dehydration train (25% of the commercial train capacity) is included.

Area 800 - Flue Gas System

Two 50% trains (each train handling 50% of the commercial train capacity) of equipment are provided in Area 800. Included are the high efficiency ash removal cyclones, waste heat recovery equipment, carbon monoxide oxidation, tertiary solids removal and power recovery by hot gas expanders. It should be noted that there exists only one line of primary particulate separation equipment, waste heat recovery equipment and carbon monoxide oxidation equipment per train in the one-fourth size plant. The size and cost of this equipment per train is identical to that of one parallel line of this equipment in the commercial size train. The total cost of two 1/2 commercial size tertiary solids removal vessels are considered equal to the cost of one commercial size unit. The hot gas expander package cost for each smaller train is taken from vendor preliminary estimates.

Area 900 - Ash Handling

This area includes heat recovery and cooling of ash, ash slurrying and treating facilities and ash slurry pumping facilities. Two 50% trains of equipment are provided (each train 25% of commercial train capacity).

Area 1000 - Acceptor Reconstitution

Two trains of equipment (each train 75% of commercial train size) including transfer and storage of reject acceptor, hydration, inert separation facilities, and calcination are provided. A single cooling tower package at 25% of commercial capacity is included. The reconstituted acceptor storage and transfer to process facilities consists of a single train oversized by 10% to allow for surge storage during maintenance periods.

Area 1100 - Sulfur Recovery

One Claus unit at 25% of commercial capacity is included. The Claus unit cost is factored from the commercial unit cost.

Area 1200 - Ammonia Recovery

This area includes one train and is estimated based on handling 10% more than one-fourth commercial flow. U.S. Steel provided a new cost estimate for the Phosam-W Process unit in the 1/4 scale plant. An additional 10% overdesign is included in the estimate for the ammoniacal water holding tank to allow for Phosam-W unit maintenance.

Area 1300 - SO_2 Removal

Two 100 % capacity trains are provided for SO_2 removal. Each train would be 50% of capacity of the commercial size train.

Utilities

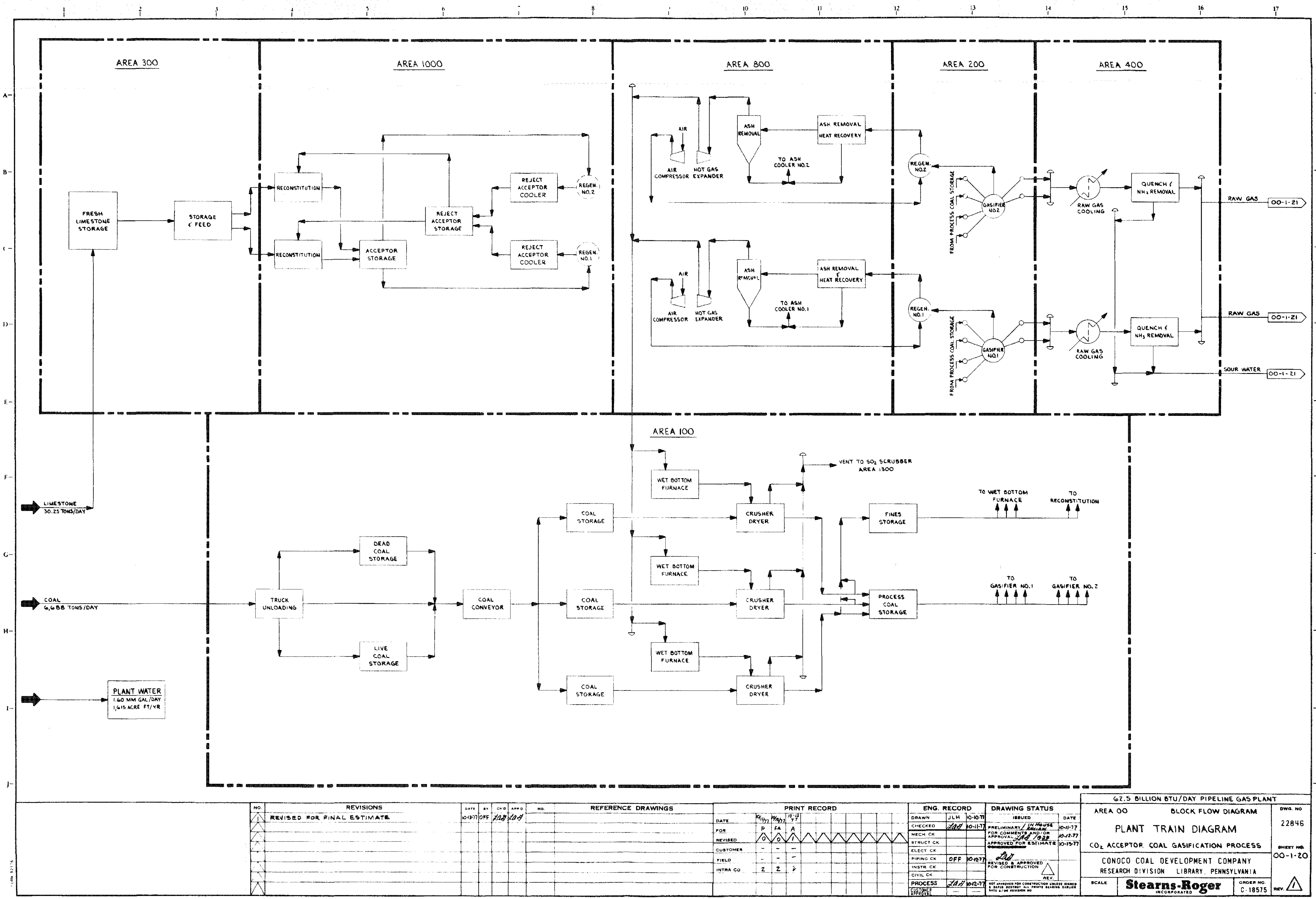
Since no new utilities balance was possible due to the time available, an arbitrary 30% of commercial capacity was used to estimate the costs of Raw Water Supply and Treating, Waste Solids Disposal, Steam System, Electrical Power Generation and Miscellaneous Utilities. Two 100% capacity turbo-generators (each turbo-generator handling 25% of the commercial electrical load) were requoted by the vendor for the one-fourth plant estimate. The plant cooling water system cost is based on 25% of the commercial capacity.

