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MASTER

PIPELINE GAS DEMONSTRATION PLANT

Phase I.

ABBREVIATED VERSION OF

PROCESS EVALUATION REPORT
CONCEPTUAL COMMERCIAL PLANT

SECTION 3

PROCESS DESIGN STUDIES

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PROCESS EVALUATION REPORT
CONCEPTUAL COMMERCIAL PLANT

This Process Evaluation Report (PER) contains the results and recommendations of comprehensive analyses and studies which were made to optimize the ICGG Commercial Plant Baseline Process Concept for producing synthetic pipeline gas (SPG) from coal. Design studies to optimize the thermal efficiency and economic attractiveness of the COGAS Process Areas of the plant were conducted along with design studies and trade-off studies of available process subsystems to complement the COGAS Process Areas. The results, recommendations and description of the work accomplished in developing the PER are contained in six separately bound sections of the PER which are summarized as follows:

<u>Section 1 - Executive Summary</u>

This section provides an overview of the total PER and presents results, recommendations and conclusions in a brief format.

<u>Section 2 - Process Analysis - Commercial Plant Concept</u>

This section gives a brief description of plant size, configuration, feedstocks, operating conditions, products and by-products for the Commercial Plant Concept.

<u>Section 3 - Process Design Studies</u>

This section describes various design studies which were conducted to optimize the COGAS Process Areas and other plant areas.

<u>Section 4 - Trade-off Studies</u>

This section describes those trade-off studies which were made to select processes which would best complement the COGAS Process Areas and provide the most efficient and economical Commercial Plant Concept.

<u>Section 5 - Baseline Process Concept</u>

This section describes the ICGG Commercial Plant Baseline Process Concept which was developed from the Tentative Baseline Design (TBD) and was used as a standard for measuring the results of the Process Design Studies and Trade-off Studies.

<u>Section 6 - Baseline Economic Evaluation</u>

This section describes an economic evaluation which has been performed for the Commercial Plant Baseline Process Concept, described in Section 5.

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TABLE OF CONTENTS

3.0 PROCESS DESIGN STUDIES

3.1 AREA 102 - COAL PREPARATION

3.1.1 Summary

Introduction

Two process design studies were done for Area 102. Coal drying and pressurization studies were performed to determine ways to reduce capital costs, reduce operating costs, increase reliability, reduce maintenance expenses, increase flexibility and reduce technical risks while trying to improve plant thermal efficiency. Coal pressurization studies were performed to examine the use of the Fuller-Kinyon pumps to feed ground coal to pyrolysis, in place of lock hoppers, and to determine the impact of their use on the gas selling price. Initially it was intended to make the comparisons on a technical basis only, which included environmental consideration. The studies were later extended to include economics.

Alternates Considered

Three alternates were considered for coal drying to determine the optimum amount of drying to be done ahead of pyrolysis, the type of dryer to be used, and the optimum source of heat to accomplish the drying. The coal drying options were:

- No drying ahead of pyrolysis.
- Mild drying ahead of pyrolysis to reduce the coal moisture content from 10% to 7% by weight.
- Thorough drying ahead of pyrolysis to reduce the coal moisture content from 10% to 3% by weight.

The heat source for drying the coal was either hot flue gas generated by the combustion of coal, or hot flue gas from Flue Gas Power Recovery, Area 110. Fluid-bed and flash drying equipment were considered.

Mild coal drying, from 10% to 7% moisture in a fluid-bed dryer using hot flue gas produced by the combustion of coal, is the TBD or Base Case. The two alternatives are compared to this case, although further study showed that a more elaborate fluid-bed drying scheme would be required to accomplish the drying specified.

In coal pressurization, the Fuller-Kinyon dry solids feeding system was compared to the lock-hopper feeding system used as the baseline in the TBD.

3.1 AREA 102 - COAL PREPARATION

3.1.1 Summary

Evaluation

The different coal drying schemes were evaluated quantitatively on the basis of installed costs and operating costs; and qualitatively on the basis of technical or degree of development, flexibility, turndown, and impact on other areas. The evaluation report on this study is in Subsection 3.1.2.

Coal pressurization systems were compared on the basis of capital costs, operating costs, commercial experience, reliability, flexibility and system impact. The complete evaluation of coal pressurization is in Subsection 3.1.3.

Recommendations

The final coal drying selection was based on economics because there were no outstanding technical differences among the three processing schemes. It is recommended that the fluid-bed drying scheme shown in the TBD should be eliminated. The scheme with no drying ahead of pyrolysis is recommended, since it results in an annual savings of approximately $2.2 MM.

Gas prices may be reduced by 5.0¢/MM Btu if the Fuller-Kinyon pressurization system is used. Other factors such as degree of development, flexibility, reliability and impact on the rest of the process are judged equal except in one area. The amount of particle size attrition caused by Fuller-Kinyon pumps may be excessive for Western coals. It is recommended that the Base Case lock-hopper system be replaced by a Fuller-Kinyon coal feeding system in the Commercial Plant Conceptual Design for Illinois No. 6 coal feed and further studies are proposed to determine attrition rates for Western coals with Fuller-Kinyon pumps.

3.0 PROCESS DESIGN STUDIES

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Summary

The intent of the Dryer Design Study was to determine the optimum amount of drying to be done ahead of pyrolysis in Area 102, the type of dryer and the optimum source of heat to accomplish the drying. This was to be done for each of two coals: Illinois No. 6 Seam, and a Western subbituminous. Pittsburgh No. 8 coal contains only 4% moisture as received, and did not warrant a separate drying step.

The following vendors of drying equipment and/or drying systems were solicited for information:

- Cameron Engineers (flash dryer)
- Williams Patent Crusher and Pulverizer Co. (flash dryer)
- FMC - Materials Handling System Division (fluid bed dryer)
- Bepex Corp. - Strong Scott Division (flash dryer)
- Combustion Engineering - Raymond Division (flash dryer)

For the Illinois coal at 10% moisture, three cases were selected for the trade-off study to cover the complete range of possible choices. One case was not to dry the coal at all; a second case corresponded to the process indicated in the TBD where the coal was mildly dried to 7% moisture and heated to 130°F; and the third case was to thoroughly dry the coal to 3% moisture, allowing it to heat up to 200°F.

For the Western subbituminous coal at 25% moisture, three cases were also considered. One case consisted of no drying of the coal; a second case consisted of drying the coal to a moisture level of 10% using flue gas from Area 110 at approximately 850°F; a third case consisted of drying the coal to 3% moisture, using 850°F flue gas supplemented by coal firing to bring the flue gas temperature to about 1600°F.

Budget estimates have been received from vendors for the systems studied, or for similar systems, so that costs can be extrapolated. The cost impact on other areas has been estimated. Capital costs are summarized in Table 3.1.2-1.

The major impact of the coal drying on operating costs is in energy efficiency and the utilization of waste heat. Differences in thermal performance are expressed as equivalent tons/day of coal. The impact on waste heat recovered from Area 110 and Area 115 is included. Operating costs are summarized in Table 3.1.2-1.

3.0 PROCESS DESIGN STUDIES

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Table 3.1.2-1

Drying System Costs - Commercial Plant
Three Trains

SYSTEM	ILLINOIS COAL: TBD	ILLINOIS COAL: Flue Gas Drying	ILLINOIS COAL: No Drying	WESTERN COAL: No Drying	WESTERN COAL: Flue Gas Drying	WESTERN COAL: Flue Gas and Coal Fired Drying
Coal Moisture - Wt %						
to Area 2	10	10	10	25	25	25
to Pyrolysis	7	3	10	25	10	3
Coal temperature °F	130	200	60	60	200	200
Equivalent T/D of coal required for drying and process heat. (as received)	160	138	174	1,099	966	753
Total Fixed Capital $M (Includes drying system as well as incremental costs incurred in pyrolysis, waste heat recovery & utilities)	20,279	19,366	10,940	60,111	47,871	44,703
Yearly costs 10MM'$ (330 days/yr. Capital at 25% of fixed capital)	5.070	4.834	2.735	15.078	11.968	11.175
Direct Annual Cost 10 MM'$/yr. Illinois @ $22.50/ton Western @ $8/ton	1.188	1.0246	1.292	2.901	2.550	1.988
Total 10 MM'$/yr.	6.258	5.859	4.027	17.929	14.523	13.163
Impact on gas price ¢/MM BTU	7.98	7.470	5.13	22.86	18.51	16.78

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Summary - continued

For the Illinois coal, it is recommended that no drying be done in Area 102. For the Western coal, it is recommended that drying be done in a flash dryer to a moisture level of 3%.

Performance Specifications

A number of different drying levels were considered in the study. The general performance specifications were:

1. If coal is exposed to the atmosphere it must be limited to a temperature less than 140°F in order to avoid oxidation.

2. The coal entering pyrolysis must be crushed and screened to meet the specified size consist.

3. Drying heat may be supplied by flue gas from Area 110.

4. If supplementary heat is required, it should be provided by coal combustion.

5. Flue gas used for drying shall contain a maximum of 2% oxygen.

6. The coal may not be heated above 220°F in a 2% oxygen environment.

Vendor Responses

Processes Considered

The process recommended as a result of the trade-off study is not unique to one particular vendor but rather a specification of desired operating conditions. A number of different vendor responses were received which met the requirements, using either a flash dryer or fluid bed dryer. The responses do not directly correspond to each other as they were received at different times and are for different feed rates, moisture levels and drying conditions. However, an attempt has been made to put the budget estimates on the same basis by comparing them for the same evaporation rates

3.0 PROCESS DESIGN STUDIES FE-2012-096
MAY 1980

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Vendor Responses

Processes Considered - continued

and common auxiliary equipment. The estimates received were of adequate accuracy for the performance of a trade-off study.

Process Description

Three process cases have been considered for each of the two coals; Illinois No. 6 and Western subbituminous. These are listed below and described separately in the following paragraphs.

Illinois No. 6 Coal	Fig. No.
o Fluid bed drying to 7% moisture (TBD)	3.1.2-1
o Flash drying to 3% moisture	3.1.2-2
o No drying	3.1.2-3
Western Subbituminous Coal	
o No drying	3.1.2-4
o Flash drying to 10% moisture with flue gas	3.1.2-5
o Flash drying to 3% moisture with flue gas and coal combustion	3.1.2-6

Illinois Coal

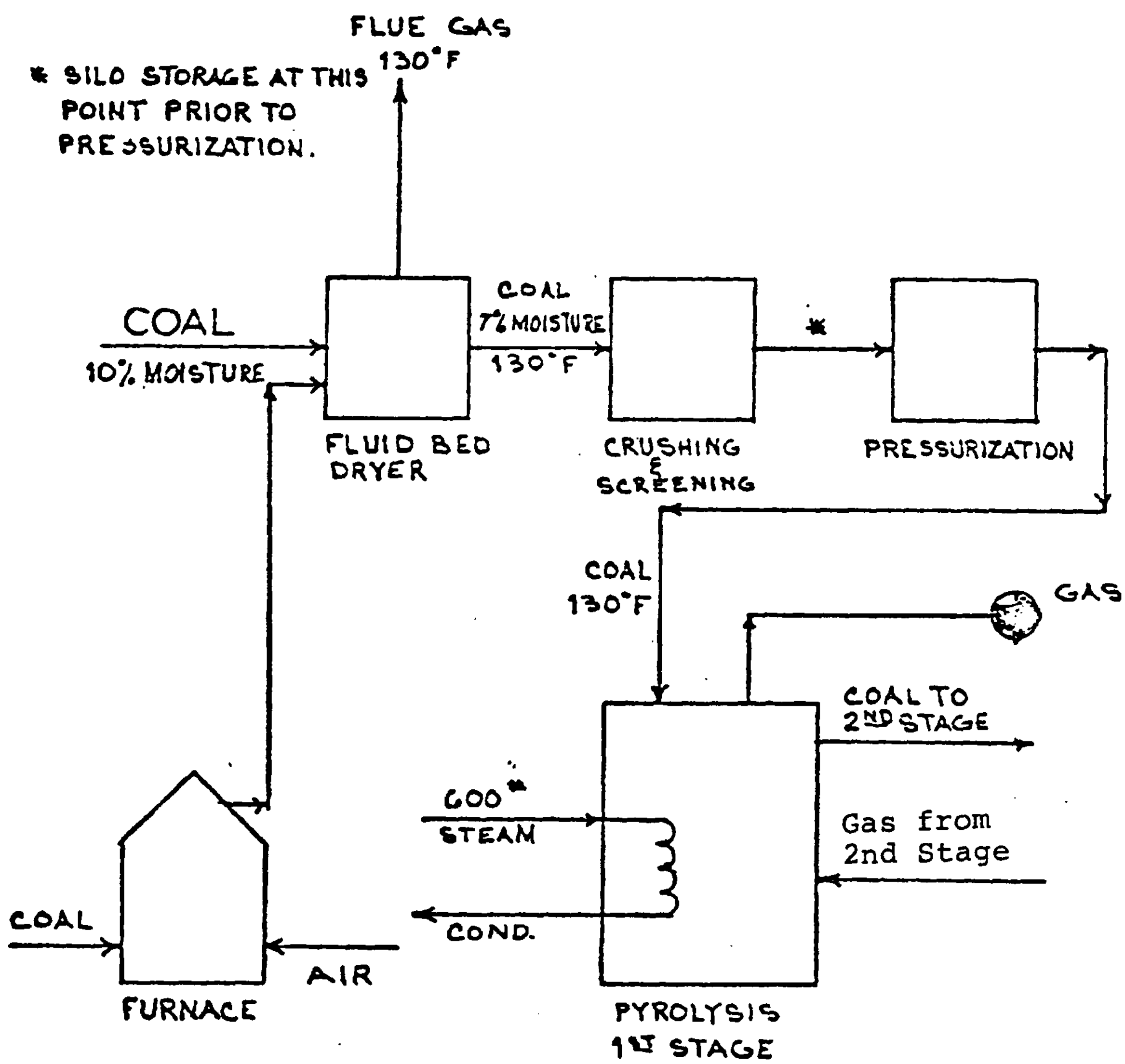
For the process shown in Figure 3.1.2-1, Illinois Coal (corresponding to the TBD) is dried to 7% moisture in a fluid bed dryer with flue gas derived from a coal-fired dryer furnace. The coal has its moisture content reduced from 10% to 7% and leaves the dryer at 130°F, a temperature low enough to permit exposure to the atmosphere without excessive oxidation. This level of drying is sufficient to insure that the coal is free-flowing, but requires the bulk of the water

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Figure 3.1.2-1

Illinois Coal-TBD Drying to 7% Moisture



3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Vendor Responses

Process Description

Illinois Coal - continued

to be removed in pyrolysis. The coal is then crushed and screened, pressurized and introduced to the first pyrolysis stage where the remaining moisture is removed and the coal is heated to 350°F. Heat is supplied from steam coils and hot gas from the second stage.

The process shown in Figure 3.1.2-2 is for Illinois coal dried to 3% moisture, a case in which the bulk of the drying is done ahead of pyrolysis. In order to achieve this low moisture level, the coal must be crushed and screened prior to drying. For drying the smaller sized coal, a flash dryer is used where the heat is supplied by flue gas at 850°F from Area 110. The coal leaves the dryer at 200°F and 3% moisture and requires inerting and protection from the atmosphere to prevent oxidation. It is stored, pressurized and introduced into the first pyrolysis stage. The remaining moisture is removed and the coal is heated by condensing steam in tubes in the pyrolysis vessel. The pyrolysis steam usage in this case is less than for the 7% drying case, but the flue gas used for drying is not available for raising steam in Area 110.

Figure 3.1.2-3 depicts the flow sheet for the non-drying case. Coal, as received from Area 101, is crushed, screened, pressurized and introduced directly into the first pyrolysis stage. Steam condensing in tubes in the first pyrolysis stage provides the main source of heat for removing moisture and preheating the coal. This case uses the maximum amount of pyrolysis steam, but requires no external coal firing or 850°F flue gas, and the entire Area 102 drying system is eliminated.

Western Coal

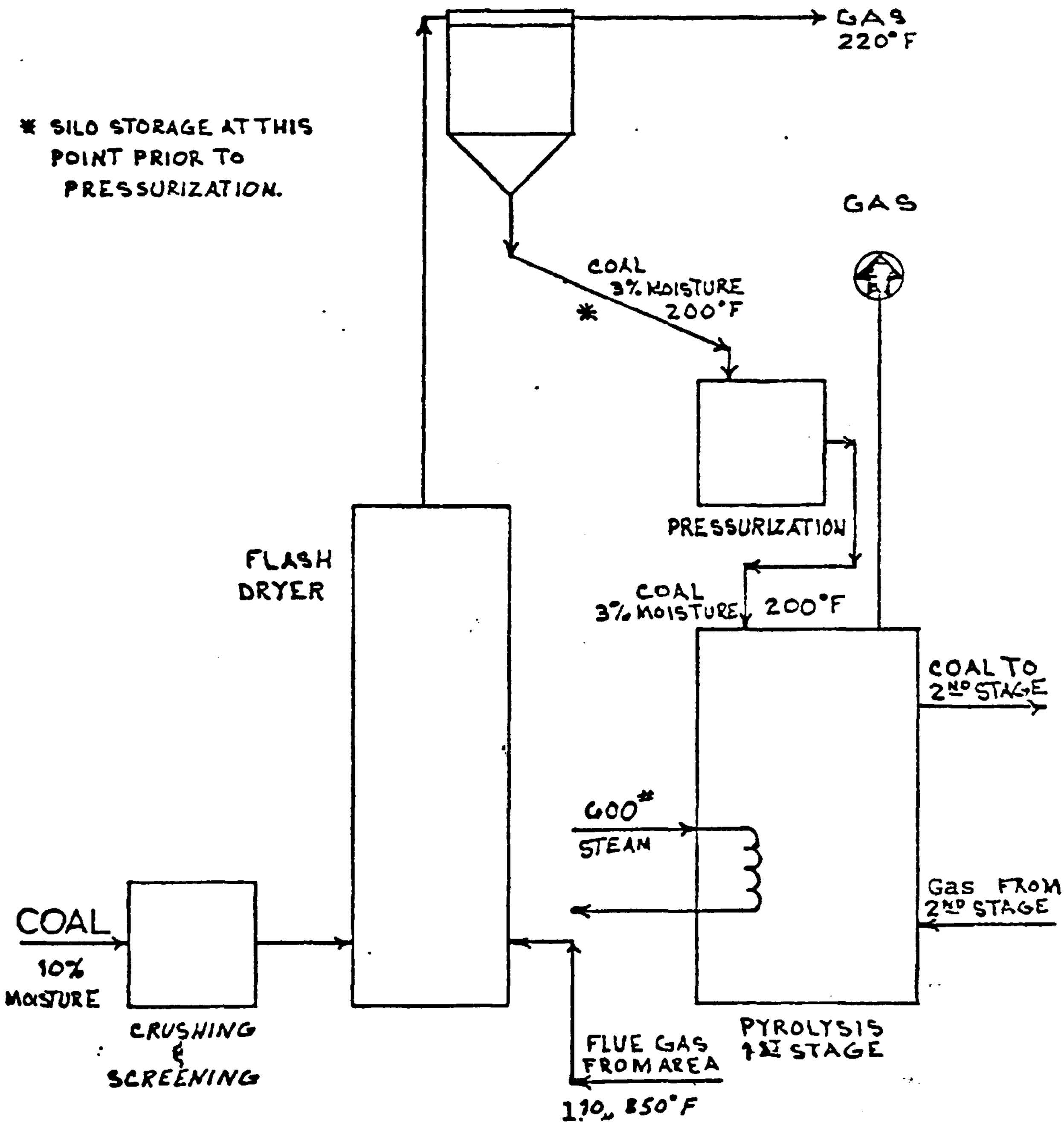
Figure 3.1.2-4 for the no-drying case of Western coal shows a process similar to the no-drying case for Illinois coal. The primary difference is the much higher moisture level of the Western coal (25%) compared to the Illinois coal (10%), which puts an extraordinary heat load on the steam tubes in the first pyrolysis stage.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Figure 3.1.2-2