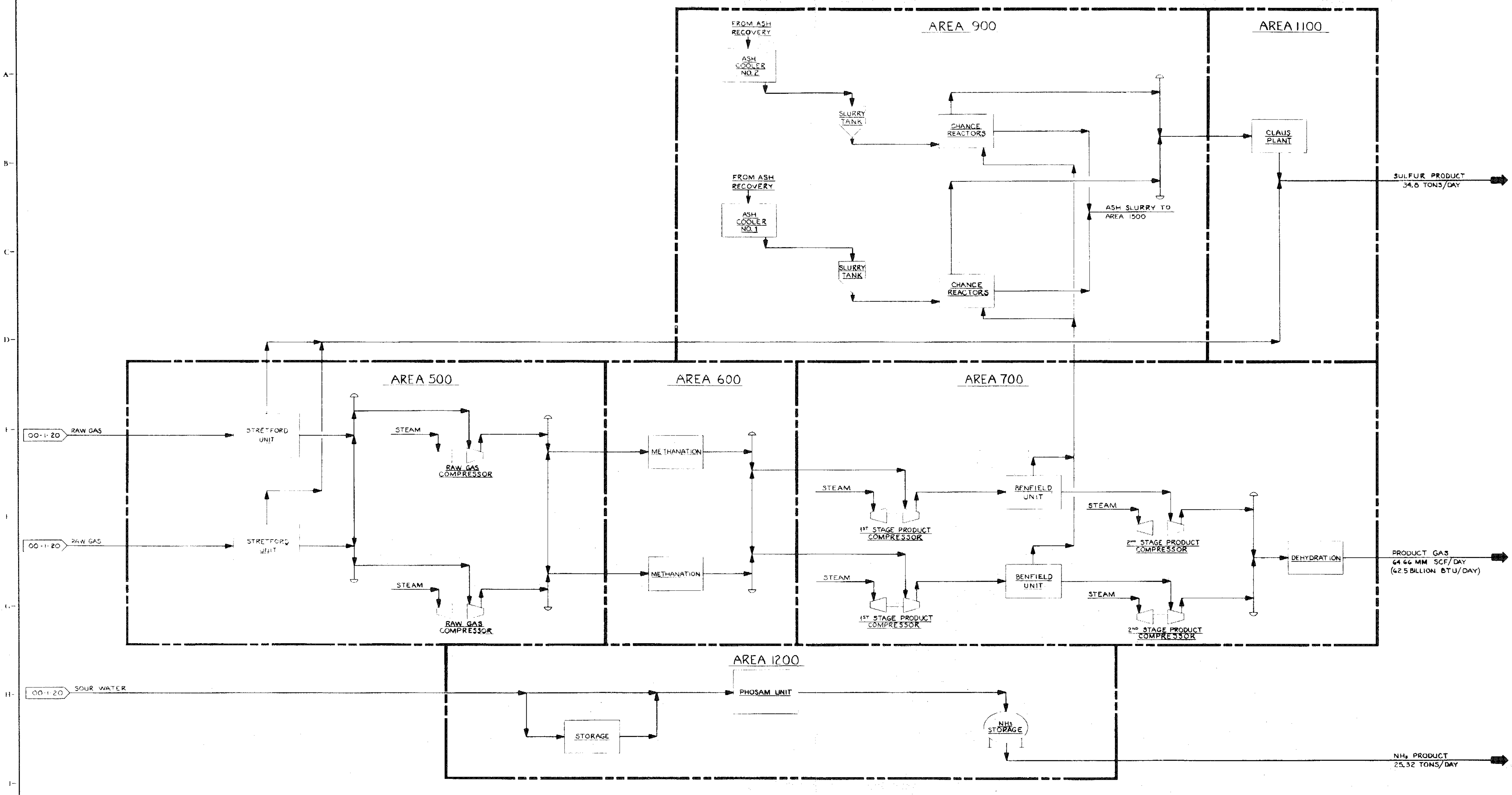
Area 1900 - General Facilities

This area includes the plant flare, system and storage facilities that are not included within a process area. An arbitrary 28% of commercial capacity was used as a basis for costing this area.

Area 2000 - General Facilities

These facilities were scaled down as judgment directed in accordance with plot area being reduced to 40% of the commercially sized plant, reduced vehicle requirements, and smaller buildings.


AREA 300
FRESH LIMESTONE STORAGE
STORAGE & FEED
AREA 1000
RECONSTITUTION
RECONSTITUTION
ACCEPTOR STORAGE
REJECT ACCEPTOR STORAGE
REJECT ACCEPTOR COOLER
REJECT ACCEPTOR COOLER
REGEN. NO. 2
REGEN. NO. 1
AREA 800
AIR
AIR COMPRESSOR
HOT GAS EXPANDER
ASH REMOVAL
ASH REMOVAL & HEAT RECOVERY
TO ASH COOLER NO. 2
TO ASH COOLER NO. 1
AREA 200
REGEN. NO. 2
REGEN. NO. 1
GASIFIER NO. 2
GASIFIER NO. 1
FROM PROCESS COAL STORAGE
AREA 400
RAW GAS COOLING
QUENCH & NH_3 REMOVAL
RAW GAS
00-1-21
SOUR WATER
AREA 100
LIMESTONE 30.25 TONS/DAY
COAL 6,688 TONS/DAY
PLANT WATER 1.60 MM GAL/DAY 1,615 ACRE FT/YR
TRUCK UNLOADING