Illinois Coal Flash Drying to 3% Moisture

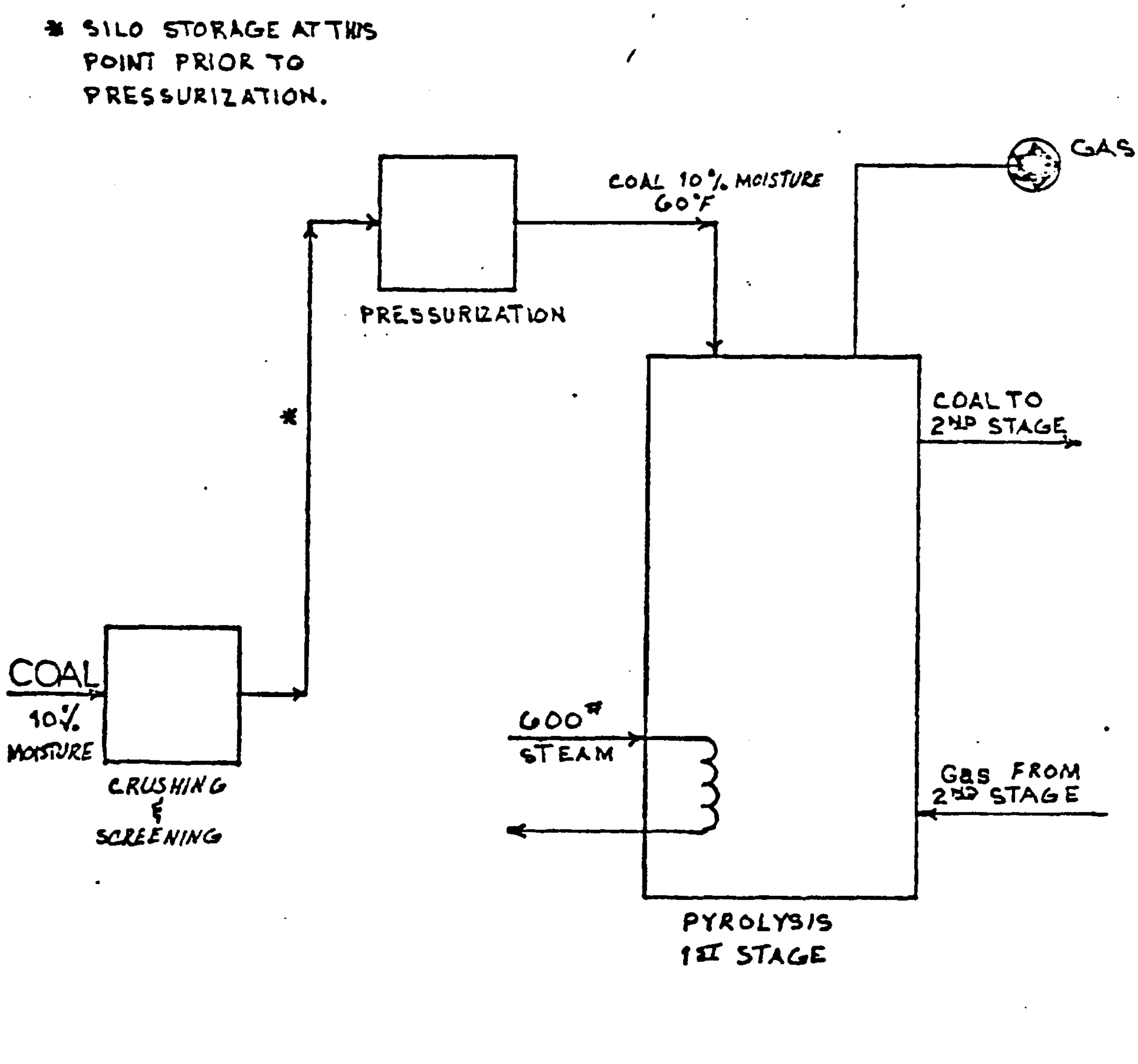


3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Figure 3.1.2-3

Illinois Coal No Drying

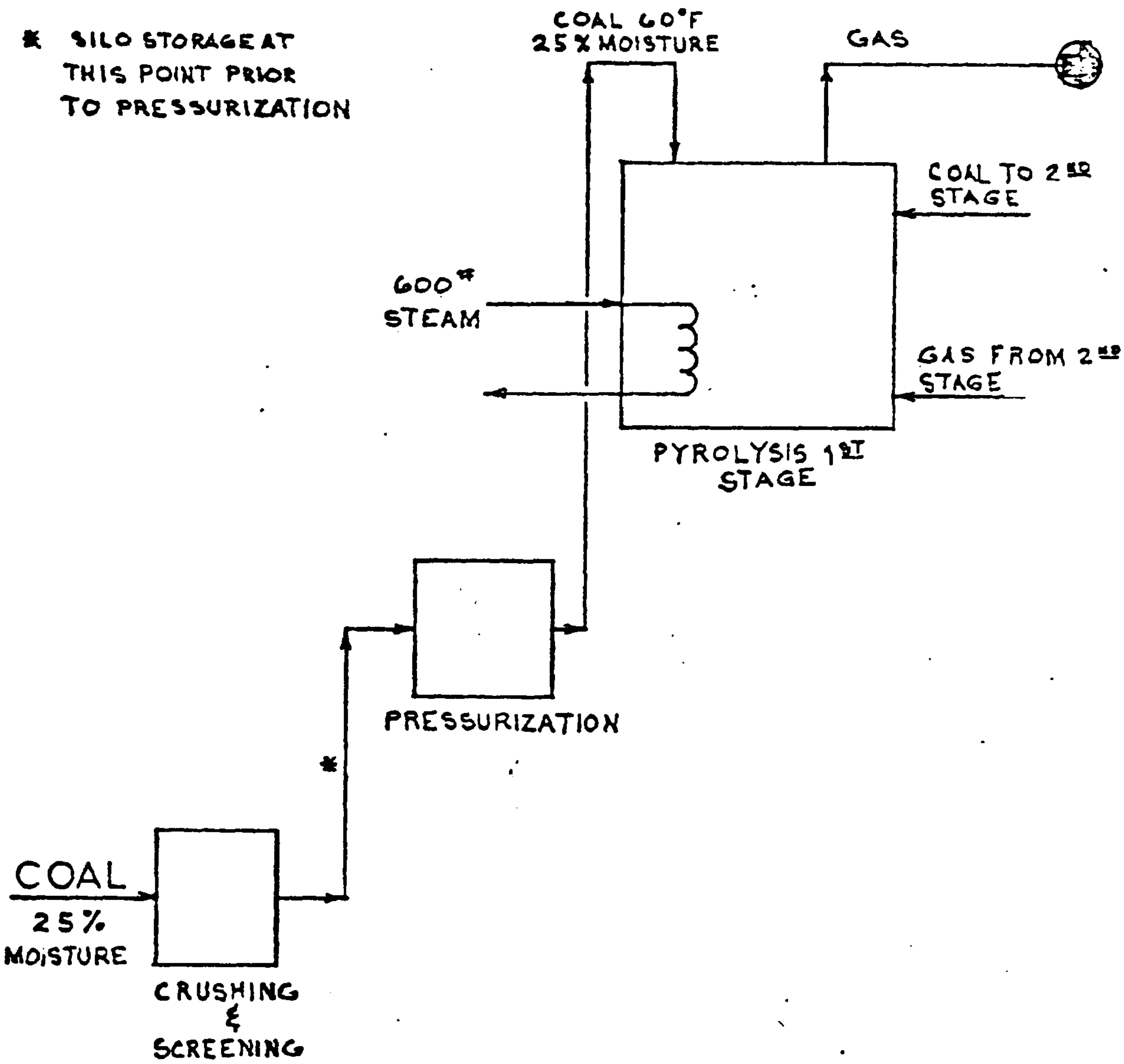


3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Figure 3.1.2-4

Western Coal No Drying Case



3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Vendor Responses

Process Description

Western Coal - continued

In Figure 3.1.2-5, the flow sheet for drying Western coal to 10% moisture is shown. This case would be similar to the no-drying case for Illinois coal in its effect on pyrolysis drying load. The coal is crushed and screened and introduced into the base of a flash dryer. Heat for moisture removal is supplied by flue gas at 850°F from Area 110. The coal leaves the top of the flash dryer at 200°F, is separated from the gas and dropped into an enclosed and inerted storage silo. It is then pressurized and introduced into the first stage of pyrolysis where the rest of the moisture is removed by steam condensing in tubes in the pyrolysis vessel. This process uses less pyrolysis steam than the no-drying process, but the flue gas used for drying is not available for raising steam in Area 110.

The process for drying Western coal to 3% moisture is shown in Figure 3.1.2-6. Crushed and screened coal is introduced into the flash dryer. In order to supply enough heat to remove the moisture without causing the gas flow to become excessively high, the 850°F flue gas is augmented by the gas from a coal-fired furnace such that the mixed gas enters the dryer at 1600°F. Thereafter, the dried coal leaves the dryer at 200°F, is separated from the gas, dropped into a storage silo from where it is pressurized and introduced into pyrolysis. This process puts the smallest heat load on the pyrolysis system but requires the external combustion of coal to supply drying heat.

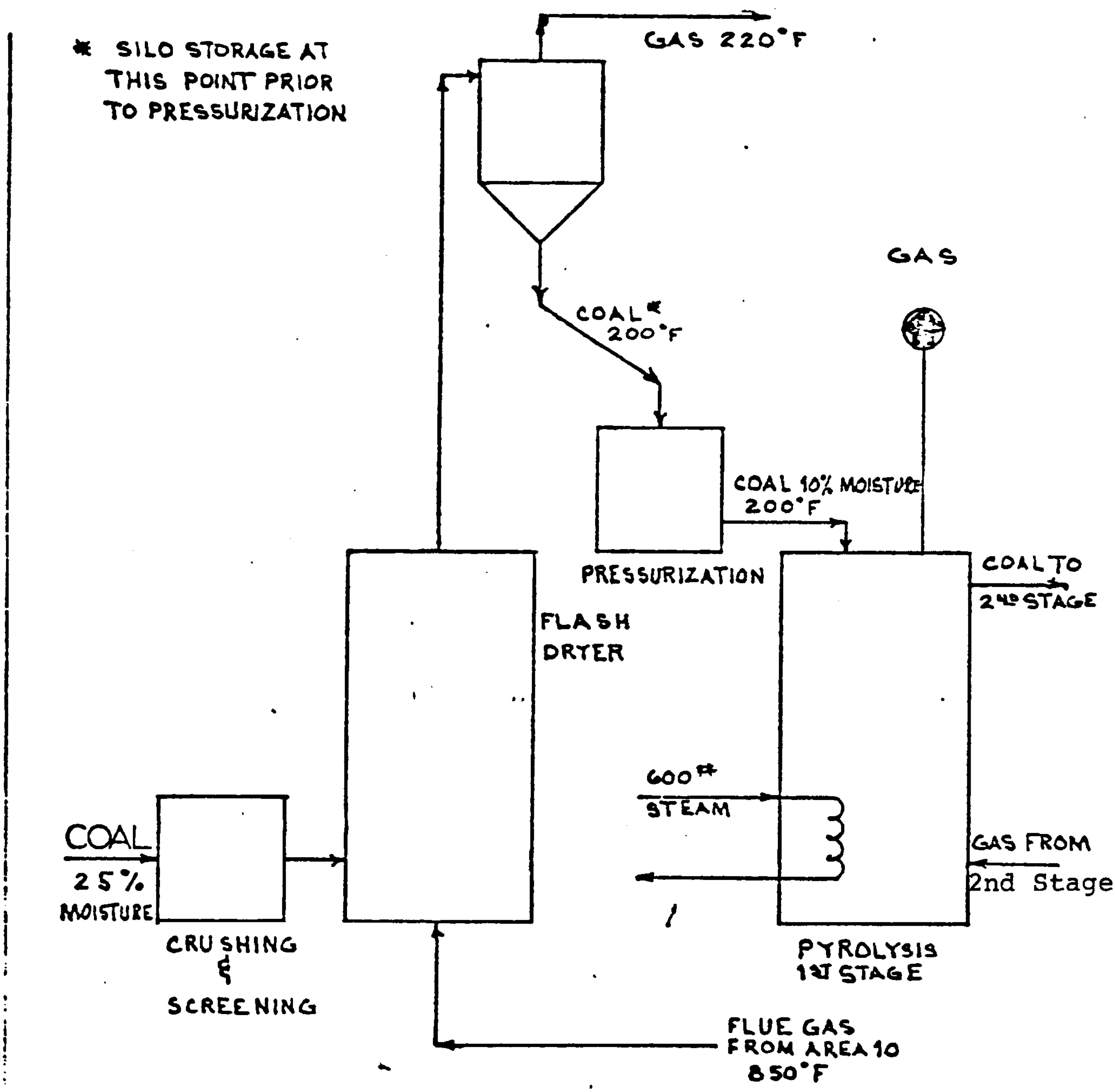
Pittsburgh coal is not dried, since at 4% moisture it already is dry. For all coals, and for all cases, an auxiliary coal-fired fluid bed dryer would be desirable to dry coal prior to crushing and screening. This dryer is not necessary for the nominal design conditions, but is to be used when the as received coal has excess surface moisture, or frozen moisture. The auxiliary dryer will take care of the excess heat load and remove excess surface moisture to insure that the coal is free-flowing to the extent necessary for crushing, screening and pressurizing.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Figure 3.1.2-5

Western Coal
Drying with Flue Gas to 10% Moisture

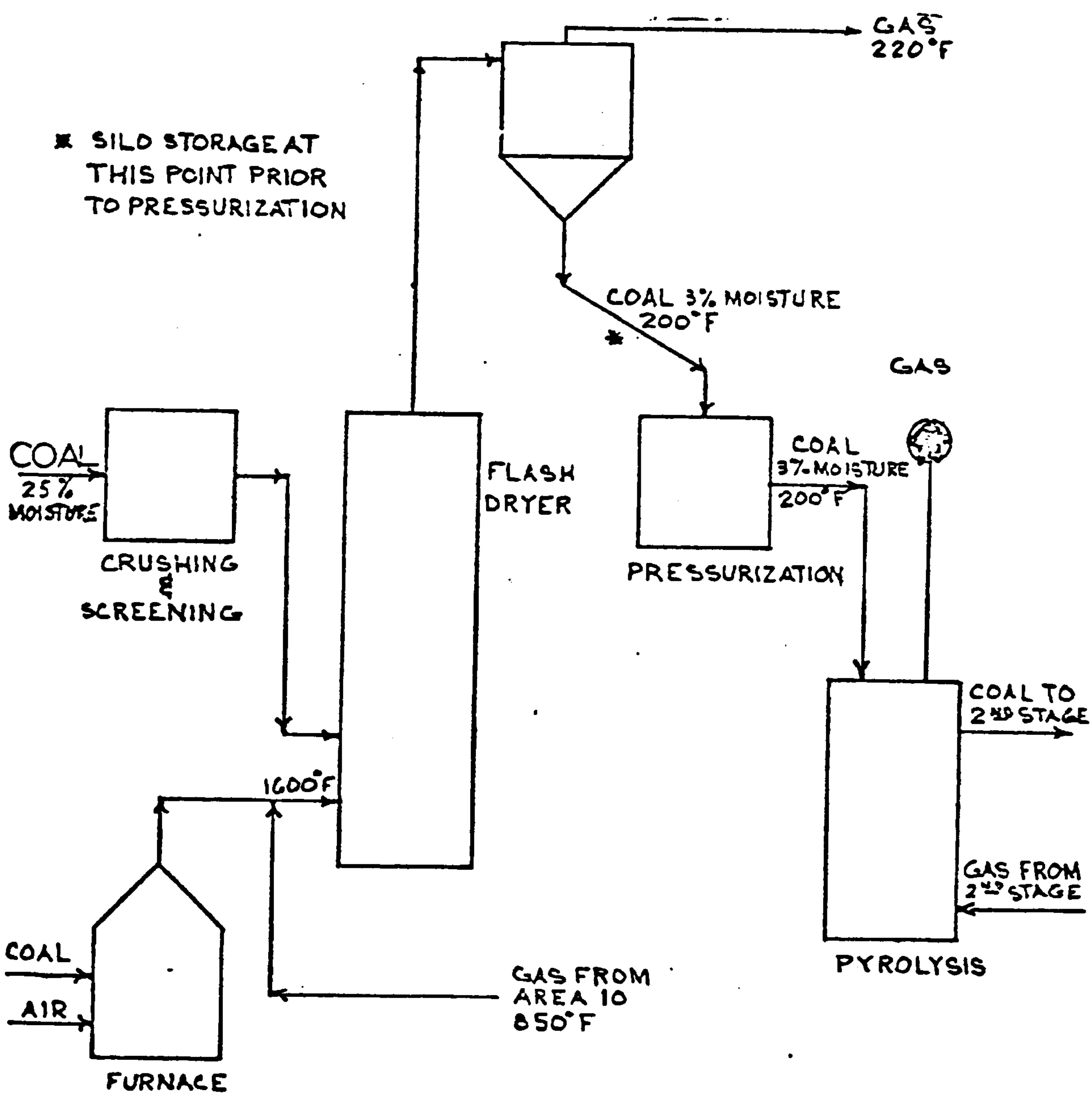


3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Figure 3.1.2-6

Western Coal
Drying with Flue Gas and Coal Firing to 3% Moisture



3.1 AREA 102 - COAL PREPARATION

3.1.2 - Dryer Design Study

Vendor Responses

Process Description - continued

Control Scheme

The details of the control scheme have not been considered as they do not have a significant impact on the dryer trade-off study. However, the major control criteria with flash drying is to maintain the gas flow rate above the minimum required for lifting the largest particle out of the dryer. This must be done even when changing conditions of the coal feed dictate that less heat should be provided. If, for example, there is a decrease in the moisture content or feed rate of the incoming coal, drying with the same amount of hot gas will cause overheating of the coal. Less heat must be provided at a constant volumetric gas flow rate. There are a number of ways to accomplish this:

The gas may be cooled by raising steam.
More flue gas at 850°F, and less high temperature gas can be used.
A third, cool, inert gas may be blended into the drying gas.
An additional heat load may be imposed on the system (i.e. water sprays) to absorb the excess heat in the flue gas.
Common practice is to recycle the dryer flue gas to cool the drying gas.

Similar control problems exist in fluid bed dryers, as there is a relationship among volume of gas, heat content and residence time. However, the residence time and gas velocity are usually not as critical as in the case of flash dryers, and control is usually accomplished by turning down the dryer furnace or limiting the hot gas source.

Process Flexibility

The flash drying system is extremely flexible. Generally, flash dryers may be operated easily at 150% of rated capacity if enough hot gas is available. Most of the drying is done in the first few feet of the column and there is sufficient height so that particles moving at higher than design velocities have an adequate time to dry. Turndown of flash dryers is usually limited to

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Vendor Responses

Process Flexibility - continued

10%, but by allowing for the recycle of flue gas, essentially an infinite turndown can be achieved without danger of overheating the coal.

Equipment Description

Three identical trains are provided. The equipment for each train is described below:

Flash Dryer

The flash dryer tube is 11' in diameter and 40' high. The primary separator is 35' in diameter and the overall structure height is 75'. This is the dryer size for drying Western coal and Illinois coal to 3% moisture. A slightly larger diameter is required for drying Western coal to 10% moisture.

Fluid Bed Dryer

The auxiliary fluid bed dryer has a 10' x 18' cross section.

Screens

The screen capacity with recycle is approximately 500 TPH per train. Nine 4' x 15'-2" surface Hummer screens are required. These will have dust covers and explosion proof motors.

Crushers

The coal crusher will be an impact mill, having a rated capacity of 600 TPH.

Commercial Experience

The Parry Entrainment Dryer, has been used as a model for the Flash drying alternative. The first commercial

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Evaluation Criteria and Process Impact - continued

which is taken into account in the study. In the Tentative Baseline Design, coal is heated to 130°F and dried to 7% moisture in the dryer. The remaining moisture removal and additional heating of the coal is done in the first stage of pyrolysis. The heat required is supplied by condensing 600 psig steam in coils in the first pyrolysis vessels. The 600 psig steam is produced from flue gas in a waste heat boiler, after the expander in Area 110. Heat not required for raising 600 psig steam is used to heat boiler feed water for making 1500 psig superheated steam. Thus, by changing the amount of drying done in the coal dryer, the quantity of 600 psig steam and also the amount of 1500 psig steam is changed.

The 600 psig steam which condenses in the pyrolysis vessel heating coils is removed and flashed to produce 50 psig steam for feed water preheating and deaeration. This decreases the amount of low pressure steam required from back pressure turbines and more power can be produced in condensing turbines, decreasing the requirements for 1500 psig superheated steam. In addition, if more or less flue gas from the expander is used to dry coal, then this also affects the amount of boiler feed water and 1500 psig steam that can be produced. Thus, while flue gas may be used as a source of heat for drying coal, instead of firing coal, this implies that less 1500 psig steam can be produced and some coal must be combusted in an auxiliary boiler to produce the required amount of 1500 psig steam.

In carrying out the design studies, all of these various impacts have to be considered both from a fuel efficiency and capital cost standpoint.

Generally, the thermal effects have been expressed as equivalent tons/day of coal. If coal has to be combusted in an auxiliary boiler to raise steam, it is assumed to be combusted at a 75% efficiency. Aside from coal usage, other direct operating costs are negligible, or are negligibly different for alternative processes. Indirect

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Evaluation Criteria and Process Impact - continued

costs are estimated to be 25% of the total fixed capital, which in turn is estimated to be 2.25 times the equipment cost. Aside from the cost of the dryer system, capital costs for other areas which are impacted are also included. These include waste heat boilers for 1500 psig and 600 psig steam, boiler feed water heaters, heating coils for the pyrolysis vessel, the pyrolysis vessel if it changes in size, and the size change in the auxiliary boiler for 1500 psig steam.

One impacted area that was not considered was in liquid waste treatment. In the no-drying case, water removed with coal leaves the system with the gases from the first pyrolysis stage, goes through the oil ("glop") recovery system and then to the liquid waste disposal system. In the drying case, most of the water removed leaves the system with the dryer flue gas. Since the flue gas is sent to the SO_2 removal system, the water contained in this gas is also eventually condensed and is routed to the liquid waste removal system. Thus, there will either be an additional load on the oil recovery system or flue gas cleanup system, depending on which drying system is selected. This effect was not investigated based on the assumption that the increased cost of each system is approximately equal.

The total installed cost thus reflects all of the potentially variable capital items associated with the alternative and not just the drying system.

Selection of Dryer Type

A secondary consideration in this study was the selection of the appropriate drying equipment to suit the alternate drying process being evaluated. All contacts with vendors were related to exploration of flash drying systems and comparison against fluid bed dryers previously proposed.

While a fluid bed dryer may be used to remove inherent moisture as well as surface moisture, those cases which remove inherent moisture require that the coal be crushed

FE-2012-096
MAY 1980

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Selection of Dryer Type - continued

prior to drying.[1,4] In drying crushed coal, the fluid bed dryers operate in the dilute phase fluidization mode and can accomplish drying to the same extent as flash dryers; however, for this application flash dryers are less costly and thus preferred over fluid bed dryers.

Generally, flash dryers will be used on coal of smaller sizes than is the case for fluid bed dryers and they are favored when drying is to be done at low moisture levels.

Table 3.1.2-2 is a summary of budget estimates received by dryer vendors. These were received at different times and reflect different operating conditions and different coals. In carrying out the trade-off studies, the costs were adjusted to the same operating conditions for each coal by scaling according to the heat load. Some estimates included auxiliary equipment and all costs were put on a common basis for comparison purposes.

Results for Illinois Coal

Three alternatives have been considered for Illinois coal, including the Base Case which is that used in the TBD. The other two cases are to not dry at all, and to dry to 3% moisture in a flash dryer using flue gas as the heat source.