DEAD COAL STORAGE
LIVE COAL STORAGE
COAL CONVEYOR
COAL STORAGE
WET BOTTOM FURNACE
CRUSHER DRYER
VENT TO SO_2 SCRUBBER AREA 1300
FINES STORAGE
PROCESS COAL STORAGE
TO WET BOTTOM FURNACE
TO RECONSTITUTION
TO GASIFIER NO. 1
TO GASIFIER NO. 2
REVISIONS
REVISED FOR FINAL ESTIMATE
REFERENCE DRAWINGS
PRINT RECORD
ENG. RECORD
DRAWN JLH 10-10-77
CHECKED 10-11-77
PIPING CK OFF 10-12-77
PROCESS 10-12-77
DRAWING STATUS
PRELIMINARY / IN HOUSE REVIEW 10-11-77
FOR COMMENTS AND/OR APPROVAL 10-12-77
APPROVED FOR ESTIMATE 10-13-77
REVISED & APPROVED FOR CONSTRUCTION
NOT APPROVED FOR CONSTRUCTION UNLESS SIGNED & DATED. DESTROY ALL PRINTS BEARING EARLIER DATE &/OR REVISION NO
62.5 BILLION BTU/DAY PIPELINE GAS PLANT
AREA 00 BLOCK FLOW DIAGRAM
PLANT TRAIN DIAGRAM
CO_2 ACCEPTOR COAL GASIFICATION PROCESS
CONOCO COAL DEVELOPMENT COMPANY
RESEARCH DIVISION LIBRARY, PENNSYLVANIA
Stearns-Roger INCORPORATED
DWG. NO. 22846
SHEET NO. 00-1-20
ORDER NO. C-18575



NO.	REVISIONS	DATE	BY	CH'D	APP'D
1	REVISED FOR FINAL ESTIMATE	10-13-77	OFF	LAD	LAD

REFERENCE DRAWINGS

PRINT RECORD: DATE / FOR / REVISED / CUSTOMER / FIELD / INTRA CO.

ENG. RECORD		
DRAWN	DEH	10-11-77
CHECKED	LAD	10-11-77
MECH CK.		
STRUCT. CK		
ELECT. CK.		
PIPING CK.	OFF	10-12-77
INSTR. CK.		
CIVIL CK.		
PROCESS	LAD	10-12-77
CUSTOMER APPROVAL	—	—

DRAWING STATUS ISSUED	DATE
PRELIMINARY In House Review	10-11-77
FOR COMMENTS AND/OR APPROVAL LAD 1079	10-12-77
APPROVED FOR ESTIMATE	10-13-77
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NOT APPROVED FOR CONSTRUCTION UNLESS SIGNED & DATED. DESTROY ALL PRINTS BEARING EARLIER DATE &/OR REVISION NO.

62.5 BILLION BTU/DAY PIPELINE GAS PLANT

AREA 00 BLOCK FLOW DIAGRAM

PLANT TRAIN DIAGRAM

CO_2 ACCEPTOR COAL GASIFICATION PROCESS

CONOCO COAL DEVELOPMENT COMPANY
RESEARCH DIVISION - LIBRARY, PENNSYLVANIA

SCALE

Stearns-Roger INCORPORATED

DWG. NO. 22846
SHEET NO. 00-1-21
ORDER NO. C-18575
REV. 1

PLANT INVESTMENT

The total plant investment for the CO_2 Acceptor Process Plant producing 62.5 Billion Btu/Day of pipeline gas at Beulah, North Dakota, is estimated to be $309,780,000 in mid-1977 dollars. The breakdown of this cost into the various areas is found in Table 1 in this appendix.

This estimate is of a preliminary nature and was developed with the use of exponential factors on the changed capacities. From this, new equipment costs for the one-fourth commercial capacity plant were generated. The direct field cost installation factors of the commercial plant were upgraded, to allow for the smaller equipment, and then applied to the new equipment costs to obtain the new one-fourth capacity estimate.

The labor productivity for the North Dakota area was assumed to be 100% as used in the commercial plant estimate. It was assumed that an adequate local labor pool exists and that no camp would be required. Current applicable union wage rates were used for all crafts except finishers and teamsters; these unions had not reached a new agreement when the estimate was prepared and an average escalation was assumed. The average direct craft rate is $10.13 per hour. All labor is based on a forty-hour week with casual overtime only.

Subsistence and travel pay, in addition to normal fringe benefits, have been included where appropriate.

Indirect Field Costs were calculated at 125% of Direct Field Labor Cost. A sales and use tax of 3.0% was included and casual premium pay of 3.0% of Direct Field Labor was added. To this subtotal, engineering costs and contractor fees were added at 14% based on individual plant areas and Stearns-Roger historical data on similarly sized jobs.

Contingency was added to the Total Field and Engineering Cost (including fees, sales tax and premium pay) at 25%. The accuracy range of the estimate is $\pm$ 30%.

The costs of spare parts, catalysts and chemicals, license fees, start-up, interest during construction, working capital, client overhead and administrative costs are not included. Refer to the section entitled "Capital and Operating Costs".

TABLE 1

PLANT INVESTMENT

(In Millions of Dollars)

62.5 BILLION BTU/DAY PIPELINE GAS PLANT

AREA NO.	ITEM	AREA COST
100	Coal Preparation	45.52
200	Gasification	59.88
300	Gasifier Services	7.16
400	Raw Gas Cooling	11.50
500	Raw Gas Treating	15.60
600	Methanation	19.44
700	Product Treating and Compression	16.25
800	Flue Gas System	35.17
900	Ash Handling	3.50
1000	Acceptor Reconstitution	15.98
1100	Sulfur Recovery	2.04
1200	Ammonia Recovery	7.74
1300	SO_2 Removal	21.18
1400	Raw Water Supply and Treating	2.94
1500	Waste Solids Disposal	0.79
1600	Steam System	7.65
1700	Electrical	10.06
1800	Miscellaneous Utilities	7.75
1900	Offsites	2.10
2000	General Facilities	17.53
	TOTAL PLANT INVESTMENT	309.78