TBD (Figure 3.1.2-1)

The scheme presented in the TBD is not viable as presented. The FMC Materials Handling System Division has indicated that in order to remove moisture to 7%, a much longer residence time than the usual 10 minutes in a fluid bed dryer would be required when allowing the coal to heat up to only 130°F. They stated that the coal should be heated to a higher temperature and then cooled, or that rotary dryers be used to provide a longer residence time. However, since the other alternatives are being compared to the TBD, the TBD case has been retained even though it is unrealistic. In making the economic evaluation, it has been assumed

3.0 PROCESS DESIGN STUDIES

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Table 3.1.2-2

Drying Systems Considered
(In Year End 1977 Dollars)

Vendor	A	B	C	D	E
Coal	Western	Western	Western	Western	Illinois
Dryer type	Flash	Flash	Fluid bed	Flash	Flash
TPD	9600	9600	10000	2400	3000
% moisture Feed	28	25	23	25	10
Product	3	3	2.5	10	3
Inlet gas temp.	1600	1100	860	860	
Outlet gas temp.	220	240	290	225	
Outlet coal temp.	200	190	290	190	
Evaporation mlb/hr	206	181.4	175.2	33.3	18.04
System includes	Dryer, furnace, coal crushers, screens, conveyors, cyclones, blower controls, erection supports	Dryer, furnace cyclone, bag filter, feeder, hopper controls, motors, pneumatic separation	Dryer, fan stack dust collectors & duct handling controls erection	Dryer, fan crusher, feeder ducts	Feed, hopper feeder, dryer, furnace, ducts, cyclones, blowers, dust collection controls, pulverizer
Does not include	Dust collection		Coal conveyors, crushers, screens motors	Dust collection, supports, motors, screens	
Budget estimate	$2,076,000	$5,400,000	$8,000,000	$1,000,000	$3,500,000
Scaled to evaporation capacity	$2,076,000	$6,130,000	$8,800,000	$2,960,000	$15,100,000
Adjusted for common equipment*	$2,076,000	$5,800,000*	$7,800,000	$2,960,000	$11,000,000

- Common equipment are those items included with the flash dryer system A.
- Estimate of Vendor B estimate is high because 4 dryers are required per train due to capacity limitations.
- It is not clear why the estimate of Vendor E is so high. They appear to be oversized for drying Illinois coal.

COGAS 3/A-24

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Results for Illinois Coal

TBD (Figure 3 1.2-1) - continued

that the drying could be accomplished as specified in the TBD. The actual economics would be worse than calculated since more costly drying equipment is actually required.

In the TBD, the total amount of superheated steam produced in the Flue Gas Oxidizer, per train, was 1,023,556 lb/h; and 4,443 lb/h (53.3 TPD) of coal was combusted in the dryer furnace.

Flue Gas Drying to 3% Moisture (Figure 3.1.2-2)

For drying to 3% moisture using flue gas, and heating the coal to 200°F, the heat load on the first pyrolysis stage is greatly reduced. However, a portion of the flue gas, 536,992 lb/h at 850°F, is diverted to coal drying and is not available for heating 1500 psig boiler feedwater and steam. The total coal requirement to provide the needed steam is 46 TPD for drying with flue gas. However, the 53.3 TPD of coal used in the TBD dryer furnace are not longer required so there is a net savings of coal of 7.3 TPD for a single train or a total of 21.9 TPD for the full plant.

No Drying (Figure 3.1.2-3)

For the case where no drying is done, 168,280 lb/h of 600 psig steam are required for the steam coils in the first stage of pyrolysis to supply the additional heat load. Since the incremental amount of steam required is produced from flue gas, there is less heat available to make 1,500 psig boiler feed water and steam. In order to raise enough 1,500 psig steam in an auxiliary boiler to make up the difference between the TBD and no-drying case (38,949 lb/h) 6,195 lb/h (74.3 TPD) of coal have to be combusted at 75% efficiency. However, the condensate from the medium pressure steam is greater in the non-drying

3.1 AREA 102- COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Results for Illinois Coal - continued

No Drying (Figure 3.1.2-3) - continued

case by 48,075 lb/h and the credit for the sensible heat is equivalent to 1,364 lb/h of coal (16.4 TPD). Thus, there is a net requirement for the no-drying case of 57.9 TPD of coal. However, 53.3 TPD of coal had to be combusted in the dryer furnace in the TBD, so the incremental coal usage is 4.6 TPD. All of the figures quoted above are for a single train of a commercial plant. The net coal usage for the entire plant is then 13.8 TPD additional for the no-drying case, as compared to the TBD.

Comparison of Alternates

The net coal usage is only one part of the study. The impact on the different configurations on capital costs were considered. The impacted areas are the drying system as well as the heating tubes in the pyrolysis reactor, the steam waste heat boiler, 1,500 psig feed water heater, and the 1,500 psig auxiliary boiler. The heat recovery system is completely changed. In addition to changing the amounts of steam and boiler feedwater produced, the temperature of the flue gas also changes in the three cases considered. So the amount of heat transfer area required for each application is changed due to a change in load as well as a change in log mean temperature differences. In the Flue Gas Oxidizer, the total load is constant, although there is a slight change in the temperature difference. This has been ignored,as it was not possible to do a complete analysis of the Flue Gas Oxidizer. The process parameters and equipment coast of the impacted areas are summarized in Table 3.1.2-3. These are not complete system costs for the dryer and for heat recovery, but only reflect the specific areas impacted.

3.0 PROCESS DESIGN STUDIES

FE-2012-096
May 1980

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Results for Western Subbituminous Coal

Three alternatives have been considered for the Western subbituminous coal. The coal is assumed to have a total moisture content of 25%. In one case, it is assumed to be flash dried to 10% moisture using flue gas. In another case it is flash dried to 3% moisture using flue gas augmented by a high temperature gas from coal combustion to bring the drying gas to 1600°F, and the third case is not to dry the coal at all.

It is assumed that flue gas is available for drying at 865°F, and that the pounds of flue gas available per pound of MAF coal is the same as for the Illinois coal as in the TBD. This assumption is not crucial, but only provides a simple reference point for establishing the heat recovery relationship in the flue gas recovery area. The process parameters and equipment costs of the three cases are summarized in Table 3.1.2-4 for the Western subbituminous coal. The methodology is the same as described for the Illinois coal, and the numerical values are not itemized here. A heat balance around the dryer is established to determine flue gas and coal requirements. The volume of gas required sets the dryer size. Based on the temperature and moisture content of the coal sent to pyrolysis, the additional steam to pyrolysis is calculated. This 600 psig steam is produced

3.0 PROCESS DESIGN STUDIES

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Table 3.1.2-3
Drying of Illinois Coal - Single Train
(8620 TPD dry coal)

		TBD	Flash Dryer	No Drying
1.	Type of drying	TBD	Flash Dryer	No Drying
2.	Dried coal temperature, °F	130	200	60
3.	Moisture of coal to pyrolysis WT%	7	3	10
4.	Flue gas required for drying, lb/h	0	536,992	0
5.	Coal required for drying, lb/h	4443	0	0
6.	Medium pressure steam required for pyrolysis, lb/h.	Base	-76.925	+48.075
7.	1500 psig steam produced in Flue Gas Oxidizer, lb/h	903,351	894,351	864,582
8.	Coal required for 1500 psig steam in auxiliary boiler, lb/h	0	1438	6195
9.	Coal equivalent of condensate, lb/hr	0	2398	(1364)
10.	Net coal requirement, lb/h	4443	3836	4831
11.	Equipment Costs - M$			
	a. Coal dryer	500	1020	-
	b. Dryer furnace	250	-	-
	c. Bag collector	1800	1800	-
	d. Cyclones, blower & other equipment associated with dryer	843	Included with dryer	-
	e. Steam tubes in pyrolysis vessel	77.7	23	112
	f. 600 psig steam waste heat boiler	126	41.3	191.3
	g. 1500 psig BFW heater	59.7	40.3	77.7
	h. 1500 psig auxiliary boiler	1,110	1140	1239.3
	i. Total equipment	3004.3	2864.7	1621

See notes on next page.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Notes to Table 3.1.2-3

Line

1. The TBD uses a fluid bed dryer. This is unrealistic, but in this case is for cost comparison.

2, 3 These moisture levels and temperatures are mutually attainable.

4. Based on dryer heat balance.

5. This is the amount of coal required by a dryer heat balance.

6. This steam provides the heat to remove the remaining water in pyrolysis and to heat the coal to 350°F.

7. After the amount of flue gas used for drying coal and raising 600 psig steam is established, the amount of 1500 psig raised in the Flue Gas Oxidizer can be calculated.

8. It is assumed that steam can be raised from coal with a 75.0% efficiency. The steam is raised in an auxiliary boiler such that the amount 1500 psig steam produced is the same for all cases.

9. This is a credit for the sensible heat in the 600 psig condensate, which is flashed to low pressure steam. The numbers represent the increment compared to the TBD. In the no-drying case their coal usage represents a credit.

11 a-d No furnace is required when flue gas is used for drying. For the no-drying case, the other equipment associated with the drying system is not required.

e The tube surface reqûired is based on the 600 psig steam required for heating.

f,g The costs for the waste heat boiler and feed water heater are adjusted for heat duty as well as log mean temperature difference.

h The auxiliary boiler is a single unit and the cost per train is taken as 1/3 of the total cost.

11. Dryer system costs are based on vendor proposals, other equipment costs are scaled from ICGG Proposal Vol. 5.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Table 3.1.2-4

Drying of Western Coal - Single Train of a Commercial Plant (8808 TPD of Dry Coal)

	No Drying	Drying with Flue Gas	Drying with Flue Gas & Coal Firing
1. Type of Drying			
2. Dried coal temperature, °F	60	200	200
3. Moisture of coal to pyrolysis wt.%	25.5	10	3
4. Flue gas required for drying lb/hr	-	1,359,000	631,000
5. Coal for drying (dry) lb/hr	-	-	18,087
6. M.P. steam required in pyrolysis lbs/hr	504,258	166,071	81,318
7. 1500 psig steam produced lbs/hr	663,800	686,400	865,230
8. Additional 600 psig steam required from auxiliary boiler lb/hr	82,000	-	-
9. Coal required for steam dry lbs/hr	38,531	25,461	0
10. Coal equivalent of condensate lb/hr	15,750	5,167	2,500
11. Net coal required (dry) lb/hr	22,781	19,994	15,587
TPD	273	240	187
12. Equipment costs - M$			
a. Auxiliary boiler	9433	5967	0
b. Steam tubes in pyrolysis	416.0	137	67
c. 1500lb steam	2608	2702	3,000
d. M.P. steam WHB	3640	893.3	222.
e. BFW heating	0	32.7	114.3
f. Drying system	0	1087	1020
g. Dryer furnace	0	0	556
h. Pyrolysis vessel	1298	1043.3	1043.3
i. Bag filter for drying system	0	600	600
j. Total equipment	8905.3	7092	6656

See notes on next page.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Notes to Table 3.1.2-4

1. In order to dry to 3% moisture, it is necessary to use a higher temperature in the drying gas than for drying to 10% moisture. By augmenting the 850°F flue gas with coal firing, a drying gas temperature of 1600°F is achieved.

2,3 These moisture levels and temperatures represent typical values achieved in drying Western subbituminous coals under the conditions indicated.

4. Based on dryer heat balance.

5. Coal required to bring drying gas temperature to 1600°F.

6. This is the heat required to evaporate water remaining in coal and to heat coal and water to 350°F.

7. After the flue gas used for drying coal and making 600 psig steam is calculated, the amount of 1500 psig steam that can be raised is calculated.

8. When the coal is not dried, the pyrolysis heat load is so large that there is not enough heat in the flue gas to make all of the 600 psig steam required, and some must be raised in an auxiliary boiler.

9. It is assumed that steam can be raised from coal with an efficiency of 75%. The steam is raised in an auxiliary boiler such that the 1500 psig steam for all three cases is the same.

10. This is a credit for the sensible heat in 600 psig condensate, which can be flashed to low pressure steam. Again, it is assumed that the conversion of coal to steam is 75% efficient.

11. a. The cost for the auxiliary boiler is an incremental cost above the cost for the 3% drying case, and includes the capacity required for the additional 600 psig steam in the no-drying case. Additionally, 1/3 of the cost is assigned to each train.

    b. The tube surface required is based on the 600 psig steam used for heating.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Notes to Table 3.1.2-4 - continued

c. The waste heat boiler for raising 1500 psig steam is part of the Flue Gas Oxidizer. The total heat load is the same in all cases, but there is a greater temperature difference for heat transfer for the no-drying case and the flue gas drying case leading to a reduced heat transfer area.

12. Dryer system costs are based on vendor proposals; other equipment costs are scaled from the ICGG Proposal, Vol. 5.

d,e The costs for the 600 psig WHB and BFW are adjusted for heat load and log mean temperature difference (LMTD). For the no-drying case in the WHB, the log mean temperature difference (136°F) is small compared to the log mean temperature difference (428°F) in the drying to 3% moisture case. This explains the high cost of the WHB. Because of the low LMTD, this may not be the best way to raise steam.

f. The flash dryers for the two drying cases are the same since the 3% moisture case uses a 1600°F gas for drying. For drying to 10% moisture with flue gas only, the diameter of the dryer is slightly increased. .

g. Furnace is required only for drying to 3% moisture.

h. For the no-drying case, the diameter of the pyrolysis vessel has to be increased to accommodate the steam tubes.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Results for Western Subbituminous Coal - continued

from the flue gas. Based on the flue gas required for drying and making steam, the amount of 1500 psig steam raised is calculated. Using the no-drying case as a reference, additional steam is produced in the auxiliary boiler so that the output of 1500 psig steam is the same for all cases.

No Drying (Figure 3.1.2-4)

Crushed coal is fed without drying directly to the first pyrolysis stage. The heat required for moisture removal and coal heating is supplied by 600 psig steam raised from a waste heat boiler in Area 110. Even though no flue gas is used for drying the coal, there is not enough heat available in the Area 110 flue gas to produce all of the 600 psig steam required, and some must be raised in an auxiliary boiler. This alternative eliminates the dryer system but incurs very high costs for raising steam and for providing steam heat in the first pyrolysis stage.

Flue Gas Drying (Figure 3.1.2-5)

In this alternative, use is made of flue gas to dry and heat coal in a flash dryer, prior to pyrolysis. Some of the high costs associated with raising steam and heating coal in pyrolysis are eliminated, but at the added expense of a drying system. Since the moisture level of the coal introduced to pyrolysis is 10%, a significant amount of 600 psig steam is still required.

Flue Gas and Coal Fired Drying

In order to achieve a moisture content of 3% by flash drying, the 850°F flue gas must be augmented by coal firing to bring its temperature up to 1600°F. While the cost of an auxiliary furnace is incurred, the added cost of a larger dryer is eliminated because more heat is available for a given volume of gas. Some coal has to be combusted externally, but the flue gas is no

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Results for Western Subbituminous Coal

Flue Gas and Coal Fired Drying - continued

longer used in producing 600 psig steam and may be used to heat 1500 psig boiler feedwater and make 1500 psig steam.

Comparison Alternatives

As in the cases for Illinois coal, the impact on capital costs was also considered; including the heating tubes in the first pyrolysis stage, waste heat boiler, feedwater heater, auxiliary boiler and Flue Gas Oxidizer. The size of all these units is not the same for all cases, and cost differences are indicated in Table 3.1.2-4.

Cost Comparison

Total Installed Cost

In some cases equipment costs have been established for a system, and in some cases they have been expressed as an incremental cost relative to a reference case. In order to choose among alternate systems, it is the incremental portion of the cost that is significant. Equipment costs, or incremental costs, are multiplied by a factor of 2.25 to get an estimate of the total fixed capital required. These costs for three cases for each of the coals are summarized in Table 3.1.2-5.

Significant cost differences in equipment relate to the basic trade-off of the cost of a direct heat transfer system (fluid bed or flash dryer) compared to an indirect heat transfer system (raising steam in waste heat boilers and condensing steam in tubes). For the Illinois coal, it is generally more advantageous to eliminate the external drying system and incur the cost of indirect heating. For Western coal, where the heat loads are much larger, the direct heating system offers cost advantages. When

3.0 PROCESS DESIGN STUDIES

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Table 3.1.2-5

Equipment Cost Summary
3-Train Commercial Plant Basis

Coal Type	Illinois		
Type of Drying	TBD	Flue gas to 3% moisture	None
Equipment M$	9,013	8,594	4,863
Total Fixed Capital M$	20,279	19,336	10,940

Coal Type	Western		
Type of Drying	None	Flue gas to 10% moisture	Flue gas plus coal to 3% moisture
Equipment M$	26,716	21,276	19,868
Total Fixed Capital M$	60,111	47,871	44,703

Note: Costs for the Western and Illinois coal are not to be compared with each other, as they do not reflect common equipment. The significant comparisons are the incremental costs among the cases considered for each coal type.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Cost Comparison

Total Installed Cost - continued

all of the drying is done in pyrolysis, for example, the cost of waste heat boilers and steam tubes in the first pyrolysis stage becomes excessive. For Illinois coal, the case requiring the least capital is the no-dryir.. case, requiring only about half as much fixed capital ($11,000,000) as for the other two cases ($19,000,000 and $20,000,000) on a 3-train basis.

The least capital required for Western coal is for drying to 3% moisture using flue gas and coal firing. On a 3-train basis, this alternative costs approximately $3,000,000 less than drying to 10% moisture and about $15,000,000 less than the no-drying case.

Operating Cost

The major difference in direct operating expenses among the alternative configurations is in equivalent coal usage. Differences in labor and power for motors, fans, pumps, etc. is judged to be negligibly different among the various alternatives, although operating labor would be less for the no-drying cases. Indirect operating costs are estimated to be 25% of fixed capital as previously calculated. For Illinois coal a cost of $22.50/ton is assumed, and for Western coal a cost of $8/ton is assumed. The annual operating cost is based on 330 days per year of operation, and is summarized in Table 3.1.2-6.

Differences in direct operating costs for Illinois coal are not large. Basically, a certain amount of heat has to be supplied to dry and heat the coal, and cost differences are based on slight differences in how the heat is supplied. When the moisture is removed in pyrolysis, it leaves the system at a higher temperature (360°F) than if it were removed in a fluid bed dryer (140°F) or a flash dryer (220°F). The difference in the sensible heat of the water vapor leaving the system accounts for most of the difference in thermal requirements, and thus for most of the difference

Table 3.1.2-6

Annual Operating Costs
3-Train Commercial Plant Basis

Coal Type	Illinois			Western		
Drying Case	TBD	Flue gas to 3% moisture	No Drying	No Drying	Flue gas to 10% moisture	Flue gas plus coal to 3% moisture
Incremental coal due to drying - T/D	160	138	174	1099	966	753
Tons/year	52,800	45,540	57,420	362,670	318,780	248,490
Cost of coal - MM $/yr	1.188	1.0246	1.2920	2.901	2.550	1.988
Annual fixed costs MM $/yr	5.070	4.834	2.735	15.028	11.968	11.175
Total Annual Costs MM $/yr	6.258	5.859	4.027	17.929	14.523	13.163
Impact on Gas Cost ¢/MM Btu (78441 x 10⁹ Btu/yr)	7.98	7.470	5.13	22.86	18.51	16.78

Note: Costs for Illinois and Western coal should not be compared to each other. The significant comparisons are the incremental costs among the cases considered for each coal type.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation

Cost Comparison

Operating Cost - continued

in equivalent coal usage and direct operating costs. Other variations are due to differences in the temperature of the flue gas leaving the system. In Area 110, flue gas used for heating boiler feedwater leaves at 250°F. Drying will be a more efficient use of heat when flue gas from the dryer leaves the system at temperatures less than 250°F.

For Illinois coal, the direct operating cost is very nearly the same for all cases, but the total operating cost is least for the no-drying case (approximately $4,000,000 per year on a 3-train basis compared to approximately $6,000,000 for each of the other two cases).

For Western coal, drying to 3% moisture leads to the lowest direct operating cost as well as lowest total operating cost (approximately $13 MM/yr compared to $14.5 MM and $18 MM/yr for the other two cases).

Technical Comparison

All of the systems considered (flash drying, fluid bed drying, and drying in pyrolysis) are state-of-the-art and involve little technical risk. Other criteria such as turndown, operating flexibility, and impact on other plant areas are also considered and summarized in the Technical Review Matrix, Table 3.1.2-7. The technical comparison is not done on a case by case basis but on a common equipment basis. Thus, the no-drying system has all of the drying being done in pyrolysis and applies to those cases for Illinois and Western coal. The fluid bed dryer is used only in the TBD. The flash dryer, using flue gas, only applies to drying Illinois coal to 3% moisture and Western coal to 10% moisture. The flue gas and coal fired drying applies to drying Western coal to 3% moisture.

Table 3.1.2-7

Technical Review Matrix

System	No Drying	Fluid Bed Dryer	Flash Dryer Flue Gas Only	Flash Dryer Flue Gas & Coal Fired
Degree of Development	high (1)	high (1)	high (1)	high (1)
Technical Risk	low	low	low	low
Turndown	moderate (2)	high (3)	high (5)	high (5)
Operating Flexibility	moderate (2)	high (3)	high (4)	high (5)
Impact on other plant areas	high (2)	moderate (3)	moderate (4)	low (5)

Notes: (1) All systems are well developed state-of-the-art.

(2) Paradoxically, not drying at all may produce a less flexible system in respect to turndown and operating flexibility, due to the fact that the drying is actually done in pyrolysis by the use of 600 psig steam. Changing the amount of 600 psig steam used will alter the operation of the flue gas heat recovery system and operation of the auxiliary boiler. Therefore, the turndown ratio and operating flexibility depends on other system components.

(3) Fluid bed dryer turndown is essentially based on the turndown of the furnace supplying the hot gas. Usually a 50% turndown can be achieved.

(4) The flash dryer can be infinitely turned down, provided that provision is made for recycling cool gas from the dryer exhaust. This is generally done. The impact on the rest of the plant is that the amount of flue gas required from the expander changes and more heat must be taken from it at some other point.

(5) Operating a flash dryer with flue gas, as well as coal firing, minimizes the impact on other plant areas, as variations can be accommodated by changing the amount of coal firing.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

Technical and Economic Evaluation - continued

Technical Comparison - continued

The categories of turndown, operating flexibility, and impact on other areas are interrelated. Each of the systems considered has operating flexibility when considered by itself. It is the impact on other plant areas which will tend to limit flexibility of a particular system, as each of the systems depends on receiving heat from some other point in the process. When all of the drying is done by condensing steam in the first pyrolysis stage, the heat load must match the 600 psig steam available. A change in heat load will have an impact on the entire power recovery train in Area 110. Similarly, when a portion of the drying heat load is accomplished by direct heating with flue gas, the flue gas available must match the heating demand. There is slightly less impact on other plant areas when using direct flue gas heat since this has less of a direct impact on the operation of the waste heat boilers. When coal firing is used to supply some of the drying heat, this minimizes, but does not eliminate, the impact on other plant areas, since the amount of coal firing can be independently raised.

Based on the technical comparison, there is no compelling reason for selecting one alternative over another, and the recommendation is based on the cost comparison.