CAPITAL AND OPERATING COSTS

INTRODUCTION

The summary of capital and operating costs as well as gas cost for the CO_2 Acceptor plant is shown in Table 2. In addition, there are tables in this section showing capital requirements, annual operating costs, and annual maintenance costs. Basis of economic data and evaluation was the C.F. Braun & Company Guidelines report (10).

When comparing various gasification processes, gas cost is an important consideration. Cost of gas depends upon many factors, e.g., investment amortization, taxes, operating costs, etc. This section of the report discusses those factors.

FINANCING METHOD

There are two potential methods of financing the gasification plants. These are the utility financing and private investor financing methods. Because the gas costs are of interest to various groups, the costs are presented using both methods.

The following parameters apply to the two financing methods:

	Utility	Private
Project life, years	20	20
Debt - equity ratio	75/25	100% Equity
Return on equity, %	15	12 DCF
Federal Income Tax, %	48	48

Depreciation for the utility method is 20-year straight line depreciation on plant investment, allowance for funds used during contruction, and capitalized portion of start-up costs. Depreciation for the private investor method is 16-year sum of the years digits depreciation on total plant investment.

CAPITAL REQUIREMENT

Capital requirements and calculation methods are shown in the Table 3. Cost of initial charge of catalyst and chemicals reported in the table was calculated using 25% of the commercial plant's initial catalyst charge and 30% of the initial chemicals charged to the commercial size plant.

ANNUAL OPERATING COSTS

Annual operating costs are shown in the Table 4. Process operating manpower

CAPITAL AND OPERATING COSTS - continued

was estimated in accordance with complexity and number of units being operated. Maintenance labor is taken as 50% of total maintenance cost. On-stream factor is 0.9. All costs, wages, and product values are mid 1977 values.

MAINTENANCE COSTS

Annual maintenance costs shown in the Table 5 are calculated as a percent of installed unit cost. Maintenance factors are as shown in the table.

CATALYST AND CHEMICALS

Annual catalyst replacement and chemical make-up and consumption costs are defined in annual Catalyst and Chemical Consumption Costs, Table 6.

GAS COST

A chart showing the variation of gas cost with coal cost is included in this section.

TABLE 2

SUMMARY OF CAPITAL AND OPERATING COSTS

(In Millions of Dollars)

62.5 BILLION BTU/DAY PIPELINE GAS PLANT

	UTILITY FINANCING		PRIVATE FINANCING
	AVERAGE GAS COST	FIRST YEAR GAS COST	CONSTANT GAS COST
Capital Costs			
Total Plant Investment	$309.78	$309.78	$309.78
Initial Charge of Catalysts and Chemicals	1.28	1.28	1.28
Allowance for Funds Used During Construction	52.27	52.27	52.27
Paid-Up Royalties	0.30	0.30	0.30
Start-Up Costs	9.02	9.02	9.02
Working Capital	7.13	8.16	8.57
TOTAL CAPITAL REQUIREMENT	$379.78	$380.81	$381.22
Operating Costs			
Raw Materials	11.26	11.26	11.26
Catalysts and Chemicals	1.25	1.25	1.25
Labor			
Process Operating Labor	1.94	1.94	1.94
Maintenance Labor	6.83	6.83	6.83
Supervision	1.75	1.75	1.75
Administration and General Overhead	6.31	6.31	6.31
Supplies			
Operating	0.58	0.58	0.58
Maintenance	6.83	6.83	6.83
Local Taxes and Insurance	8.36	8.36	8.36
TOTAL GROSS OPERATING COSTS/YEAR	$ 45.11	$ 45.11	$ 45.11
TOTAL BY-PRODUCT CREDITS	$ 1.41	$ 1.41	$ 1.41
TOTAL NET OPERATING COSTS/YEAR	$ 43.70	$ 43.70	$ 43.70
AVERAGE GAS COST, $/MMBTU	4.35		
FIRST YEAR GAS COST, $/MMBTU		5.56	
CONSTANT GAS COST, $/MMBTU			6.03

TABLE 3

CAPITAL REQUIREMENTS

(In Millions of Dollars)

62.5 BILLION BTU/DAY PIPELINE GAS PLANT

		UTILITY FINANCING		PRIVATE FINANCING
		AVERAGE GAS COST	FIRST YEAR GAS COST	CONSTANT GAS COST
Total Plant Investment		$ 309.78	$ 309.78	$ 309.78
Initial Charge of Catalysts and Chemicals				
Catalysts	$0.76			
Chemicals	0.52			
		1.28	1.28	1.28
Paid-Up Royalties		0.30	0.30	0.30
Allowance for Funds Used During Construction (Total plant investment x average spending period in years x 9%)		52.27	52.27	52.27
Startup Costs (20% of total annual gross operating costs)		9.02	9.02	9.02
Working Capital (14-day inventory of raw materials + materials and supplies at 0.9% of total plant investment + net receivables at 1/24 annual gas and by-product revenue at calculated sales price)		7.13	8.16	8.57
TOTAL CAPITAL REQUIREMENT		$ 379.78	$ 380.81	$ 381.22

TABLE 4

ANNUAL OPERATING COSTS

(In Millions of Dollars)

62.5 BILLION BTU/DAY PIPELINE GAS PLANT

Raw Materials		
Coal (6,688 tons/day @ $5.00/ton)	$10.98	
Limestone (30.25 tons/day @ $20.00/ton)	.20	
Char	.07	
Dead Burned Limestone	.01	
Total Raw Materials		$11.26
Catalysts and Chemicals		
Catalyst (replacement)	$ 0.76	
Chemicals (makeup and consumption)	0.49	
Total Catalysts and Chemicals		$ 1.25
Labor		
Process Operating Labor (32 men/shift @ $7.30/hr)	$ 1.94	
Maintenance (50% of total maintenance)	6.83	
Supervision (20% of direct labor)	1.75	
Total Labor		$10.52
Administrative and General Overhead		
(60% of labor)		$ 6.31
Supplies		
Operating (30% of process operating labor)	$ 0.58	
Maintenance (50% of total maintenance)	6.83	
Total Supplies		$ 7.41
Local Taxes and Insurance		
(2.7% of total plant investment)		$ 8.36
TOTAL GROSS ANNUAL OPERATING COST		$45.11
By-Product Credits		
Sulfur (31.12 long tons/day @ $25/long ton)	$ 0.25	
Ammonia (25.32 ton/day @ $140/ton)	1.16	
Total By-Product Credits		$ 1.41
TOTAL NET ANNUAL OPERATING COSTS		$43.70

TABLE 5

ANNUAL MAINTENANCE COSTS

(In Millions of Dollars)

62.5 BILLION BTU/DAY PIPELINE GAS PLANT

AREA NO.	ITEM	AREA COST	MAINTENANCE FACTOR PERCENT	ANNUAL MAINTENANCE COST
100	Coal Preparation	45.52	6	2.73
200	Gasification	59.88	6	3.59
300	Gasifier Services	7.16	3	0.21
400	Raw Gas Cooling	11.50	5	0.57
500	Raw Gas Treating	15.60	3	0.47
600	Methanation	19.44	3	0.58
700	Product Treating & Compression	16.25	3	0.49
800	Flue Gas System	35.17	5	1.76
900	Ash Handling	3.50	6	0.21
1000	Acceptor Reconstitution	15.98	6	0.96
1100	Sulfur Recovery	2.04	3	0.06
1200	Ammonia Recovery	7.74	3	0.23
1300	SO_2 Removal	21.18	6	1.27
1400	Raw Water Supply & Treating	2.94	1	0.03
1500	Waste Solids Disposal	0.79	6	0.05
1600	Steam System	7.65	1	0.08
1700	Electrical	10.06	1	0.10
1800	Miscellaneous Utilities	7.75	1	0.08
1900	Offsites	2.10	1	0.02
2000	General Facilities	17.53	1	0.17
	TOTAL	309.78		13.66