Recommendation

Based on the economic comparison, the following recommendations are made: For Illinois coal, the conceptual design indicated in the TBD is changed to eliminate the drying system. This recommendation is based on the fact that not drying coal ahead of pyrolysis will result in an annual saving of approximately $2.2 MM. This is accomplished without increasing the technical risk but with some impact on other plant areas. However, the same impact exists for all alternatives, and its effects can be minimized.

For Western coal the recommended drying alternative is to flash dry the coal to 3% moisture using a combination of flue gas and coal products of combustion. This alternative saves $4.7 MM/yr compared to drying with flue gas only. In addition, the respective capital savings are $15.3 MM and $3.11 MM vs. the no-drying case.

3.1 AREA 102 - COAL PREPARATION

3.1.2 Dryer Design Study

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3.0 PROCESS DESIGN STUDIES

FE-2012-096
MAY 1980

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Summary

The intent of this study was to examine the use of Fuller-Kinyon pumps as an alternative to the lock hopper pressurization presented in the TBD. The TBD lock hopper system was based on a conceptual design and budget estimate received from Babcock and Wilcox. Other lock hopper vendors who had previously been solicited were: Petrocarb and DUCON. Prior to the TBD, the Babcock and Wilcox system was selected as being the preferred lock hopper system.

A conceptual design and budget estimate was received from the Fuller-Kinyon Company for a pressurization system. The trade-off study evaluated the pertinent factors in the lock hopper and Fuller Kinyon pump systems. This evaluation included capital costs, operating costs, commercial experience, reliability, flexibility and system impact.

Differential capital and annual operating costs for both systems are summarized below, for a single train of a commercial plant. All dollars are at year end 1977.

System	Lock Hoppers	Fuller-Kinyon Pumps
Equipment Cost \$M	2,399	578
Total Fixed Capital \$M	5,398	1,300
Capital Related Cost \$M/yr	1,720	390
Direct Operating Cost \$M/yr	278	199
Total Annual Cost \$M/yr	1,898	589
Impact on Gas Price ¢/MM Btu	7.25	2.25

Based on the cost impact, the Fuller-Kinyon pressurization system is the preferred. Other factors such as degree of development, reliability, flexibility and impact on the rest of the process are judged to be equal for the two systems, except in one area. The amount of particle size attrition caused by the Fuller-Kinyon pump may be excessive for the Western subbituminous coal. This may be determined in a test program proposed by Morgantown Energy Research Center, or at the Fuller-Kinyon laboratory in Pennsylvania.

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Summary - continued

For the Commercial Plant, operating on Illinois or Pittsburgh coal, the Fuller-Kinyon system is recommended.

Performance Specification

The pressurization system shall be capable of pressurizing and delivering the coal to the first stage of pyrolysis in order to elevate the coal and to allow for pressure drop during transport, the system should operate at a nominal pressure of 50 psig. In order to allow for extraordinary demands, the system should have the capability of pressurizing to 75 psig. The system should also be capable of operating at 250°F.

Process Description

The lock hopper system is illustrated in Figure 3.1.3-1. Coal is pneumatically conveyed to a surge hopper with a one-hour capacity. Coal is discharged from the surge hopper through a rotary valve and flows by gravity into one of two parallel lock hoppers. Each lock hopper is sized for 20 minutes coal holdup. After the first lock hopper is charged with coal, the top valve is closed and inert gas is fed into the lock hopper until the pressure is raised to about 60 psig. The coal from the lock hopper is then discharged into a fluidizing hopper, and to a conveying line where it is picked up by transport gas. When the lock hopper is empty of coal, a bottom valve closes. The empty, pressurized lock hopper then has its pressure equalized with the second lock hopper, which has previously been charged with coal and is at atmospheric pressure. This is done by opening a valve in an interconnecting line. The cycle is then repeated.

After the pressures are equalized, the first lock hopper is vented through a dust collector and the top valve is opened to receive another batch of coal.

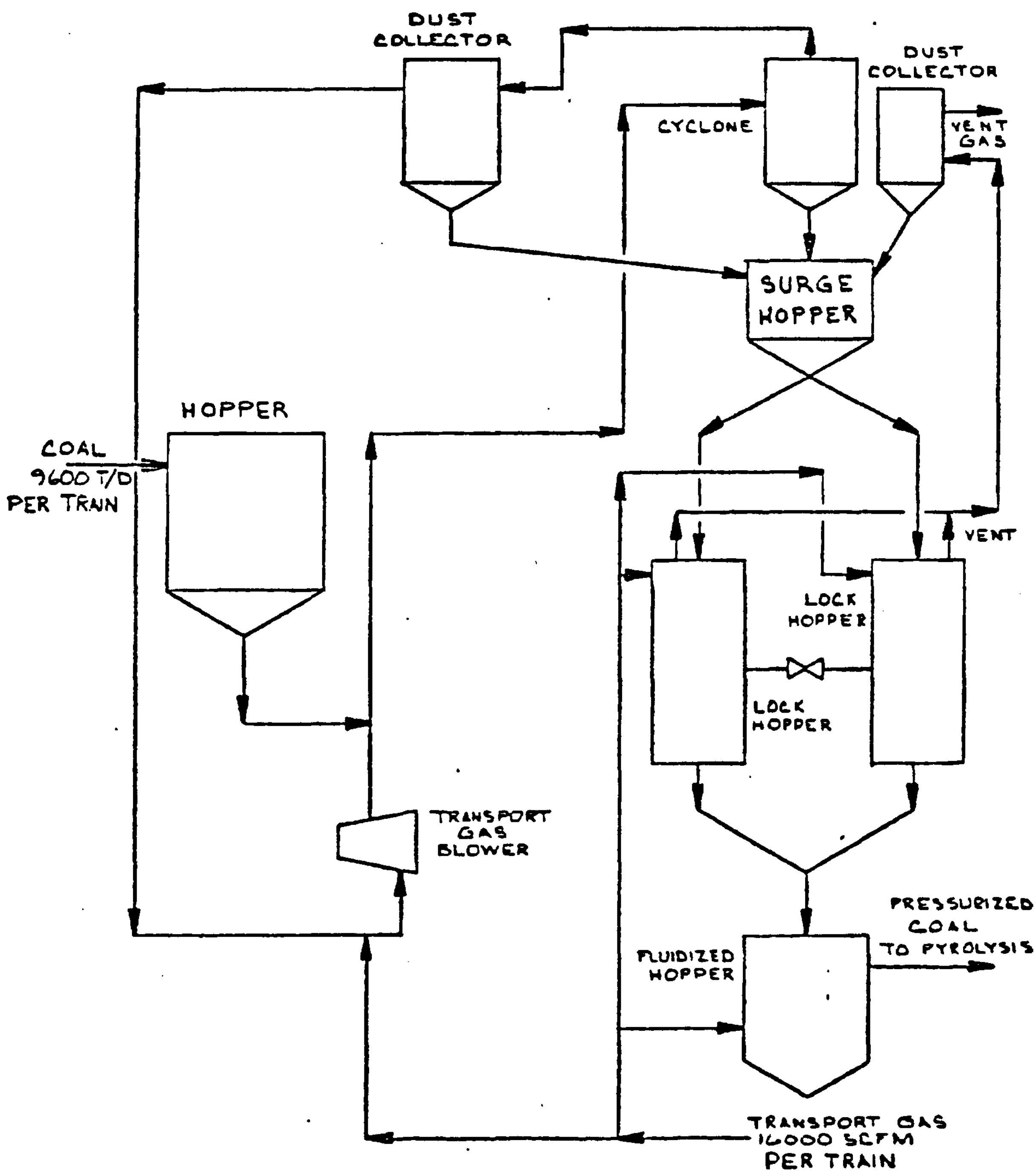
The inert gas used for pressurization and transport is obtained from the Pyrolysis First Stage Gas Scrubbing System, and is available at 160°F and 25 psig. It must be further compressed within the Pressurization System to 60 psig, and cooled to condense water vapor.

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Figure 3.1.3-1

Lock Hopper Pressurization System



3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Process Description - continued

The system using Fuller-Kinyon pumps is illustrated in Figure 3.1.3-2. The coal from the bottom of the storage silo will flow by gravity to the hopper of a Fuller-Kinyon pump. Each train of a commercial plant will require two pumps. The coal is pressurized in the pump to about 50 psig. At the outlet of the pump an inert gas is introduced to disperse the particulate matter and convey it to the first pyrolysis stage. This system eliminates one dilute phase pneumatic lift and substitutes the Fuller-Kinyon pump for an array of lock hoppers and cyclones. The inert gas comes from the same source as in the lock hopper system.

Utility Usage

Utility usage consists of power for the recirculation inert gas compressors in the lock hopper system and Fuller-Kinyon pump system. In addition, the Fuller-Kinyon pumps require power. The lock hopper system for a commercial plant will require 2400 more horsepower than the Fuller-Kinyon pump system.

Process Flexibility

The lock hopper system can readily respond to feed rate changes by going to a longer or shorter cycle time, as required. Additionally, each lock hopper is oversized, so that a larger charge may be introduced at a constant cycle time if an increase in throughput is desired. Overpressures can readily be handled, limited only by the recycle compressor capability.

The Fuller-Kinyon system can also readily respond to system changes. The pumps would be equipped with a variable speed drive to change feed rates as required. The Fuller-Kinyon pumps normally operate at pressures of around 50 psig. Higher pressures can be obtained, and are usually limited by the barrel length which acts as a pressure seal. The finer the material being pumped, the higher the pressure achieved.

3.0 PROCESS DESIGN STUDIES

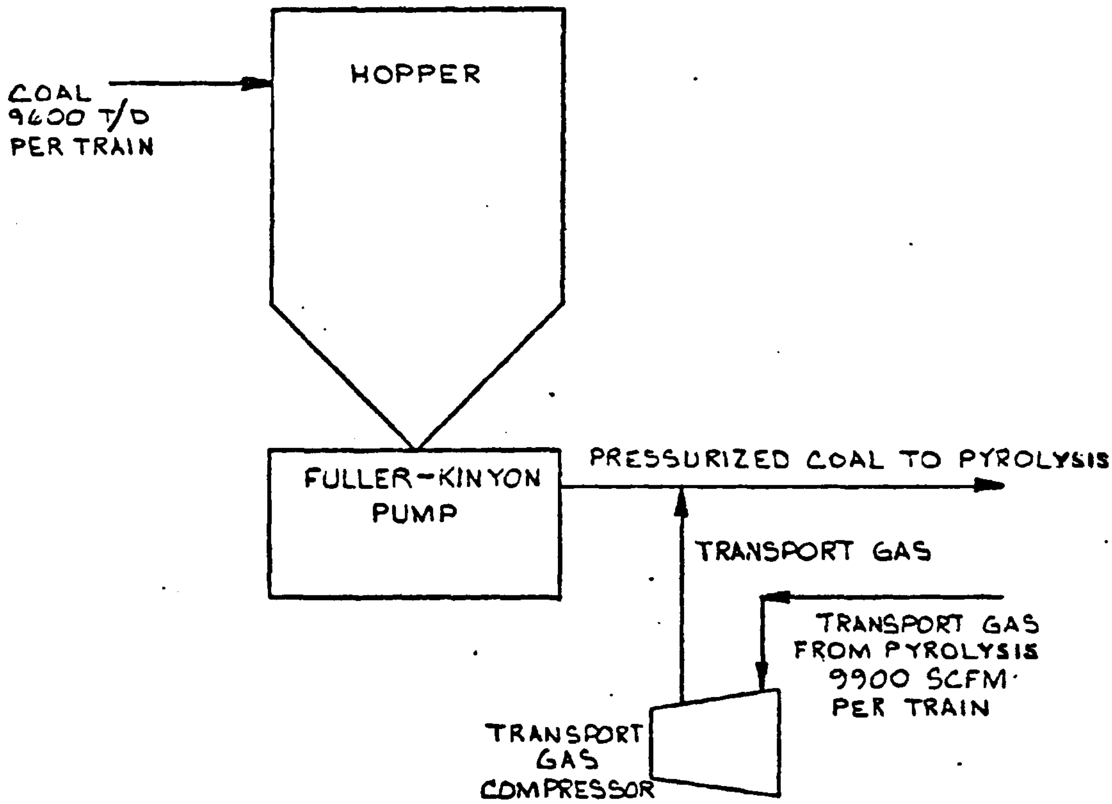
FE-2012-096
MAY 1980

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Figure 3.1.3-2

Fuller-Kinyon Pressurization System



3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Equipment Description

The equipment specified for the lock hopper system is based on the conceptual proposal from a vendor of such systems. The system proposed has a crushed coal hopper with an 8-hour storage capacity, 20 minutes holdup in each of two lock hoppers as a nominal condition, but each lock hopper can handle double the nominal amount. The feed tank also has a 40 minute holdup. The compressor for a single train is nominally designed to handle 12,000 SCFM of gas at a pressure ratio of 1.9. Also included are a bag filterhouse, screw feeder, rotary seal, load cells for the lock hoppers and controls.

For the Fuller-Kinyon system, two 300 MM pumps are specified for each train of the commercial plant, with a spare pump on each train. This could, by means of a remote control diverting valve, immediately be placed into service, in the event that an operating pump has to be taken off line for maintenance or screw replacement. Variable speed drives will be provided. Other equipment included will be a 150 HP explosion-proof motor for each pump, recycle gas compressors with motor and an electric motor operated rotary cut-off valve for bin discharge.

Operating Requirements

Manpower and catalyst/chemical make-up are not a factor in evaluating these two systems. There is no significant difference in manpower required, and catalyst and chemicals are not used.

Commercial Experience

For either the lock hopper system, or the Fuller-Kinyon pumps, the entire system is of concern, including controls, recycle compressors, pneumatic transport of coal, feeders, dust collection, and general handling of crushed coal.

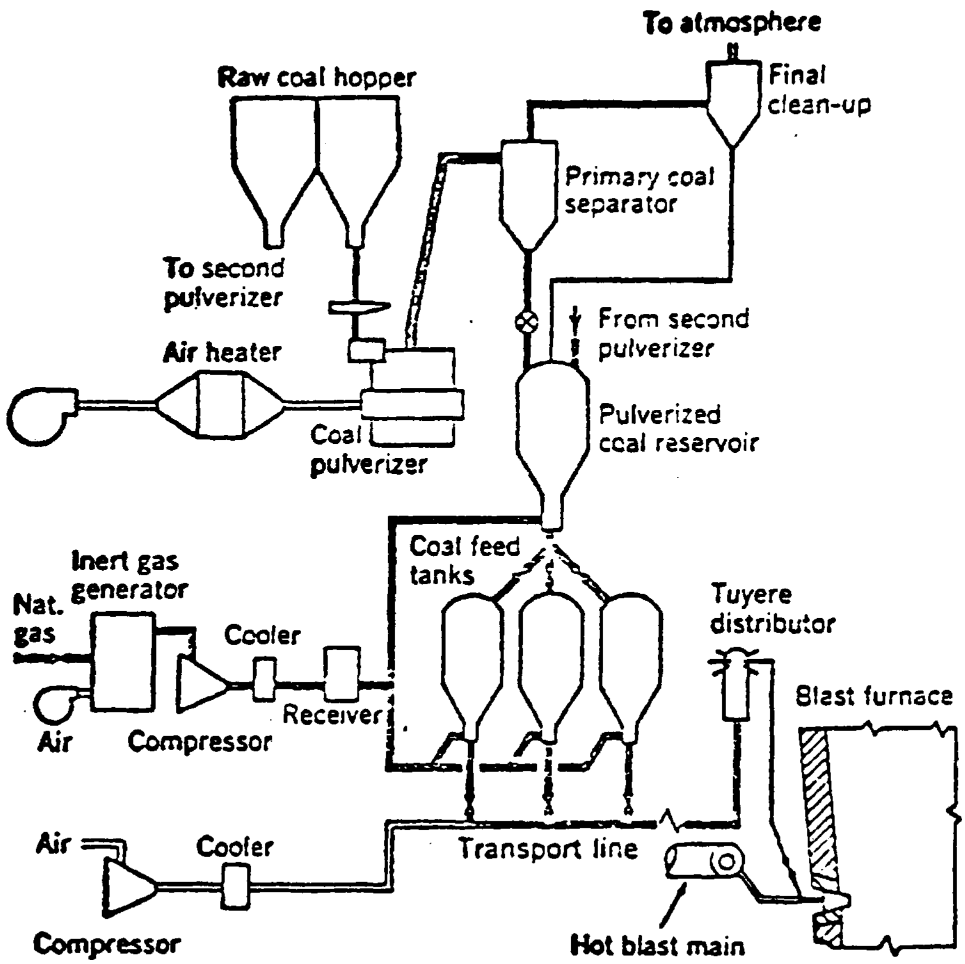
In general, lock hopper systems have been effectively proven. The Synthane Pilot Plant uses a lock hopper system designed to feed coal at 1000 psi. This system was supplied by Petrocarb. DUCON has also supplied lock hoppers for conveying solids into reactors under pressure. In one case, at the Ethyl Corporation in Baton Rouge, La., the reactor back pressure is 125 psig. Babcock and Wilcox have supplied the lock hopper system at Armco Steel's, Ashland, Ky., Works, Fig. 3.1.3-3. Here coal is injected into blast furnaces which operate at 45 psig. The system is designed to operate at 90 psig to provide pressurization and pressure for conveying the coal a distance of 340 ft.

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Figure 3.1.3-3

Babcock and Wilcox Armco Pulverized Coal Injection System



3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Commercial Experience - continued

Babcock and Wilcox's experience in conveying crushed coal on a commercial basis includes conveying 1/4" x 0 coal to their cyclone furnace installations, as well as the pneumatic conveying of pulverized coal to blast furnaces.

Fuller-Kinyon pumps, originally developed to pressurize pulverized coal, have been in use for many years to pressurize pulverized solids in many applications.

Early applications of Fuller-Kinyon pumps were to feed pulverized fuels for the firing of metallurgical furnaces, kilns and boilers. One of the first coal installations was in the mid-twenty's at Cahokia Station, the principal utility servicing St. Louis. The Fuller-Kinyon pump not only solved the explosion hazard previously inherent in handling coal, but also provided a practical means for controlling the spontaneous combustion of pulverized bituminous coal. A number of plants, including utilities, were designed to permit pump withdrawal of burning coal to empty storage bins, where the fire was extinguished.

Generally, the pumps operate at moderate pressure (about 25-75 psig) and handle large volumes (500 tons/hour of cement, which is one of the major applications). These pumps have been used on a large variety of materials including cement, cement raw materials, ores, catalysts, clays, coal, coke, gypsum, limestone, phosphate rock, silica, talc. One of the most significant applications of the Fuller-Kinyon pump is as part of a pneumatic conveying system to load and unload bulk cement barges or ships. There are approximately 150 of these installations in existence, which include fluidizing conveyors, pneumatic transport, air compressors, feeders and collecting screws, solids pumps, dust collectors, two-way and three-way valves and controls. Unloading rates are as high as 700 tons/hour.

There is only one Fuller-Kinyon commercial installation currently operating on coal, and this one is a historical curiosity. Most of the early coal pulverizing plants that used Fuller-Kinyon pumps have gone out of service. However, it was discovered that a 1922 installation in a Pennsylvania foundry was still in service on pulverized West Virginia coal. Pulverized coal is delivered by a Fuller-Kinyon pump from the pulverizing mill to two weigh bins. The discharge from either weigh bin is fed to a second Fuller-Kinyon pump which delivers the coal to the foundry via a 4-inch pipe over a distance of 700 ft. The pneumatic conveying is done at 55 psig.

3.0 PROCESS DESIGN STUDIES

FE-2012-096
MAY 1980

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Commercial Experience - continued

Although the pumps are generally used to handle a minus 50 mesh material, they can be used to pressurize larger size material. The importance of size consist is that the material forms a pressure seal, and finer material seals better. Coarser material require a longer screw. In pilot test on 1/4" x 0 coal, pressures as high as 90 psig have been achieved.

The coal pressurization tests are described in "Continuous High Pressure Lump Coal Feeder Design Study," EPRI Report AF410, March, 1977.[8] The test work was performed at the Fuller Company research department on a 4-inch Fuller-Kinyon pump. The results reported in the EPRI report indicated maximum pressurization of about 60 psig, which was the limit attainable with the existing motor. Continuation of the test program by Fuller-Kinyon resulted in pressures as high as 90 psig being attained, as reported by Jack Short of the Fuller Company.[9]

Technical and Economic Evaluation

Selection of a system depends on capital and operating costs, impact on the rest of the plant, equipment reliability, flexibility and capability to do the job.

The differenatial total installed cost for the lock hopper and Fuller-Kinyon pump systems are:

	Lock Hopper System	Fuller-Kinyon Pumps
Equipment Cost, $M	2,399	578
Total Fixed Capital, $M	5,398	1,300

Annual Operating Costs

The annual costs for a single train are summarized below:

	Lock Hopper System	Fuller-Kinyon Pumps
Power: KWH/yr	1.208×10^7	0.863×10^7
Cost, $M/yr	278	199
Capital Related Cost at 30% of Fixed Capital, $M/yr	1,620	390
Total Cost, $M/yr	1,898	589
Impact on Gas Price, ¢/MM Btu	7.25	2.25

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Technical Comparision

The significant technical comparisons are in the areas of particle size attrition, system complexity, valve wear in the lock hopper system, and screw wear in the Fuller-Kinyon pumps.

Western subbituminous coals, when dry, are very fragile and break up easily. This is anticipated to occur in the lock hopper system as well as in the Fuller-Kinyon pump system. However, because of the high shear forces generated in the pump barrel, this is potentially a more serious problem for the Fuller-Kinyon pumps. If the extent of particle size attrition is determined, then it can be compensated for by supplying a larger size consist to the pump. However, if an excessive amount of fines are produced in the pump that cannot be compensated for, the fines would have to be fed to pyrolysis.

In pumping tests on Illinois and Pittsburgh coal, the extent of particle size attrition was negligible.[8]

The Fuller-Kinyon system is much simpler than the lock hopper system in terms of equipment required and operation. In the Fuller-Kinyon system, coal is fed into the pump feed hopper, pressurized and picked up by a gas stream for pneumatic conveying. The pumps operate continuously and control is achieved by a variable speed drive.

The lock hopper system requires a whole array of vessels, cyclones, dust collectors and pneumatic lifts. The operation is intermittent and requires a fairly complex timing system for opening and closing valves for feeding coal, pressurization and venting.

Critical items in the lock hopper system are the ball valves, which are periodically opening and closing. These are usually designed for a two-year life span. However, the valves supplied to the Armco facility by Babcock and Wilcox for transporting coal to the blast furnace have been in continuous service since 1973.