Annual Maintenance Costs

Maintenance Labor	6.83
Maintenance Supplies	6.83

Average Maintenance Factor = $\frac{13.66}{309.78} = 0.0441$

TABLE 6

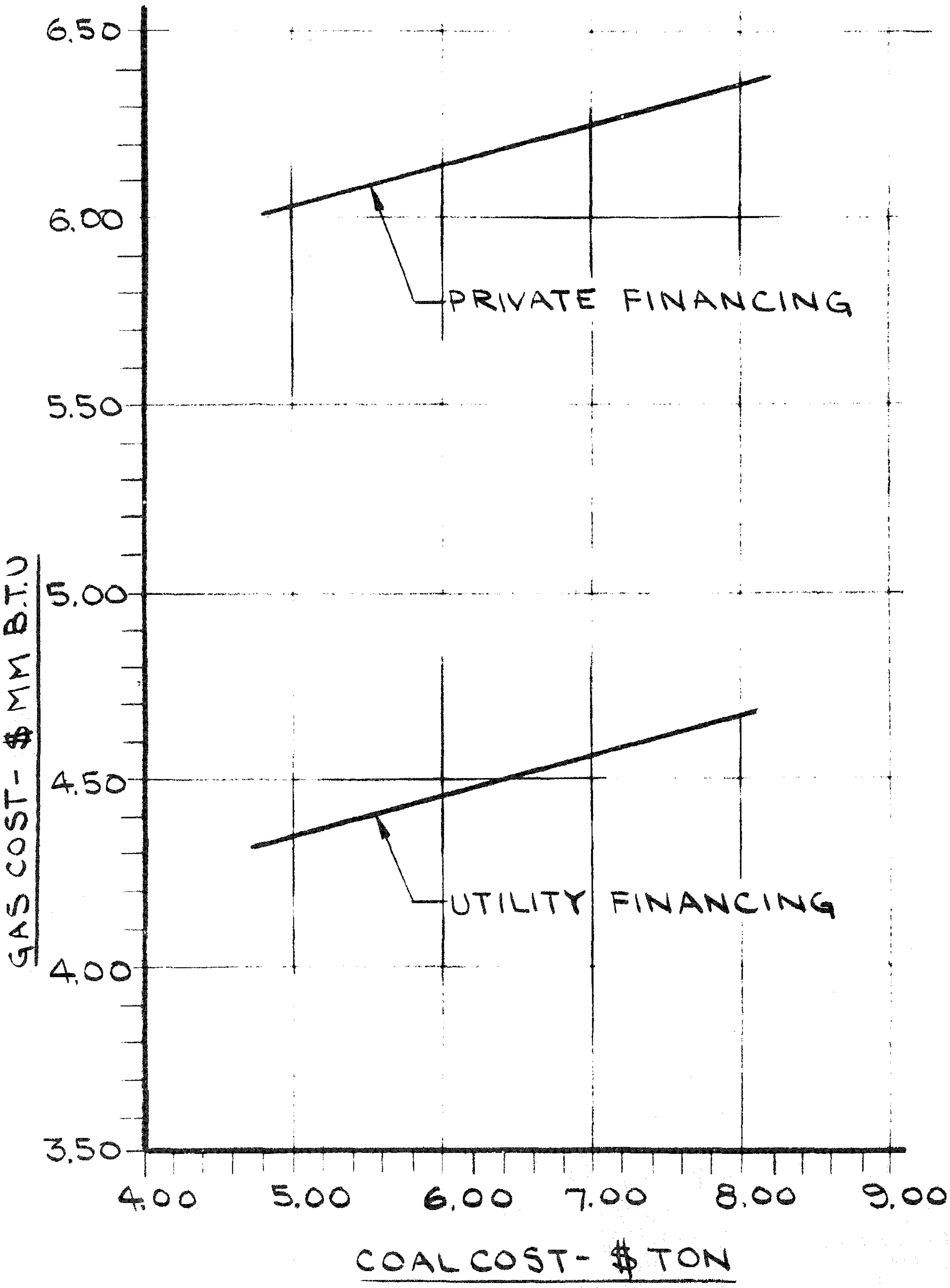
CHEMICALS AND CATALYSTS

62.5 BILLION BTU/DAY PIPELINE GAS PLANT

Catalysts Area Number	Designation	
600	Methanation	$ 758,300
1100	Sulfur Recovery	940
	Subtotal Catalysts	$ 759,240

Chemicals Area Number		
500	Raw Gas Treating (Stretford)	$ 16,935
700	Benfield Solution TEG	3,300 9,300
1200	Ammonia Recovery (PHOSAM-W)	42,900
1400	Raw Water Supply and Treating	417,345
	Subtotal Chemicals	$ 489,780
	TOTAL CATALYSTS AND CHEMICALS	$1,249,020

GAS COST VS. COAL COST
62.5 BILLION B.T.U/DAY PIPELINE GAS PLANT



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