The critical item in the Fuller-Kinyon pump is the feed screw. These are estimated to have a life span of nine months, but are easily replaceable in the field at a small cost.

The Technical Review Matrix contains a summary of these factors.

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

Technical Comparison - continued

Technical Review Matrix

	Lock Hopper	Fuller-Kinyon
Degree of Development	High	High
Technical Risk	Low	Moderate
Turndown Ratio	High	High
Operating Flexibility	High	High
Impact on Other Plant Areas	Low	Low
Reliability	High	High
Complexity	High	Low

Guarantees, Royalty and Fee

No royalties are involved in the use of either of the systems discussed. Fees, if any, would be related to the extent of engineering and/or vendor testing required.

At this stage, where budget estimates have been received for conceptual systems, specific guarantees have not been spelled out. The Fuller-Kinyon Company will guarantee their system if they supply the recycle compressors, as well as the pumps.

Recommendation

Based on the cost comparison of the two systems, which are significantly different, as well as the general operating characteristics of the two systems, which are basically similar in terms of development, reliability, flexibility, etc., the Fuller-Kinyon system is recommended for the Commercial Plant on Illinois coal. There is some question as to amount of size attrition of Western coal. It is possible that this uncertainty may be resolved as a result of a test program being proposed at Morgantown Energy Research Center. An alternate approach would be to undertake a test program, under Task IX of this contract, which would combine a crushing test, flash drying test, and Fuller-Kinyon pump test of Western coal.

3.1 AREA 102 - COAL PREPARATION

3.1.3 Coal Pressurization Design Study

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3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.1 Summary

Introduction

Ten process design Alternate studies were conducted in Area 103. Nine of the studies were done wholly by CDC in cooperation with equipment suppliers. The process control study was done by ICGG's consultant, Econergy Associates (EA). The studies were:

Pyrolyzer Heater

Solids Transport by Gasifier Gas

Mixing Section Vent

Slag Removal - Process Requirements

High Temperature Valves

Pittsburgh Coal Pyrolysis

Cyclone Requirements

Process Control

Mixing Section Design

Process Requirements - Char Cooling and Dump System

Alternates Considered and Recommendations

Pyrolyzer Heater

The pyrolyzer heater alternates were:

- Direct heating of the fluidizing gas
- Indirect heating of the fluidizing gas

Direct heating is the TBD case. Fuel oil is combusted with slightly less than the stoichiometric oxygen level to provide hot flue gas for fluidizing and heating coal in the first and second stage pyrolyzers. The flue gas from this system, after cooling and compression, is used to pressurize the coal feed lock hoppers. Gas, containing combustibles generated in the pyrolysis vessel, vented from the system is burned in the Power Generator Boiler.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

Pyrolyzer Heater - continued

The alternate scheme uses a fuel oil fired indirect heater to heat the second-stage pyrolyzer fluidizing gas with the flue gas venting to the stack. In this case CO_2 from Area 108, instead of Area 103 flue gas, is used for fluidization and solids transport gas. Only the vent gases from the lock-hopper system are incinerated to dispose of combustibles removed from the first and second pyrolysis vessels.

There were no apparent technical reasons why direct or indirect heating cannot be used. Because of higher operating and capital cost and reduced flexibility for the indirect heating system, it is recommended that the direct heating system, used in the TBD, should be retained.

Solids Transport

Two sources of gas were evaluated for transporting coal and char between pyrolysis stages. The gas sources were:

- Recycle Make-Gas
- Hot Gasifier Gas

It was concluded from the process design study that the recycle make-gas be used for the transport of the carbonaceous materials.

Mixing Section Vent*

Two routings of the vent gas from the mixing section were studied. As a result of the study it was decided to replace the TBD arrangement with the alternative because of the favorable reduction in steam usage and total gas volume.

Slag Removal - Process Requirements*

This study was undertaken to improve upon the slag removal system presented in the TBD. The improvements selected are based on the United Conveyor Corporation technology which has been successfully demonstrated for wet bottom utility boilers.

* Details of these studies are not included because of proprietary data considerations.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

High Temperature Valves

The alternates for high temperature valves for the gasifier were:

- To use existing valve types developed for fluid cracking units.
- To modify the existing valves to meet the higher operating temperatures by altering the materials of construction.
- To develop a new valve design.

The Crane Company stated that their existing hot valve design is suitable and only modifications to materials of construction are required to upgrade it to operating conditions found in the gasifier.

Pyrolysis and Gasification

The options for the Pittsburgh Coal Pyrolysis Study were:

- Pre-Oxidation of the coal
- Non-Oxidation

In the pre-oxidation case, air is injected into the first stage pyrolysis reactor to oxidize a portion of the coal and to render it non-agglomerating. Pre-oxidation has a detrimental effect on oil yield. For the non-oxidation case, an additional stage of pyrolysis is used and agglomeration is prevented by recycling an adequate quantity of inert char. Additional heaters and increased quench and compressor capacity are required for this case.

As a result of this study, the pre-oxidation system is recommended because:

- There is a better design basis for pre-oxidation in the successful COED pilot-plant operation, and thus less risk. However, the chances of success with non-oxidation are good.
- Economics slightly favor the pre-oxidation mode of operation with possible a break-even situation occuring in the future.
- Design and operation of a Commercial Plant is less complex utilizing the pre-oxidation mode of operation.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

Cyclone Requirements

For the cyclone study, alternate process conditions were not evaluated. Buell studied alternate configurations and pieces of equipment to meet the process conditions specified.

Buell determined the cyclone requirements for the pyrolysis, gasifier, combustor, and for heat and power recovery. Their report was a compilation of summary sheets, efficiency curves and arrangement drawings. It is recommended that the Buell design be adopted.

Process Control

The process design study "Analysis of the Integrated Operation of the Pyrolysis - Gasification Sections in the Gas Process for Illinois Coal Gasification Group" evaluated upsets resulting from:

- Coal feed rate
- Solids flow from one reactor to another

Equations were derived for calculating the effect of process upsets in the pyrolysis section on temperature transients which would occur if no corrective control actions were made. A computer program was developed for using these equations. This program was applied to possible upsets in the system in order to calculate temperature transients for these upsets and to estimate how quickly remedial action must be taken to avoid inoperable conditions in the pyrolysis reactors. Alternate methods of off-setting these upsets and maintaining controlled temperatures were evaluated and the preferred method was identified. A control system that could handle normal and upset conditions for Pyrolysis and Gasification was presented.

Mixing Section Design*

Alternate refractory shapes and designs were evaluated to make the mixing section more structurally stable and to compensate for thermal expansion. The process requirements and the operating conditions for the mixing section were not affected by this study.

A preliminary design developed by Harbison Walker satisfied the criteria established for the mixing section. However, modifications may be necessary when details of the support structure and the integration of the mixing section and the combustor are developed.

*Details of this study are not included because of proprietary data considerations.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

Char Cooling/Dump System

In the TBD case, an external cooling system was used to cool the char from the pyrolysis vessels and gasifier to a safe temperature when it is necessary to dump these vessels. An alternate scheme was developed where the char can be cooled in place by recycling inert gas with existing scrubbers, coolers and compressors in Areas 103 and 104.

The process specifications were developed and the feasibility of the alternate char cooling and dump system was determined. Although detailed economics were not developed, it appears that capital requirements will be reduced.

It is recommended that the alternate char cooling and dump system be incorporated into the design to replace the system shown in the TBD.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.2 Direct VS Indirect Gas Heater Design Study

Summary

A direct-fired auxiliary furnace is used in the Commercial Plant Tentative Baseline Design (TBD) to prepare an oxygen-free, inert gas to fluidize and heat the 1st and 2nd stage pyrolyzers. The gas generated from fuel oil combustion in the auxiliary furnace, along with any make gas from the first and second stages of pyrolysis, must be purged from the inert gas system to maintain a constant fluidizing gas rate. A portion of this gas is used as transport gas to the coal feed lock hopper system. The remaining gas must be vented to the utility boiler, requiring additional coal fuel to the boiler. An alternate flow scheme, which uses an indirect-fired auxiliary furnace to heat the second stage fluidizing gas, was considered. The fuel oil combustion products leave the furnace through a stack rather than being mixed with the recycle fluidizing gas. The quantity of vent gas that must be incinerated and the amount of coal burned in the utility boiler are reduced; however, the indirect-fired heater is a less efficient heat transfer source of transport gas for the coal feed hopper, namely CO_2 from Area 108, would be used.

The indirect-fired auxiliary furnace requires an additional 1182 lb/h of fuel oil and saves 3100 lb/hr. of coal to the utility boiler. This results in an increase in operating costs of $329,000 per year because of the higher cost of fuel oil. Capital equipment costs will also be greater for the alternate design due to the heat exchange surface required in the indirect-fired heater.

It is recommended that the direct-fired auxiliary furnace design should continue to be used. The proposed alternate would result in greater capital and operating costs. The direct-fired furnace provides a more flexible system in that it gives an inert gas for start-up. The alternate scheme would require an outside source of inert gas during start-up until CO_2 from Area 108 became available. The CO_2 would have to be steam stripped instead of air stripped to eliminate oxygen from this stream, which would result in higher costs for Area 108.

Performance Specification

The auxiliary furnace in each train of a commercial plant must produce an oxygen-free gas stream in sufficient quantity and at a

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.2 Direct vs Indirect Gas Heater Design Study

Performance Specification - continued

high enough temperature to fluidize the first two stages of pyrolysis in series and provide heat to the first two stages of pyrolysis. Gas leaving Pyrolyzer 1 is scrubbed and compressed and is reheated in the auxiliary furnace. The furnace is fired on product fuel oil at a rate to be controlled to maintain the desired 2nd stage pyrolyzer temperature.

Process Description

Process flow diagrams around the 1st and 2nd pyrolyzers and inert gas system for the TBD and alternate design are shown in Figures 3.2.2-1,2. In the TBD, scrubbed gas is compressed in compressor (103-42001) and fed to Auxiliary Furnace (103-44001), Fuel oil is burned sub-stoichiometricly with air in the direct-fired furnace to heat the recycle gas from the compressor discharge. The oxygen-free mixture of hot recycle gas and combustion products fluidizes and provides heat to the 2nd and 1st stage Pyrolyzers (103-33002 and 102-33001) in series. The first stage off-gases are scrubbed in Venturi Scrubbers (103-45011 and 103-45019) to remove dust and oil and to condense water. A portion of the gas is used to transport feed coal from the Lock Hopper System (102-43017) to the 1st stage pyrolyzer. The remaining gas, along with gas from the coal lock hopper system, is incinerated in the coal-fired Utility Boiler (115-44007). The oil firing rate to the auxiliary furnace is controlled to maintain the 2nd stage pyrolyzer temperature of the desired set point. The air rate to the furnace is set by an air to fuel oil ratio controller to maintain the desired degree of sub-stoichiometric burning.

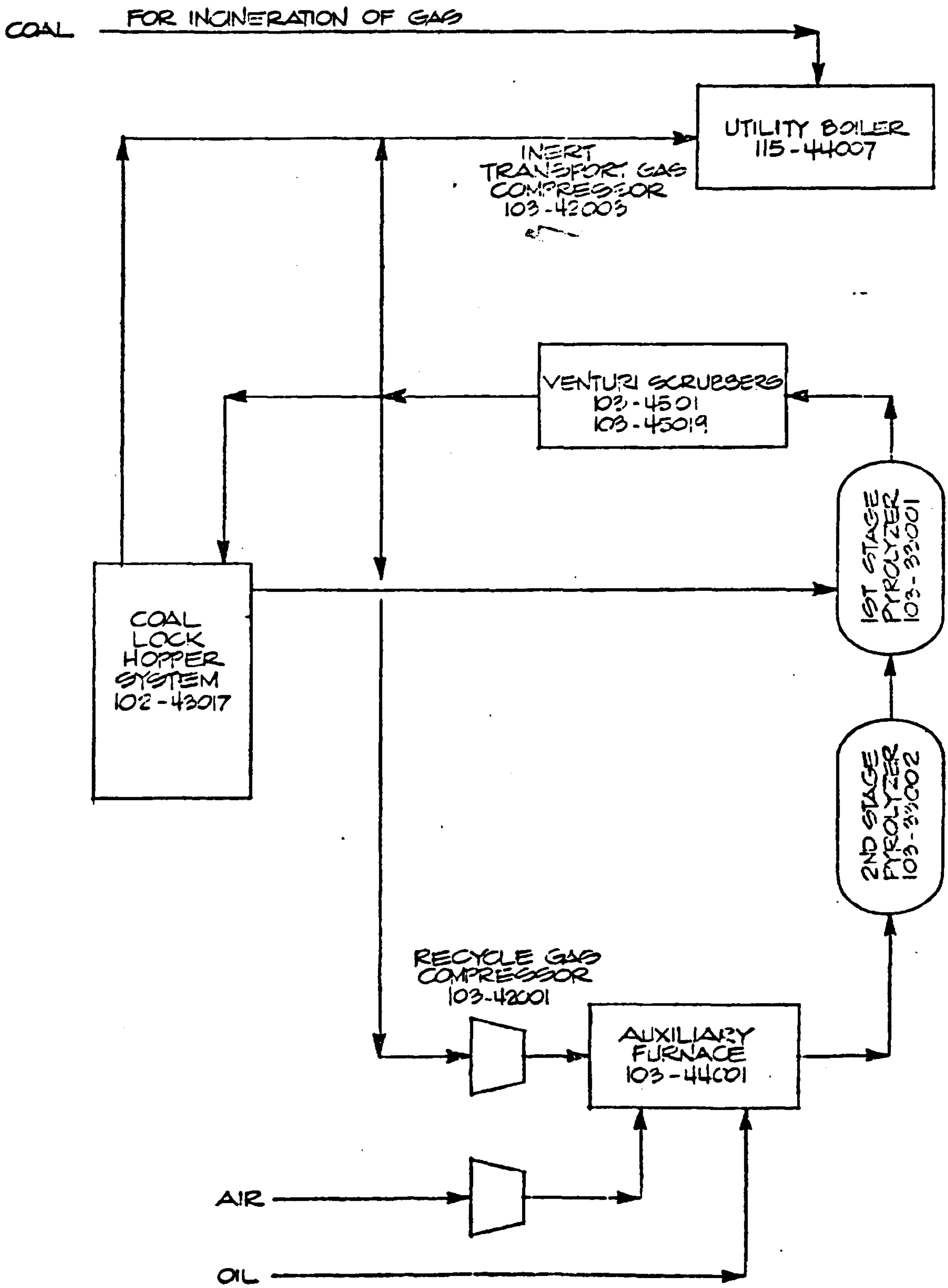
The alternate flow scheme differs from the TBD in that it uses an indirect-fired auxiliary furnace. The fuel oil combustion products leave the furnace via a stack to atmosphere rather than in a mixture with the heated recycle gas. Because of the heat loss in

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.2 Direct VS Indirect Gas Heater Design Study

Figure 3.2.2-1

Stage 2 Fluidization Gas Generation
Tentative Baseline Design

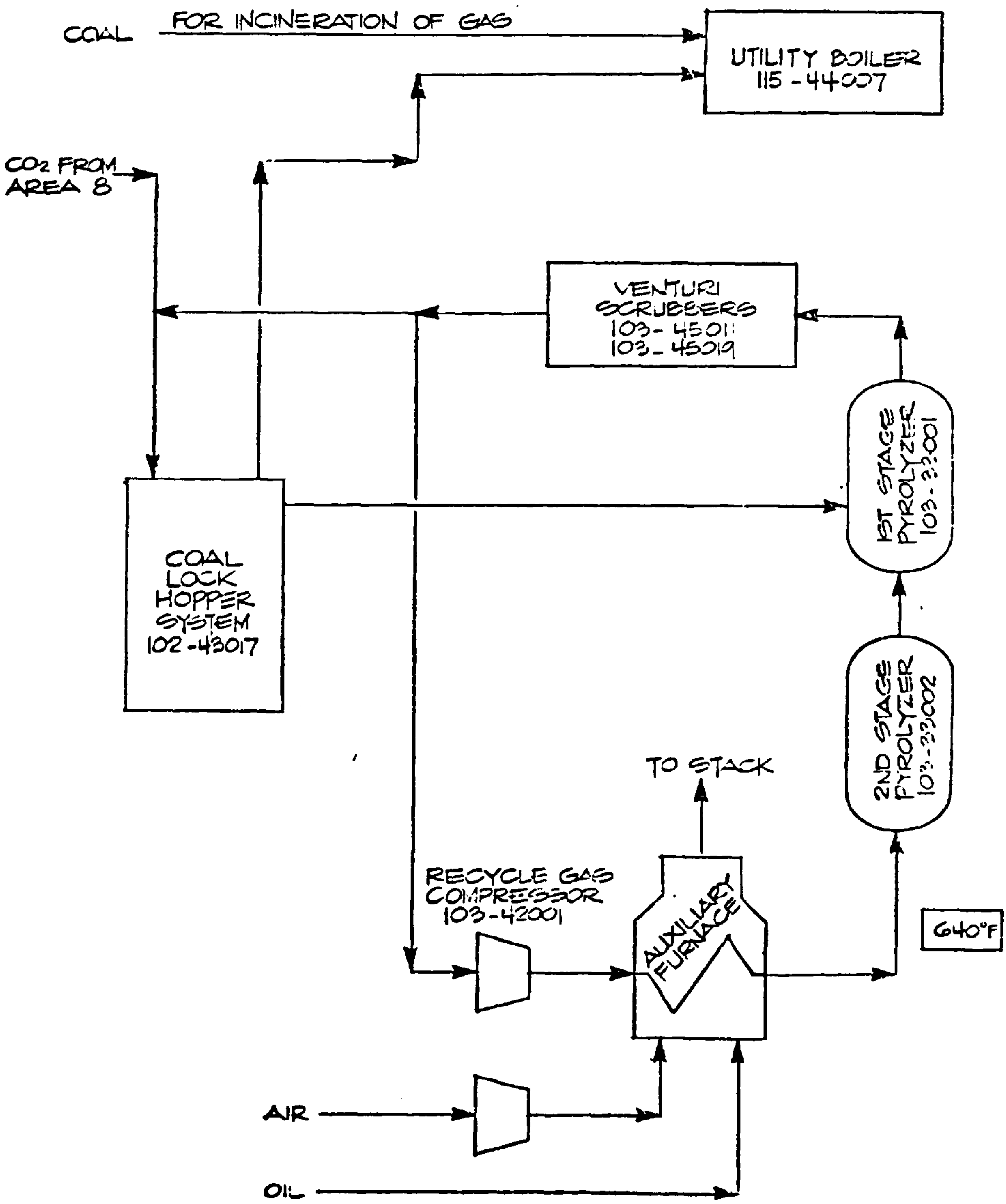


3.2.2 Direct VS Indirect Gas Heater Design Study

Figure 3.2.2-2

Stage 2 Fluidization Gas Generation

Alternate Design



3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.2 Direct VS Indirect Gas Heater Design Study

Process Description - continued

the stack gas, a higher fuel oil rate is required. Sub-stoichiometric burning is not necessary. As the increase in recycle gas volume is due to gas make in the first two pyrolysis stages alone, there is little inert gas to be vented. CO_2 from Area 108 must be used for transport gas. The vent from the coal lock hopper system is the only gas stream to be incinerated in the utility boiler. This alternate scheme uses an additional 1182 lb/h of fuel oil because the indirect-fired auxiliary heater is less efficient than the direct-fired heater, but saves 3100 lb/h of coal because less vent gas must be incinerated in the utility boiler.

Technical and Economic Evaluation

The TBD flow scheme provides a greater degree of process flexibility than the alternate case. A source of inert gas is available when needed during plant start-up. The alternate case would require a start-up inert gas system, as CO_2 from Area 108 would not be immediately available. The direct-fired furnace is a more flexible heat source than the indirect-fired furnace as the heat duty is only a function of the oil firing rate and not dependent on any heat transfer area constraint.

Both direct and indirect-fired process gas heaters are common in the chemical and petroleum industries. John Zink Company reports that it has several direct-fired gas heaters in commercial plants burning heavy No. 6 fuel oils sub-stoichiometricly without operating problems. Firing a direct-fired furnace with No. 4 fuel oil product is well within common commercial practice.

As there are no apparent technical reasons precluding the use of either the TBD or alternate scheme, the choice becomes one of economics and operating flexibility. The major concern is the higher fuel oil usage for the alternate case versus the higher coal usage in the utility boiler for the TBD. As calculated in Table 3.2.2-1, the alternate case would result in a $329,000 per year higher operating cost due to the increase in fuel oil required. The capital cost for the indirect-fired furnace would be higher than the direct-fired furnace because of the needed heat exchange surface. The alternate system would necessitate steam stripping rather than air stripping of the CO_2 in Area 108 to provide an oxygen-free CO_2 stream. Thus, the Area 108 costs would also be higher.

3.2 AREA 108 - PYROLYSIS AND GASIFICATION

3.2.2 Direct vs Indirect Gas Heater Design Study

Table 3.2.2-1

Operating Economics of Stage 2 Fluidizing Gas Generation

	TBD	Alternate
Fuel Oil Rate,[1] lb/h	0	+ 1182
Coal Rate, lb/h	0	(- 3100)
Fuel Oil Cost,[2] $/yr	0	+605,000
Coal Cost,[3] $/yr	0	(-276,000)
Total Annual Cost, $	0	+329,000

[1] TBD Case = 0
(-) = Savings

[2] At $15.40/bbl

[3] At $22.50/ton

Recommendations

Because of the more favorable economics and greater process flexibility of the TBD design over the alternate design, the TBD design should be retained.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.3 Use of Gasifier Gas for Solids Transport

The Commercial Plant Tentative Baseline Design (TBD) calls for the use of recycle make-gas from Area 104 for char transport in Area 103. After scrubbing and recompression in Area 104, a portion of the make-gas enters Area 103 for char transport. This gas leaves with the Area 103 product gas at a higher temperature, thus imposing an additional heat load on pyrolysis. It was suggested by Econergy Assoc. that the hot gasifier gas could be used for char transport. This would possibly eliminate or al least reduce the size of the recycle gas compressors and the added heat load on pyrolysis with a resulating increase in thermal efficiency. This concept has been investigated and the results are discussed below, and summarized in Table 3.2.2-1.

In the TBD the pressure of the gasifier off gas is lower than that required by some transport gas users. In some cases this can be solved by redesign of the transport gas users.

Other potential problems include controlling the flow of the hot gasifer gas to the various transport gas injection points. Many high temperature control valves would be required for each commercial plant train. A flow measurement technique would also be required. One proposal would be to inject a constant flow of tracer gas and determine the gasifier gas flow by measuring a temperature or concentration difference. Replacement of all the low pressure transport gas would also result in a significant increase in stage 3 oil concentration. Oil condensation may be a problem. It is not known how critical this may be to the pyrolysis operation without actually trying it.

On the positive side, replacement of all the low pressure transport gas with gasifier gas would result in a pyrolysis heat savings, mainly as fuel oil to the auxiliary furnace and drive steam to the recycle transport gas compressor. This translates into a thermal efficiency increase of about 1 percent and an annual operating cost savings of roughly 7¢/MM Btu.

It was concluded that because of the added technical problems associated with the use of gasifier gas for char transport, a change in the TBD is not justified.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.3 Use of Gasifier Gas for Solids Transport

Table 3.2.3-1

Use of Gasifier Gas for Solids Transport

Possible Problem Areas	Possible Solutions
Gasifier gas supply pressure too low	Redesign char transport lines
Flow Control	Use hot control valves and tracer gas for flow measurement.
Oil Condensation	May not be critical to pyrolysis operation

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Purpose

In this study the oxidation and non-oxidation cases for the Pittsburgh coal are evaluated economically against each other.

Introduction

An economic comparison was made for all plant areas affected by the two cases under study (see table 3.2.4-1). All the equipment in Areas 102, 103, 104 and 106 were sized depending on the design rate, in order to scale them up or down against the Illinois case in the ICGG proposal.[10] As an example, compressors were sized according to the volumetric flow rates and the pressure ratio.

Area 110 received a greater flue-gas rate for the non-oxidation case than the pre-oxidation case. The Area 110 total equipment cost for each case was determined by ratioing the flue gas flow to the 0.7 power times the total equipment cost as presented in the ICGG proposal. [6]

High cost vessels were sized according to diameter and height. Less costly vessels were sized only by rationing the flow rates. This can be done because, as given below, the vessel metal volume is directly proportional to the flow rate, assuming that the velocities are equal in each vessel.

$$\left(\frac{S_2}{S_1}\right)^a \times (\$\,S_1) = (\$\,S_2)$$

S_2 = New Sizing Parameter Flow Rate

S_1 = Old Sizing Parameter Flow Rate

$\$S_1$ = Old Equipment Cost Estimate from ICGG Proposal

$\$S_2$ = New Equipment Cost Estimate

a = Exponent Dependent on Type of Equipment

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Economic Comparison - continued

The pyrolysis and gasifier vessels were sized in accordance with the procedures presented above.

Table 3.2.4-1

Capital Equipment Cost

Basis: 256 MM SCFD of 950 Btu/SCF Pipeline Gas

Capital Equipment Cost: $ MM

Area	Non-oxidation	Pre-oxidation	Δ Cost	% of Total Δ Cost
102	5.7	5.4	0.3	1.3
103	79.0	71.9	7.1	30.7
104	75.5	63.9	11.6	50.2
106	6.5	5.1	1.4	6.1
110	51.7	49.1	2.6	11.7
Total	218.4	195.4	23.0	100.0

Eighty-one percent of the increase in equipment cost for the non-oxidation case is attributed to Areas 103 and 104.

In Area 103, the major increase in equipment cost for the non-oxidation case can be attributed to the fifth pyrolysis reactor, recycle make gas heater, and a new compressor for pumping inert gas to transport char from the third to second-stage pyrolysis reactor. For Area 104, increased equipment cost for the non-oxidation case is attributed to the larger recycle make gas flow to Area 103, associated equipment, and a larger hydrotreating system.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Recommendations

An economic trade-off analysis was made for the Commercial Plant to determine whether pre-oxidation or non-oxidation is the better mode for operating the pyrolysis reactors with Pittsburgh-seam coal.

Pre-oxidation of the coal with air in the first-stage pyrolysis reactor substantially reduces the agglomerating tendency of the coal. This method permitted successful COED pilot-plant operation[1] with Pittsburgh-seam coal, but reduced the oil yeild by about 20 percent.

If pre-oxidation is not used, the coal must be diluted with sufficient inert recycle char to prevent agglomeration. The non-oxidation case, but also increased equipment cost in make gas quenching and oil hydrotreating.

On this basis, the Pittsburgh-seam pre-oxidation case resulted in a net saving of $4.7 MM/yr. This reduces the pipeline gas cost by 5.3¢/MM Btu.

The COED pilot-plant experience indicates difficulty in operating with Pittsburgh-seam coal in the non-oxidation mode. It is recommended that the pre-oxidation mode of operation be used for the Demonstration Plant design because of the above economic advantage and the lower risk involved.

Table 3.2.4-2 summarizes the major items in the Pittsburgh-seam coal study:

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Table 3.2.4-2

Major Pittsburgh Coal Pyrolysis Factors

Basis: 256 MM SCFD pipeline gas at 950 Btu/SCF.

Variables	Non-Oxidation	Pre-Oxidation
Plant Areas Analyzed for Equipment Costs	102, 103, 104, 106, 110	102, 103, 104, 106, 110
Number of Pyrolysis Reactors	5	4 (air to 1st stage)
Recycle Gas Requirement	171,884 lb-moles/hr	111,571 lb-moles/hr
Coal Feed, TPD	21,330	19,530
Net Syncrude Product, lb/hr	248,553	173,468
bbl/d	15,698	10,956
Equipment Cost, $ MM		
Area: 102	5.7	5.4
103	79.0	71.9
104	75.5	63.9
106	6.5	5.1
110	51.7	49.1
Annual Cost of Equipment, $ MM	165.1	147.7
Annual Operating Cost, $ MM	77.0	89.7
Total Annual Cost, $ MM	242.2	237.5
Flexibility of Area 103	Capable of operating in pre-oxidation mode.	Remote possibility of operating in non-oxidation mode.
Technical Risk	Moderate risk because low risk operation is not proven.	Low risk based on COED pilot plant experience.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Process Description

Process Area 102 receives undried coal, grinds it to the desired size and feeds it into the first stage of pyrolysis in Area 103, which accomplishes drying and preheating of the coal.

The pyrolysis reactors in Area 103 (four for the pre-oxidation case, five for the non-oxidation) fluidize the coal at increasingly higher temperatures to remove the volatile organics which cause agglomeration of coal and to evolve oil and gas by thermal decomposition of the coal.

Char from pyrolysis is gasified in the gasifier. Make gas from the gasifier provides the main source of heat and fluidizing gas for the pyrolysis reactors.

In Area 104, the make gas exiting the pyrolysis vessels is cooled. Oil is separated from the gas and is hydrotreated to produce a saleable product. Recycle make gas is compressed for use as transport gas in Area 103.

In Area 106, a portion of the make gas from Area 104, which has been compressed and stripped of acid gases and light hydrocarbons, is reacted via the shift reaction with steam to produce hydrogen for the oil hydrotreating reactors in Area 104.

Operating Requirements

Manpower requirements, both operating and maintenance, should be about the same for both cases. Area 104 catalyst and chemical requirements should be slightly higher for the non-oxidation case because of the greater oil production.

Pilot-Plant Experience

Pittsburgh-seam coal was pyrolyzed successfully in the COED pilot plant by pre-oxidizing the coal with 0.4 to 0.5 SCF O_2 per pound of dry coal fed into the first stage of pyrolysis.[1]

The only successful COED operation using Pittsburgh-seam coal without oxidation was achieved in the 3-inch bench-scale unit. It was found that four pyrolysis stages (at 640°F, 740°F, 810°F and 950°F) and one heat-producing stage (at 1600°F) were required to pyrolyze the coal. In addition, dilution of the coal with inert char was also required to prevent agglomeration.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Annual Capital Costs

The total equipment costs for Areas 102, 103, 104, 106 and 110 are \$218.4 and \$195.4 million for the non-oxidation and pre-oxidation cases, respectively. See Table 3.2.4-3.

A factor of 2.25 was used to convert equipment cost to installed equipment cost in the ICGG Proposal.[6] Escalation was taken at 12 percent between mid-1975 and mid-1977.

Annual capital-equipment related charges have been taken as 30 percent of the installed capital cost.

Table 3.2.4-3

Annual Costs, MM \$

(Mid-1977 Dollars)

Basis: 256 MM SCFD @ 950 Btu/SCF Pipeline Gas

	Non-Oxid.	Pre-Oxid.	Non-Oxid.-Pre-Oxid.
Installed Equipment Areas 102, 103, 104, 106 and 110	165.1	147.7	17.4
Coal	183.0	167.6	15.4
Annual Revenues from By-Product Sales MM \$			
Fuel Oil	(79.8)	(55.7)	(24.1)
Lt. Hydrocarbons	(15.3)	(12.6)	(2.7)
Sulfur	(8.3)	(7.6)	(0.7)
Ammonia	(2.5)	(1.9)	(0.6)
Net Annual Cost MM \$)	242.2	237.5	4.7

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Annual Capital Costs - continued

The pre-oxidation case is economically favored by $4.7 MM per year. This results in the net difference of 5.3¢/MM Btu in pipeline gas cost for the two cases.

If it is assumed that the accuracy of the installed equipment cost is ± 25 percent, the total annual differential cost varies from $0.4 MM to $9.1 MM. The pre-oxidation case is still slightly favored if the installed equipment cost is lowered by 25 percent.

The installed equipment, coal, and fuel oil annual costs and credits are the major items in the annual cost of the two cases. If the coal and oil costs are varied separately, ± 25 percent, the following is obtained:

1. Differential annual cost, when coal cost varies ± 25 percent, is from $0.9 MM to $8.6 MM. The pre-oxidation case is still slightly favored if the coal cost is lowered by 25 percent.

2. Differential annual cost, when oil selling price varies ± 25 percent, is - $1.3 MM to $0.7 MM. The non-oxidation is favored at high oil prices and pre-oxidation favored at lower oil prices.

If installed equipment and coal costs vary favorably for the non-oxidation case, each by -25 percent, then the differential annual cost will favor the non-oxidation case by $3.5 MM/yr.

A more realistic situation would be the equipment cost down by -20 percent and coal and fuel oil costs up by +15 percent. This would produce a result slightly in favor of the non-oxidation case by $0.1 MM/yr.

Economically, the above results indicate the pre-oxidation case is marginally favored. In the near future, the balance may swing to a break-even situation or slightly favor the non-oxidation case.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Annual Product Credits

The non-oxidized case produces 4,742 bbl/d more fuel oil for sale than the pre-oxidation case. The fuel oil was credited at $15.40/bbl. A portion of the fuel oil produced is used internally in the process for heating.

Sulfur production is proportional to the total quantity of coal fed to the process and was credited at $43/long ton. The difference between the non-oxidation and the pre-oxidation cases is 55 TPD.

Light Hydrocarbons (C_2-C_4) are produced in the pyrolysis and hydrotreating sections of the plant. Some of the light hydrocarbons produced are used to upgrade the product gas to 950 Btu/SCF. and the remainder are sold. The non-oxidation case, because it produces more light hydrocarbons, is given a credit of $8,329 per day.

Ammonia production results from nitrogen removed in hydrotreating and produced in pyrolysis sections of the plant. The non-oxidation case produces 13 TPD more ammonia than the pre-oxidation case.

Steam production is an estimate of the total utilities produced less the requirements of each process. Credit was not given to either case because the differences in total steam and other utility requirements appear small and were uncertain at the time of this study.

Technical and Economic Evaluation

Steam production for the two Pittsburgh-seam coal cases was determined by calculating equivalent steam production for all major heat sources less major users. The difference in the steam available from the Pittsburgh-seam coal cases, times a charge of $3.81/M lb for 1500 psig, 850°F steam, resulted in a small credit for the non-oxidation case. Whether an actual credit will be realized for the non-oxidation case is subject to some doubt, since total utility requirements are not known as yet. In the event that such a credit is justified, a slightly more favorable economic situation for the non-oxidation case results.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Technical and Economic Evaluation

Alternate Coal Material and Energy Balance Results

A material and energy balance for Pittsburgh-seam coals has been completed. The Pittsburgh case consists of two subcases, one in which the coal is pre-oxidized with air in the first stage pyrolyzer, resulting in a higher gas yield and lower oil yield; and one in which no pre-oxidation is used. The results are shown in Table 3.2.4-4. A pipeline gas make of 256 MM SCFD is the basis for Table 3.2.4-4.

A comprehensive material balance was performed for Areas 102, 103, 110; and a "battery limits" balance for Areas 101, 104-109, 110-113 for the two Pittsburgh seam coal cases. This "battery limits" balance consisted of computing all inputs and outputs and some major internal streams. Ratioing from the Illinois case was used when appropriate. The major objective was the calculation of all product streams and major utility consumers and producers.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Table 3.2.4-4

Summary of Significant Parameters For Illinois and Alternate Coals

Areas 101 and 102

	Item	Pittsburgh Non-Oxid.	Pittsburgh-Pre Oxid.
1.	Dry Coal Input (TPD)	21,330	19,530
2.	Entering Moisture (wt%)	4.00	4.00
3.	MAF Coal Burned in Dryer (TPD)	0	0
4.	Flue Gas from Area 110 (lb mole/hr)	0	0
5.	Dry Coal into Pyrolysis (TPD)	21,330	19,530
6.	Exiting Moisture (wt %)	4.00	4.00

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Table 3.2.4-4 - continued

	Item	Pittsburgh Non-Oxid.	Pittsburgh Oxid.
	Area 103		
1.	Number of Pyrolysis Stages	N+1	1 (pretreating)
2.	Slag to Landfill ;is (TPD)	1,409	1,300
3.	Total 1000°F Steam (lb/hr)	1,299,819	1,224,542
	Area 104		
1.	Phenol to Area 113 (lb/hr)	1,432	1,623
2.	Gas to Area 105 lb mole/hr	111,222	109,516
3.	Hydrogen from Area 106 (lb mole/hr)	9,199	7,098
4.	Fuel Oil Output (lb/hr)	66,784	65,054
5.	Syncrude Output (lb/hr)	248,553	173,468

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Table 3.2.4-4 - continued

		Areas 105-109	
1.	Total Gas In (lb mole/hr)	111,222	109,516
2.	Total Gas Out (MM SCFD) (Dry)	256.4	256.4
3.	LHC for Sale (lb mole/hr)	619.3	508.6
4.	Carbon Dioxide Removal (lb mole/hr)	11,629	N/A
5.	Acid Gas to Area 112 (lb mole/hr)	4,152	N/A
6.	Gas to H_2 Generation (lb mole/hr)	10,737	8,215.1
7.	Gas to Shift & Methanation (lb mole/hr)	92,880	94,129
		Areas 111-113	
1.	Total Dirty Gas into Area 111 (lb mole/lb)	245,963	N/A
2.	Elemental Sulfur Product (lb/hr)	54,918	50,283**
3.	Total Liquor into Area 113 (lb/hr)	538,753	N/A
4.	Ammonia Product (lb/hr)	4,859	3,763**

N/A Not Available

** Ratioed from Pittsburgh Non-Oxid.

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Table 3.2.4-4 - continued

Area 115

	Item	Pittsburgh Non-Oxid.	Pittsburgh Oxid.
1.	Additional Coal Into Boiler (Basis = Illinois Case) TPD = 0	1,377***	1,276***
2.	Coal (Dry) from Area 102	0	0

Area 110

	Item	Pittsburgh Non-Oxid.	Pittsburgh Oxid.
1.	Total Process Air (lb mole/hr)	213,427	197,951
2.	1500 psig, 850°F steam lb/hr	1,174,712	1,125,613**

** Ratioed from Pittsburgh Non-Oxid.

*** Relative Numbers - to be used for comparision between the two Pittsburgh cases only

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

Commercial Plant

Design and operation of the Commercial Plant utilizing Pittsburgh-seam coal will be more complex for the non-oxidation case because of the additional equipment in Areas 103 and 104. These include an additional pyrolysis reactor, a large steam heater for recycle gas; and in Area 104 additional recycle compressors (54 percent more recycle gas), a much larger quench and cooling system, plus a larger oil hydrotreating system (20 percent larger).

Recommendations

For Pittsburgh-seam coal, the pre-oxidation system is recommended because:

1. There is a better design basis for pre-oxidation in the successful COED pilot-plant operation, and thus less risk.

2. Economics slightly favor the pre-oxidation mode of operation with possibly a break-even situation occuring in the future.

3. Design and operation of a Commercial Plant is less complex utilizing the pre-oxidiation mode of operation.

3.0 PROCESS DESIGN STUDIES

FE-2012-096
May 1980

3.2 AREA 103 - PYROLYSIS AND GASIFICATION

3.2.4 Pittsburgh Coal Pyrolysis Design Study

References

1. J. F. Jones, et al, "Char Oil Energy Development, Vol. 1, Final Report for the Period Aug. 18, 1971 - June 30, 1975" ERDA Contract No. E (49-18) - 1212 (Sept. 1975).

2. R. T. Eddinger, et al, "Char Oil Energy Development, Period of Performance, June 1962 - Dec. 1965", OCR Contract No. 14-01-001-235.

3. "Char Oil Energy Development" - Project COED, Period Covered July 1972 - June 1973, R. & D Report No. 73 - Interim Report No. 2, OCR Contract No. 14-001-1212.

4. "Char Oil Energy Development" - Project COED, Period Covered July 1973 - June 1974, R & D Report No. 73 - Interim Report Nl.5, ERDA Contract No. E (49-18) - 1212.

5. M. R. Schmid, et al, "Concluding Report on the Operation of the Process Developmemt Unit - Pyrolysis of Coal in Five Stages" PRC-708 (Oct. 13, 1969).

6. ICGG Proposal for Pipeline Gas Demonstration Plant Program, Volume 5 (Economic Assessment), 4 February 1976.

7. CDC memo, L.H. Dierdorff to M. E. Sacks, "Cost Factors for Trade-Off Studies", October 20, 1976 (Tech File 1.6.1).

8. CPD-7051, "Unit Costs and Utility Flow Diagrams, "August 19, 1977; Transmittal Letter No. 40-C, N.A. McClintock (Dravo) to R. Bloom, Jr. (CDC), September 30, 1977.

3.0 PROCESS DESIGN STUDIES

FE-2012-096
May 1980

3.2 PYROLYSIS AND GASIFICATION

3.2.5 Process Requirements - Char Cooling and Dump System

Introduction

This report probides information developed on the process requirements and design of a proposed new approach for cooling char in the original vessels before dumping the char into the storage silo. This system will avoid the additional capital investment for separate heat-exchange vessels, whose only function is char cooling.

Process Requirements

The process requirements for a commercial plant char cooling and dump system are as outlined below:

a) When required for plant maintenance or shutdown, char cooling and dump systems for the pyrolysis and gasifier vessels are to cool and dump the char within 96 hours (maximum) after initiating shutdown.

b) Char in the 1st and 2nd stage pyrolyzer vessels is not to be dumped unless cooled to a temperature of 200°F or less.

c) Char in the 3rd and 4th stage pyrolyzer vessels and the gasifier is not to be dumped unless cooled to a temperature of 400°F or less.

d) The cooling rate of refractory lined vessels should not exceed 150°F/hr.

e) An estimate was made of the quantity of char to be dumped, the vessel weights that must be cooled per train; and the initial temperatures at the start of the cooldown cycle were identified.

f) Char to be dumped into closed char silos with sufficient capacity to hold all the dumped char.

g) The system should provide a means of feeding the char back to the first stage pyrolyzer during subsequent plant start-up.

3.2 PYROLYSIS AND GASIFICATION

3.2.5 Process Requirements - Char Cooling and Dump System

Proposed Char Cooling and Dump System

In the proposed design, after plant shutdown is initiated and before hot char is dumped from the pyrolysis vessels, the char is first cooled to 200°F-400°F by means of inert gas recirculation through the pyrolysis fluidized beds. Inert gas is supplied by an Inert Gas Generator and recycled by means of the Recycle Transport Gas Compressors in Area 204. The time required for cooling the char was calculated and appears to be practical.

Whenever char in any vessel has been cooled to 200°F or lower, the char is ready for transfer to the recovered Char Silo (103-35012) through the Char Dump System. This consists of char dump or drain lines and valves from the major vessels and from all low points in the process lines carrying char, including the pneumatic conveyor to the storage silo, the recovered char storage silo and auxiliaries, and the transfer line from the char storage silo of Area 102 for plant start-up.

When tranferrring char from a dump line to the char storage silo, only one char dump line at a time may be used to avoid the need for a pressure feeder at each dump location. Time requirements of sequential dumping have not yet been determined to see if they meet the specified time requirements for the cooling dumping cycle.

To permit char drainage from all four pyrolyzer vessels down to the grid level, special char dump standpipes must be installed. Plenum drains are also necessary on all pyrolyzer vessels to empty the vessels completely.

Before dumping char from any location, a pressure of several psig must be built up at the dump point. The magnitude of the pressure required must be determined by calculation from pneumatic conveying correlations.

Conclusion

The proposed design meets process requirements for the char handling systems. Although detailed economics were not developed, it appears that capital requirements will be reduced.

Recommendation

It is recommended that the alternate char cooling and dump system be incorporated into the design to replace the system shown in the TBD.

3.2 PYROLYSIS AND GASIFICATION

3.2.5 Process Requirements - Char Cooling and Dump Systems

Proposed Char Cooling and Dump System

The TBD utilized an external cooling system for each train. Hot char from pyrolysis vessels and the gasifier is passed through the cooling system and cooled to roughly 200°F, then transferred to a char storage silo. This system is used only during plant shutdowns when char cooling is required.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.1 Summary

Purpose

The process for Oil Recovery and Treatment of pyrolysis off-gas presented in the Tentative Baseline Design (TBD) has been reviewed by CDC with the purpose of determing whether more technically and/ or economically suitable systems might be used.

Processes Considered

Efforts focused on three alternate schemes in processing:

- A study comparing the proposed three-stage Venturi scrubbing system for oil recovery and gas cooling and cleaning, as shown in the TBD, with a system utilizing two-stage Venturi scrubbing and an electrostatic precipitator substituted for the third-stage scrubber.
- A study to determine whether a vacuum distillation system for concentration and removal of entrained solids from the naphtha-free pyrolysis oil would be more advantageous than the Rotary Drum Vacuum-Precoat Filter scheme outlined in the TBD.
- A study considering the technical and economic advantages of producing and marketing fuel oil-naphtha rather than the syncrude oil produced in the TBD case.

The design studies, evaluations, and recommendations derived for each of the alternate schemes considered are presented in Subsections 3.3.2, 3.3.3, and 3.3.4.

Results

The results of the studies indicate that for the Commercial Plant Design the three-stage Venturi scrubber system should be retained for oil separation and the cooling and cleaning of pyrolysis gas; the processing of the naphtha-free pyrolysis oil using vacuum distillation will be much more advantageous, technically and economically, than utilization of the originally proposed filtration scheme; and in the consideration of marketing fuel oil-naphtha versus marketing syncrude

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.1 Summary

oil, the decided advantages of reduced fixed capital requirements, lower annual costs, and increased annual sales revenues resulting from fuel oil-naphtha production led to the recommendation for elimination of syncrude production.

3.3.2 Venturi Scrubber vs. Electrostatic Precipitator Design Study

Summary

The method proposed in the TBD used three Venturi scrubbers in series. This cleaned and cooled the pyrolysis gas, condensed the oil, and removed the fines and the oil-water-fog from the pyrolysis gas. In this trade-off study, the process schemes evaluated were:

A. A third stage Venturi scrubber as utilized in the TBD.

B. Substitution of an electrostatic precipitator for the third stage Venturi scrubber, equivalent to the system used in the COED Pilot Plant.

Cottrell Environmetal Sciences of the Research-Cottrell Corporation was contacted to provide information on performance, operating and utility requirements and equipment cost for the two process schemes being considered. An outlet particulate level of 0.02 to 0.002 grains per ACF, which Cottrell stated could be accomplished by either process scheme, was used as design guideline. Ingersoll-Rand Corporation was also contacted, and verified that good performance with centrifugal compressors could be obtained when compressing gases with a particulate level of 0.02 grains per ACF.

From the information obtained, it was concluded that the third stage Venturi scrubber scheme should be used to clean and cool the pyrolysis gas. The primary reason for this conclusion is the lower incremental plant investment and lower incremental annual cost for the third stage Venturi scrubber system. These factors are compared in Table 3.3.2-1 on a total commercial plant basis.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.2 Venturi Scrubber vs. Electrostatic Precipitator Design Study

Table 3.3.2-1

Summary Comparison -
Third Stage Venturi Scrubber vs. Electrostatic Precipitator

Process Scheme	Total Incremental Plant Investment \$MM (6/1977)	Total Incremental Annual Cost \$MM (6/1977)
A (Venturi)	3.96	1.946
B (Precipitator)	7.02	2.264
Savings	3.06	0.318

Although Cottrell indicated that both systems have about the same operating reliability, it is concluded that a scrubber system is simpler and should offer more operating dependability.

Based on the results of this study, it is recommended that the third stage Venturi scrubber, as proposed in the TBD, be used to clean and cool the pyrolysis gas.

Performance Specification

The only significant performance specification is the degree of particulate cleanup needed to insure adequate life of the downstream compressors. Ingersoll-Rand Corporation indicated that centrifugal compressors could tolerate a particulate-oil-water mist level of 0.02 gr/ACF. The impellers in these centrifugal compressors could be hard-coated to give increased erosion life. Research-Cottrell estimated that this level was achieved in the two-stage venturi, electrostatic precipitator system in the COED Pilot Plant.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.2 Venturi Scrubber vs. Electrostatic Precipitator Design Study

Process Description

Two different process schemes have been considered in this trade-off study to determine the preferred method of cleaning and cooling pyrolysis gas. The methods chosen were:

A. A third stage Venturi scrubber as utilized in the TBD.

B. Substitution of an electrostatic precipitator for the third stage Venturi scrubber.

In both process schemes, two stages of Venturi scrubbers in series are used to cool the pyrolysis gas, condense the oil, and remove the bulk of the particulates and the oil-water fog. This is done in the following manner. Pyrolysis gas from the third stage pyrolyzer at 823°F is contacted with aqueous scrub liquor in a Venturi scrubber. The water, oil and gas mixture, having been cooled to 225°F, is separated in a quench separator. The liquids are sent to an oil decanter and the gas stream is sent to a second Ventrui scrubber for additional cleaning of the gases. The water, oil and gas mixture is again separated in a quench separator; and then gas is further cooled to 160°F in heat exchangers, and water and oil are separated out. This cooled gas is sent to the third stage of cleaning.

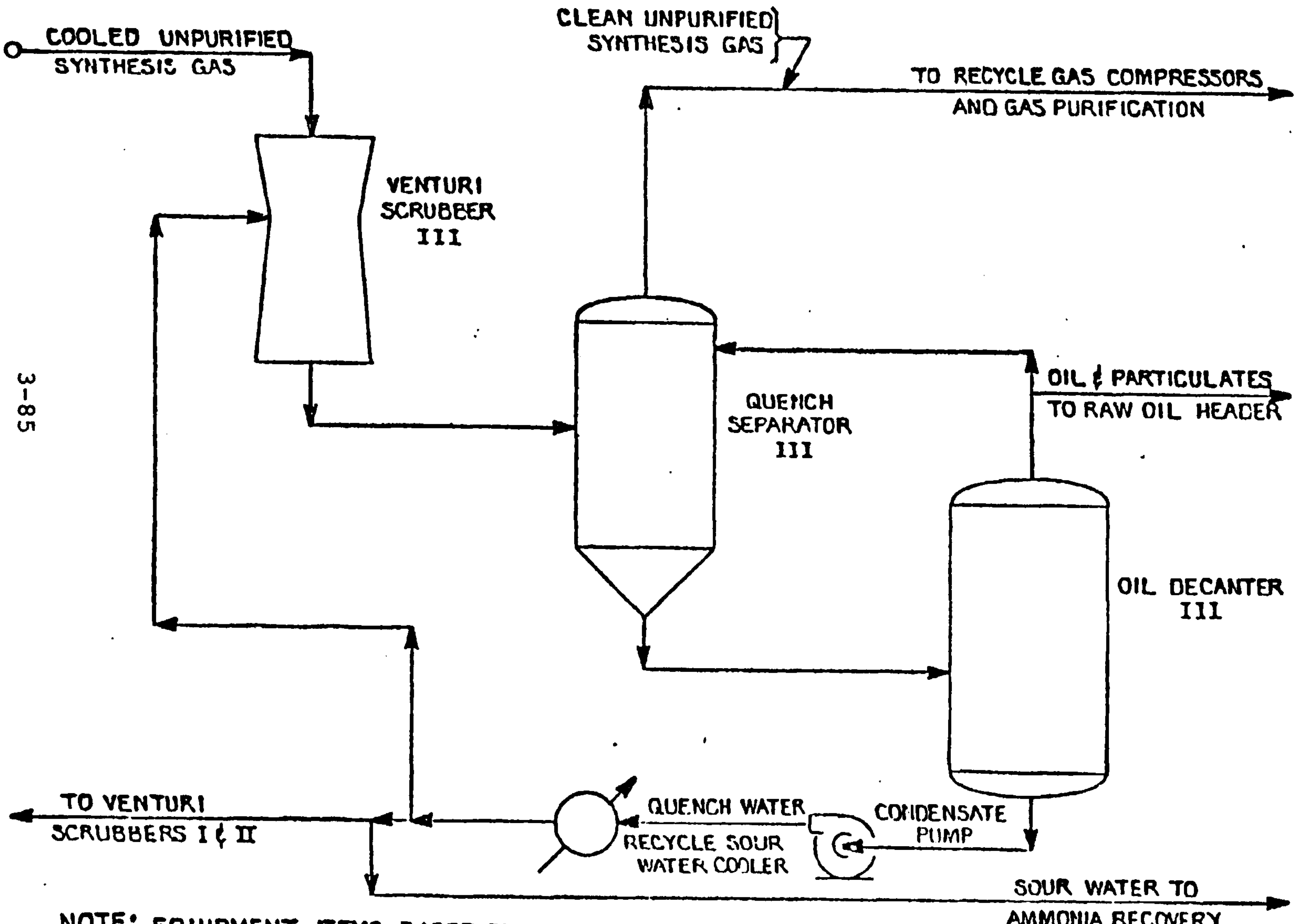
In Case A, the third stage Venturi system operates similarly to the other two Venturi stages and is shown in Figure 3.3.2-1. The mixture leaving the Venturi scrubber is separated in the quench separator. Gas leaving the quench separator is then sent to the recycle gas compressor, or the gas compressors in Area 105, before the acid gas removal system.

In the alternate Case B, the gas leaving the second stage Venturi scrubber system goes through a pressurized electrostatic precipitator for final cleaning of the pyrolysis gas as shown in Figure 3.3.2-2. Gas leaving the precipitator enters a final separator to remove the coalesced liquids. The gas leaving the separators is then sent to downstream compressors, as in the Venturi case.

Figure 3.3.2-1

Process Flow Scheme

Third Stage Venturi Scrubber

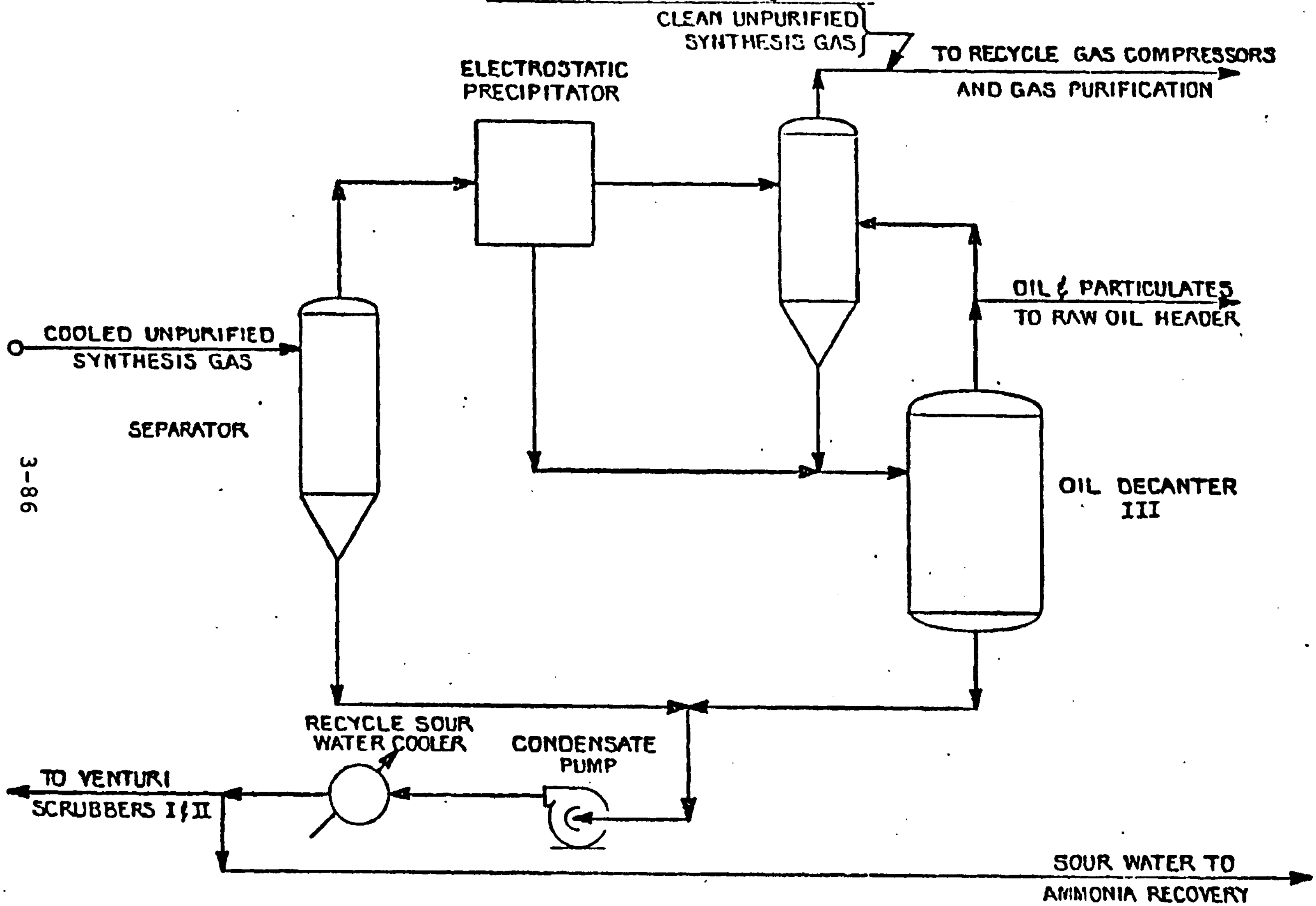


NOTE: EQUIPMENT ITEMS BASED ON TBD FLOWSHEET NO. 104-100-1.

Figure 3.3.2-2

Process Flow Scheme

Electrostatic Precipitator



NOTE: SUBSTITUTE EQUIPMENT ASSEMBLY BASED ON TBD FLOWSHEET NO. 104-100-1

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.2 Venturi Scrubber vs. Electrostatic Precipitator Design Study

Economic Comparison

Budget-type cost estimates for the equipment items required for both cases were requested and received from the Research-Cottrell Corporation. (1) (2) A comparison of incremental equipment costs of the third stage Venturi scrubber with the electrostatic precipitator is given in Table 3.3.2-2. It has been estimated that installation of an electrostatic precipitator would require an additional investment requirement of $3.1MM.

The total annual costs have also been compared for the two process schemes. The Venturi scrubber, because of its high pressure drop (55" water vs. 10"), increases the horsepower requirements of the compressors and a small amount of power is also needed to operate the water circulation pumps. The electrostatic precipitator also requires some power for its operation. The comparison of the total annual costs is given in Table 3.3.2-3. It can be seen that although the third stage Venturi uses more power (4160 KW vs. 870 KW), the large difference in the cost of capital for the total incremental plant investment offsets this. A third stage Venturi scrubber system is estimated to have an annual cost of $1.95MM vs. $2.26MM for an electrostatic precipitator. This is an annual savings of more than $0.3MM for the Venturi scrubber system.

Technical Comparison

Electrostatic precipitators and Venturi scrubbers have both been successfully used in industry to reduce particulates and mists. The electrostatic precipitator is a low pressure drop device (10" water gauge), whereas a high efficiency Venturi scrubber requires a higher pressure drop (55" water gauge). A Venturi scrubber is a simple contacting device which requires only that adequate water circulation be used in the system. Water pumps are spared to insure this requirement. For an electrostatic precipitator to work effectively, the electrical sets energizing the internal precipitator pipes must all be in a good state of repair. There are six electrical sets in each precipitator and three precipitators (one per train) are required in the Commercial Plant. (2) Research-Cottrell estimates that the operating stream factors for both units are comparable, but auxiliary equipment (pumps) are spared for the Venturi scrubbers. Sparing of equipment in an electrostatic precipitator is not done because of the large investment required.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.2 Venturi Scrubber vs. Electrostatic Precipitator Design Study

Table 3.3.2-2

Comparison of Third Stage Venturi Scrubber vs. Electrostatic Precipitator

I. Incremental Equipment Costs

	Total Horsepower (HP)	Total Delivered Equipment Cost $MM (6/1978)
A. Third Stage Venturi		
Venturi with mist eliminators	-	0.840
Water circ. pumps (100% spared)	90	0.148
Additional compressor		
Low pressure recycle	607	0.100
High pressure recycle	1,078	0.180
Gas purification	3,800	0.418
Quench separators	-	0.075
TOTAL	5,575	1.761
B. Electrostatic Precipitator		
E.S. Precipitator	870 KW	3.120

II. Total Incremental Plant Investment*	Third Stage Venturi $MM	E.S. Precip. $MM	Diff. $MM
Venturi	3.96	-	
E.S. Precipitator	-	7.02	
Additional Installed Cost			3.06

*2.25 x Equipment Cost = Installed Cost.(3)

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.2 Venturi Scrubber vs. Electrostatic Precipitator Design Study

Table 3.3.2-3

Comparison of Third Stage Venturi Scrubber vs.

Electrostatic Precipitator

Total Annual Cost

	A Third Stage Venturi Scrubber $MM (6/1977)	B Electrostatic Precipitator $MM (6/1977)	(A-B) Difference $MM (6/1977)
1. Electric Power (4160 KW & 870 KW)	0.758	0.158	+0.600
2. Cost of capital for total incremental plant investment*	1.188	2.106	-0.918
TOTAL	1.946	2.264	-0.318 (Savings)

*Cost of capital = Investment ($) x 0.30.[3]

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.2 Venturi Scrubber vs. Electrostatic Precipitator Design Study

Technical Comparison - continued

Electrostatic precipitators at elevated pressures have had only limited use in industry. However, Research-Cottrell stated that this is not a design problem but one of economics. It was not necessary to further investigate operability problems of the pressurized electrostatic precipitator because it is already economically disadvantageous to use.

Recommendations

The process gas cleanup and particulate removal scheme used in the TBD case is for three Venturi scrubbers in series. Based on the results of this trade-off study, it is recommended that the third stage Venturi scrubber be used in the Revised Tentative Baseline Design (RTBD).

References

(1) Research-Cottrell letter November 15, 1977; information transmittal on scrubber and quencher equipment recommendations and costs.

(2) Research-Cottrell letter July 28, 1977; budget cost figure for electrostatic precipitators.

(3) L. H. Dierdorff to Route List, CDC Memo, "Economic Factors for Trade-Off Studies," November 2, 1977. (CDC-G392)

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Summary

The oil leaving the Ebullated Bed Reactor contains char solids which escaped the cyclones in the Area 103 pyrolysis vessels. Once naphtha and some light fuel oil is separated in flash drums, the heavier oil containing these solids must be processed to produce solids-free oil. An economic and technical analysis was made of the following two alternate schemes for solids removal to determine which was preferable for the Conceptual Commercial Plant design:

A. Rotary drum vacuum precoat filtration of the naphtha-free oil as proposed in the TBD.(1)

B. A vacuum distillation system which concentrates the solids in the vacuum column bottoms, as proposed in the design prepared by Hydrocarbon Research, Inc. (HRI).(2)

Distillation is economically more attractive than filtration for producing a solids-free fuel oil product. Plant investment and annualized costs are much lower for distillation than for filtration. The economics of the two different process schemes are summarized in the Table 3.3.3-1. Because of the absence of filtration data on these product oils, the filtration scheme was evaluated at the two different assumed filtration rates indicated.

Table 3.3.3-1

Economic Comparison of Alternates

Alternate Case	Investment $MM (6/1977)	Total Annual Cost $MM (6/1977)
A. Filtration (11 gal/h/ft^2 flux)	10.2	4.4
A. Filtration (22 gal/h/ft^2 flux)	8.7	4.0
B. Distillation	4.6	2.2
Savings		
Distillation over Filtration (11)	5.6	2.2
Distillation over Filtration (22)	4.1	1.8

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Summary - continued

Distillation is technically preferable to filtration since filtration of coal-derived oils is still in the developmental stage. The proposed distillation scheme is based on highly developed technology used in petroleum refining operations for producing different temperature range fuel oil fractions. Solids are removed by distillation as a pumpable slurry, whereas a filter cake is produced in the filtration scheme. Solids removed by filtration are removed on conveyors and sent to auxiliary power generation as a supplementary fuel. The slurry obtained from distillation can be used in the process in the Combustor of the Gasification section. Because filters are high maintenance items, they must be periodically disassembled. This presents safety and health problems if extreme care is not taken to avoid exposure to oil fumes or contact with the oil.

Another advantage of distillation over filtration is the improvement in the product quality. The total liquid product from distillation has lower nitrogen and iron levels than the product produced by filtration. In addition, the heavy ends (+900°F) have been removed by the distillation.

Seal oil, required for operation of the Ebullated Bed Reactor, is also produced from the process oil in the distillation scheme. In the filtration scheme, seal oil must be purchased and dried to specifications.

Based on the economic and technical advantages vacuum distillation has been shown to have, the proposed design prepared by HRI is recommended for producing solid-free fuel oils and the seal oil needed for the Ebullated Bed Reactor. This design should be used in the RTBD and other future designs.

Performance Specifications

The main solids removal performance specification is to produce a fuel oil product containing a maximum of 0.1 weight percent solids, the maximum sediment level permitted for No. 2 grade fuel oils according to ASTM specifications.[3] This removal of solids from the fuel oil must be obtained with a minimum loss of oil.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Performance Specifications - continued

In the distillation scheme, the vacuum gas oil column produces the seal oil required for operation of the Ebullated Bed Reactor. This oil must meet the HRI specifications for moisture, boiling point range and dielectric properties to make it suitable for use in the canned motor pumps required for the Ebullated Bed Reactor. Since the filtration scheme does not produce a seal oil, an equivalent oil must be purchased from a refinery and dried over molecular sieves to specification.

Discussion

When the COED Pilot Plant was in operation, raw pyrolysis oils containing 4 to 10 weight percent solids were filtered in the COED Pilot Plant filtration system. A 6 square foot rotary drum pressurized precoat filter was used in the COED Pilot Plant. This filtering system operated very successfully on six different oils derived from coals ranging from subbituminous to high volatile A bituminous. In all cases, the solids content in the filtrate was less than 0.1 weight percent.

Filtration of the raw pyrolysis oils was performed at 350°F (4) in order to achieve reasonable filtration rates (10 gal/h/ft^2). At 350°F the viscosity of the pyrolysis oil was less than 5 centipoise. In the process scheme being evaluated, the oil being filtered has been hydrogenated in the Ebulated Bed Reactor which reduces the oil viscosity. Since the filtration rate is proportional to the viscosity, it was estimated that the flashed oil from the Ebullated Bed Reactor has fluid properties at 150°F equivalent to the COED oils at 350°F. Because actual filtration rate data were not available on the ebullated bed product, two filtration rates (11 and 22 gal/h/ft^2) were considered in this trade-off study to determine the cost effect of filtration rate.

The lower value represents a conservative value for which there is some experimental basis. The higher value represents an optimistic rate that might be achievable by optimization of filtration conditions if sufficient oil was available to conduct a thorough experimental program. While there is a significant cost effect of filtration rate, neither assumption gives the filtration case an economic advantage. A vacuum precoat filter was considered instead of the pressurized precoat rotary drum since it is considerably cheaper (about 1/3 the cost).

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Discussion - continued

In the HRI design prepared under a lower tier subcontract to CDC, a distillation scheme is used which concentrates the particulates in a +900°F heavy oil fraction. All fuel oils are taken as distillation products with the heavy oil and solids being removed as a pumpable oil slurry. The distillation scheme has the added advantage of producing an intermediate vacuum gas oil which is used as the seal oil required for operation of the Ebullated Bed Reactor. Since no seal oil is produced by the filtration scheme, it must be purchased from a refinery. The distillation process scheme proposed by HRI is commonly used in the petroleum industry to separate crude oil into different boiling range fractions.

Process Description

The ebullated bed hydrotreating process is shown as a block on the Process Flow Schemes, Figure 3.3.3-1 and Figure 3.3.3-2. The process, in the block, involves passing the liquid pyrolysis oil product through the Ebullated Bed Reactors with the vapor and liquid phases formed and the phases being separated in High and then Low Pressure Flash Drums. The vapor stream, containing naphtha and light fuel oil, is further processed to recover the naphtha and fuel oil. The liquid streams leaving the flash drums are mixed and contain the solids carry-over from hydrotreating. It is this mixture of oil products that is to be processed by either distillation or filtration to remove the solids.

Figure 3.3.3-1

Process Flow Scheme

Oil Distillation

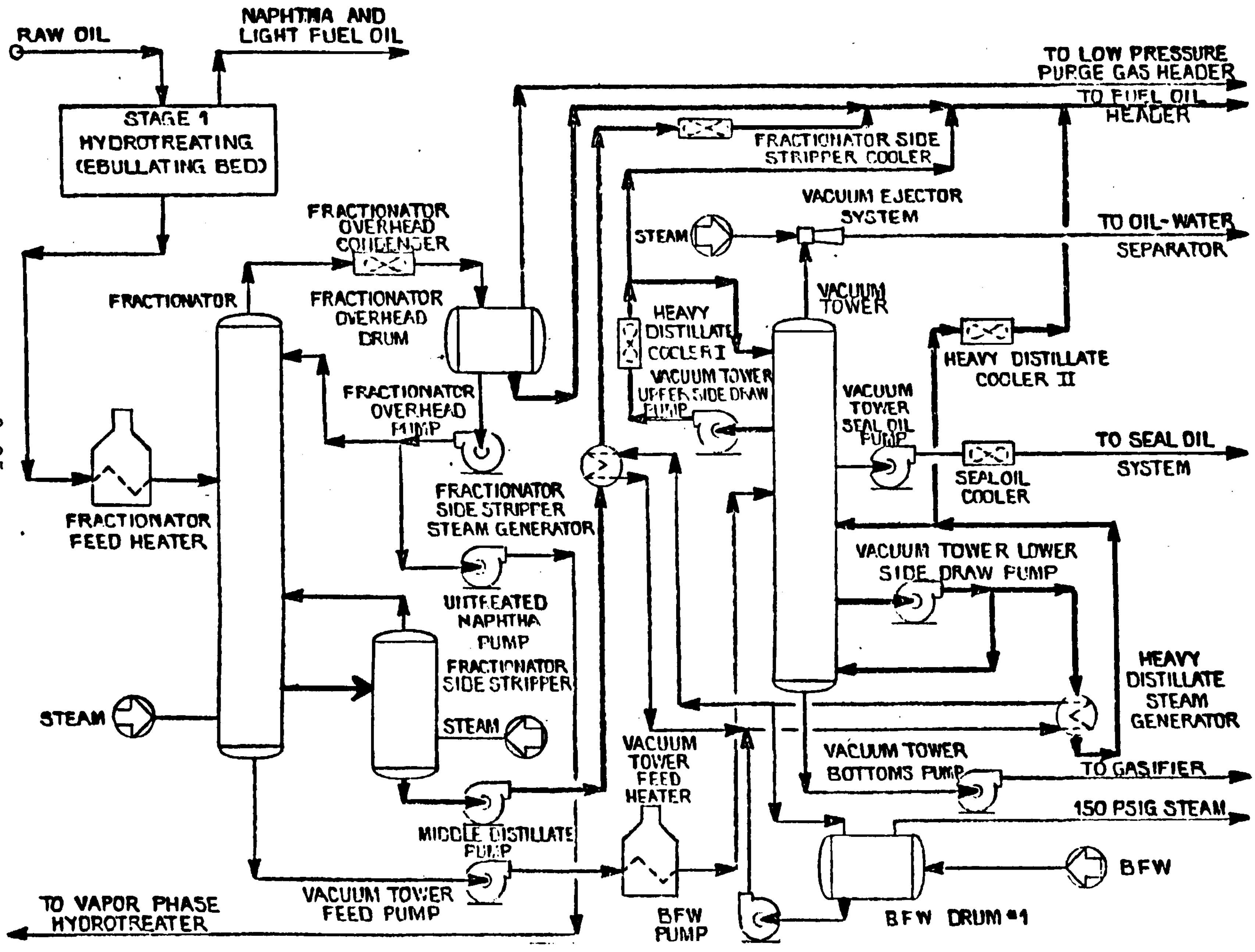
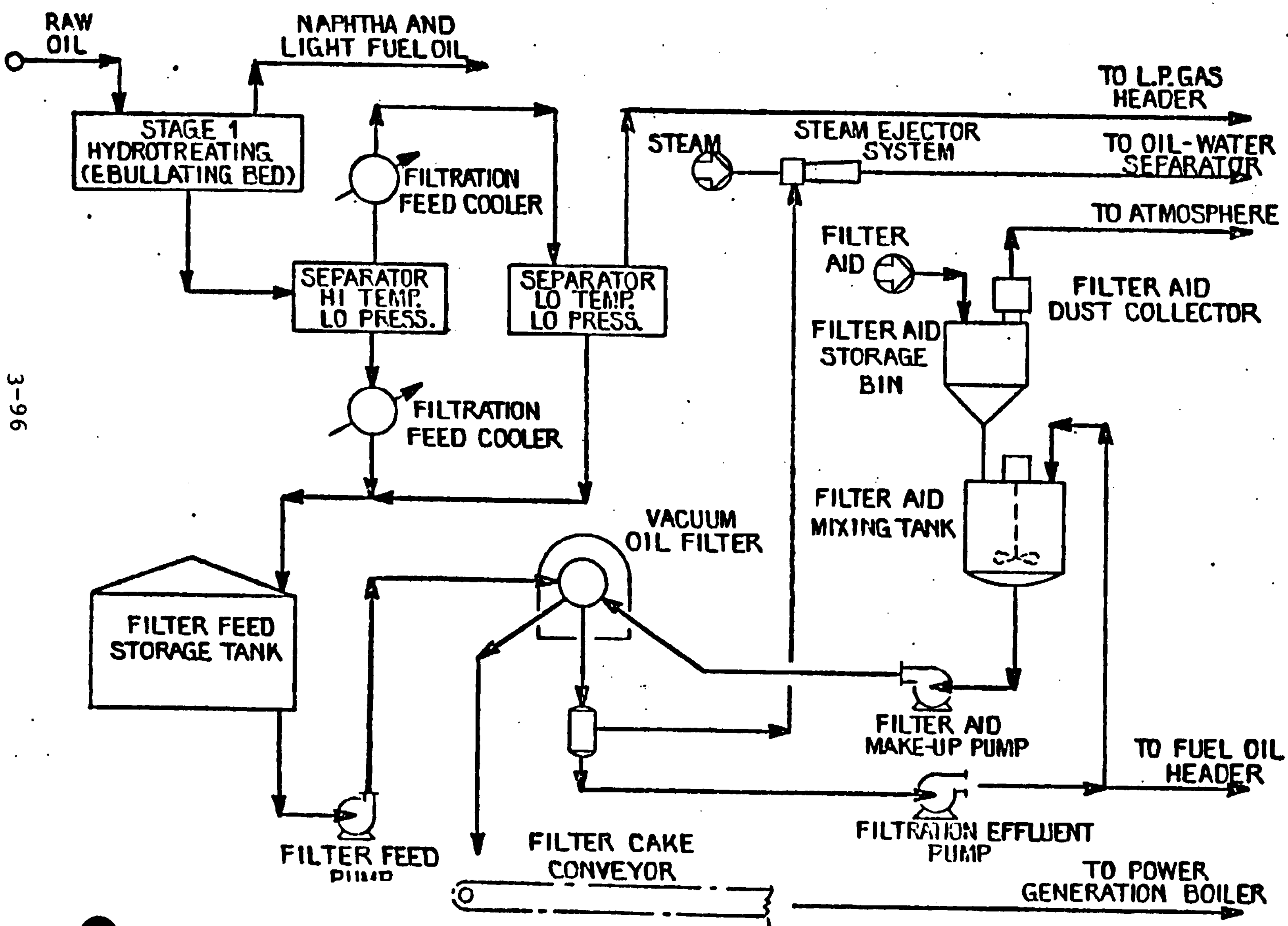


Figure 3.3.3-2

PROCESS FLOW SCHEME

OIL FILTRATION



3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Process Descriptions - continued

Case A - Filtration

As shown in Figure 3.3.3-2, the oil containing the solids is reduced in pressure and the flashed mixture is separated in the High Temperature - Low Pressure Separator. Gases leaving this separator are cooled and fed to the Low Temperature - Low Pressure Separator. Gases from this second separator go to the Low Pressure Gas Header. Condensed oils from this separator are mixed with the cooled, condensed oils from the High Temperature Separator and the mixed oil is sent to intermediate storage and then to filtration. Rotary Drum Precoat Vacuum Filters are used to remove the solids from the oil. The solids-free fuel oil product is then sent to the Fuel Oil Headers. Since the filter cake cannot be pumped, it is sent by conveyors for burning in the auxiliary power generation system.

Case B - Distillation

In this process scheme, shown in Figure 3.3.3-1, additional heat is added to the oil stream containing the solids in the Feed Fractionator Furnace. The preheated feed is reduced in pressure directly ahead of the Fractionator. Water, fuel gas, and naphtha are removed in the column overhead and separated in the Overhead Drum. A middle distillate side stream is removed from the Fractionator and distilled in a Side Stream Stripper. The middle distillate fuel oil is recovered from the Side Stream Stripper as the bottoms product. The bottoms from the Fractionator are sent to the Vacuum Tower after preheating in the Vacuum Tower Feed Heater. An overhead and two side stream fuel oils are obtained from the Vacuum Tower. In addition, the seal oil needed for the operation of the Ebullated Bed Reactor is obtained as a side stream from the Vacuum Tower. The bottoms removed from the Tower is the +900°F oil residue containing the solids. This oil/solids slurry is pumped to an intermediate storage tank in the Gasification section, Area 103, where it will be consumed as fuel.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Economic Comparison

Estimated capital investment requirements for the filtration and distillation schemes, Cases A and B respectively, are summarized in Tables 3.3.3-2 and 3.3.3-3. To estimate the cost of the filtration area, the equipment costs were estimated from the equipment costs given in the ICGG Proposal(1) for the Conceptual Commercial Plant. Since filtration data on these hydrotreated oils were not available, two different filtration rates were considered for determining the cost effect on capital investment. The higher filtration rate results in reduced filtration area and fewer filters required. The investment costs for the distillation scheme were obtained from distillation cost data on petroleum refining plants.(5) The comparison of plant investment shows the distillation scheme required about one half of the capital investment for the more optimistic filtration scheme (Table 3.3.3-4).

The total annual cost (Table 3.3.3-5), which includes the capital related costs, chemicals, and utilities, is much lower for the distillation scheme. Chemicals and utilities for filtration will cost over $500,000 more than for distillation. In addition, the higher investment cost for filtration results in higher capital related costs.

In the filtration scheme, seal oil must be purchased for operation of the Ebullated Bed Reactor; whereas in the distillation scheme, the seal oil is obtained from the Vacuum Tower. The seal oil is recovered and ultimately becomes part of the fuel oil product, which in the filtration scheme increases the fuel oil yield; however, the fuel oil value is $1.40 per barrel less than the cost of the seal oil ($15.40 vs. $16.80). Consequently, the cost difference of $540,000 between the purchased seal oil and the additional fuel oil product must be taken as an operating expense. The only costs for Case B that are higher than those for Case A are for steam and fuel, totaling $550,000 higher.

The result of the economic comparison is that the distillation scheme offers a capital saving of $4.1 - $5.6MM and an annual cost saving of $1.8 - $2.2MM.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Table 3.3.3-2

Filtration Scheme, Case A
Estimated Plant Investment

Assumed Filtration Rate - $gal/hr/ft^2$	11	22
Item	Equipment Cost, $MM	Equipment Cost, $MM
Filtration Feed Coolers	0.284	0.284
Hi Temp., Low Press. Flash Separators	0.030	0.030
Low Temp., Low Press. Flash Separator	0.020	0.020
Unfiltered Oil Storage	0.674	0.674
Rotary Drum Vacuum Filters (7) (5)[1]	0.910	0.570
Steam Jets With Condenser	0.171	0.123
Filtration K.O. Drums	0.032	0.023
Filter Aid Storage	0.035	0.035
Filter Aid Dust Collector	0.035	0.035
Filter Aid Mixing Tank	0.024	0.024
Product Oil Surge Tank	0.070	0.070
Seal Oil Storage	0.177	0.177
Filter Cake Conveyor	0.050	0.050
Misc. Pumps, Mixer and Tanks	0.100	0.100
Total Equipment Cost (12/75)[4]	2.612	2.215
Add Cost Escalation to 6/77 @ 12%[2]	0.313	0.266
Total Equipment Cost (6/77)	2.925	2.481
Total Plant Investment (3.5 x Total Equipment Cost)[3]	10.240	8.683

1) Number of filters required.
2) CE Plant Cost Index.
3) The factor was recommended by Dravo to estimate installed cost from equipment cost in the filtration area.
4) Equipment costs from ICGG Proposal.(7)

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Table 3.3.3-3

Distillation Scheme, Case B
Estimated Plant Investment

	Investment Cost, $MM
Low Pressure Distillation System[1]	
Cost includes:	
Distillation Column Side Stream Stripper All battery limits process facilities Heat exchangers to cool product streams Central Control System	1.18
Vacuum Distillation System[1]	
Cost includes:	
Distillation Column All battery limits process facilities Three stage steam jet system Heat exchangers to cool the product streams Central Control System	0.82
Two additional side streams at 10% for each	0.16
Total Investment (6/73 basis)	2.16
Add Cost Escalation[2]	
6/73 to 1/75 @ 38%	0.82
1/75 to 6/77 @ 27%	0.81
Total Investment (6/77, Gulf Coast)	3.79
Location Factor (Illinois) - add 20%	0.76
Total Investment (6/77, Illinois)	4.55

1) Costs obtained from Reference (5), June, 1973 basis.
2) Cost escalation factors from Central Engineering, FMC Corporation.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Table 3.3.3-4

Comparison of Plant Investment,
Distillation vs. Filtration

Plant Investment	$MM
Case A	
Filtration (11 gal/hr/ft^2)	10.2
Filtration (22 gal/hr/ft^2)	8.7
Case B	
Distillation	4.6
Differential Investment	
Savings, Distillation over Filtration (11)	5.6
Savings, Distillation over Filtration (22)	4.1

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Table 3.3.3-5

Annual Cost Comparison
Filtration vs. Distillation

	Case A Filtration $MM/yr	Case A Filtration $MM/yr	Case B Distillation $MM/yr
Filtration Rate gal/hr/ft^2	11	22	-
Chemical and Utilities			
Steam (600 psi $1.56/1000 lb) (150 psi $1.23/1000 lb)	0.10	0.10	0.21
Electric Power 2.3¢/KWH	0.02	0.02	0.07
Cooling Water 7¢/1000 gal	0.06	0.06	0.08
Filter Aid $125/ton	0.64	0.64	-
Seal Oil $16.80/gal	0.54	0.54	-
Fuel $2.50/MM Btu	-	-	0.44
Total Chemical and Utilities	1.36	1.36	0.80
Cost of Capital for Investment*	3.06	2.61	1.38
Total Annual Cost	4.42	3.97	2.18
Annual Savings			
Distillation over Filtration (11)			2.2
Distillation over Filtration (22)			1.8

*Cost of Capital = Investment ($) x 0.30.[8]

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Technical Comparison

Filtration of heavy oils is still in the development stage. Fuel oils leaving the Ebullated Bed Reactor would be expected to filter more easily than the COED raw oil. However, adequate experimental data on fuel oil filtration were not available. Filters are high maintenance items which must be periodically opened and the oil/solids filter cake must be handled in conveyors. This presents a safety and health problem which is not present in closed equipment used in the distillation scheme. The largest available rotary drum vacuum filters have about 500 ft^2 of surface. Because of the low filtration rates obtained with coal-derived oils, multiple filters will be required to operate in parallel. In addition to the filters on filter cycle, one filter would be operating simultaneously on precoat cycle. One spare filter must also be available. For the filtration rates assumed in this study, the number of filters and filter area required are given in Table 3.3.3-6.

Table 3.3.3-6

Filtration Rate vs. Number of Filters Required

	Filtration Rate	
	11 $gal/hr/ft^2$	22 $gal/hr/ft^2$
Total No. of Filters	7	5
Filtration Area/Filter-ft^2	500	400
No. of Filters in Operation at Plant Capacity	5	3

Solids from the distillation scheme are removed as a pumpable oil/solids slurry. In filtration, the filter cake obtained must be removed by a conveyor system. The filter cake is used as a supplementary fuel in the auxiliary boiler for power generation, whereas the oil/solids slurry may be burned in the process in the Combustor of Pyrolysis and Gasification, Area 103.

The product quality of the fuel oils obtained by the distillation scheme is superior to the oil obtained by filtration since all fuel oils are taken as distilled product fractions. Both schemes give an acceptable level of solids (<0.1 wt. %) in the product. This is based on the recommended level of sediment of No. 2 and heavier

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

Technical Comparison - continued

fuel oil as specified in the ASTM standards manual.[3] The heavy metals (i.e. Fe) are almost completely removed from the fuel oil product obtained by distillation. In addition, the +900°F material is also removed from the fuel oil obtained by distillation but remains in the filtered product oil. In distillation, this +900°F oil is removed with the solids, whereas in filtration the oil removed with the solids contains the total fuel oil range product (400°F to EP).

Recommendations

Since both the economic and technical factors favor the distillation process scheme, it is recommended that this design, as proposed by HRI, be utilized in the RTBD and in subsequent plant designs.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.3 Oil Filtration vs. Vacuum Distillation Design Study

References

(1) ICGG, "Pipeline Gas Demonstration Plant - Phase I, Tentative Baseline Design - Conceptual Commercial Plant," Report No. FE-2012-0004, ERDA Contract No. EF-77-C-01-2012, August 5, 1977.

(2) Hydrocarbon Research, Inc., "H-Oil and Hydrotreating of COGAS Pyrolysis Oil," Engineering Design Package, COGAS Commercial Plant, February 2, 1978.

(3) ASTM, "Annual Book of ASTM Standards - Petroleum Products and Lubricants," Part 23, 1976.

(4) "Char Oil Energy Development - Project COED," ERDA R & D Report No. FE/1212/F, Volume 1, Final Report for the Period August 18, 1971 through June 30, 1975, Contract No. E (49-18)-1212, September, 1975.

(5) Gary, J. H. and G. E. Handwerk, "Petroleum Refining Technology and Economics," Marcel Dekker, Inc., New York (1975).

(6) Nongbri, G. and J. Gendler, "H-Oil Processing of COED Pyrolysis Oil from Western Kentucky Coal for COGAS Development Company - Runs 84-267 and 84-268," Hydrocarbon Research, Inc., Report No. L-12-58-100-501, January 15, 1978.

(7) ICGG Proposal for Pipeline Gas from Coal Demonstration Plant Program, January 21, 1976; in response to ERDA RFP No. E(49-18)-2012.

(8) L. H. Dierdorff to Route List, CDC Memo, "Economic Factors for Trade-Off Studies," November 2, 1977 (CDC-G392).

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naptha vs. Syncrude Design Study

Summary

In the TBD, two stages of hydrotreating are used to upgrade the raw pyrolysis oil to a high quality synthetic crude oil (Case A). Stage I is an ebullating bed hydrotreating step, and Stage II is a fixed-bed hydrotreater. An alternte process scheme (Case B) was considered which eliminates the second stage, fixed-bed hydrotreater and produces low-sulfur fuel oil and naphtha products instead of syncrude. This configuration requires less hydrogen and reduced investment in hydrogen plant and hydrotreating, while producing liquids with a higher market value than syncrude.

The results of the study indicate technical and economic advantages for the fuel oil-naphtha case. Fixed capital requirements are reduced by approximately $60MM. Annual cost savings of $19.5MM, mainly capital-related, will be realized, while annual sales revenue will be increased by $14.1MM/yr, mainly because of higher market prices for fuel oil and naphtha than for syncrude. The combined effect is a $33.6MM annual cost advantage for the fuel oil and naphtha case. This reduces the required selling price of the pipeline gas product by approximately 41¢/MM Btu. Overall process thermal yield is increased by 0.25 percentage point, and coal feed rate for the same pipeline gas production rate is reduced by about 1.0 percent.

Although the Case B products are not as highly upgraded as the Case A syncrude, below, they meet current product specifications.

Table 3.3.4-1

Product Properties - RTBD

	Gravity °API	Sulfur wt %	Nitrogen wt %
Raw Oil	-4	2.1	1.1
Fuel Oil	13.9	0.03	0.36
Naphtha	49.0	<0.01	<0.1
Syncrude	27	0.04	0.10

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Summary - continued

It is recommended that the syncrude process scheme (Case A) used in the TBD be replaced by the fuel oil-naphtha configuration (Case B) for the TRBD, because of the substantially better economics, and because of process simplification realized from the elimination of second stage hydrotreating.

Performance Specifications

Raw pyrolysis oil, after recovery and dehydration, is a heavy oil which contains char solids, and is pumpable at or above 150°F.

Hydrotreating of the raw oil is carried out for the following purposes:

Improve salability and value (price) of oil product.

Reduce sulfur and nitrogen content to environmentally acceptable levels.

Reduce oil viscosity and increase the API gravity, thus improving pumpability and storage stability.

Facilitate removal of char solids in the oil.

In the TBD, two stages of hydrotreating (ebullating and fixed-beds) were employed to produce a 27° API gravity syncrude with 0.04 percent sulfur and 0.10 percent nitrogen. However, reduced depth of hydrotreating in a single stage, with lower hydrogen consumption, also produces high value products which meet the purposes listed above, and at lower cost.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Process Description

Two different process schemes are considered in this trade-off study to determine the preferred low sulfur, marketable oil by-products. These are:

Case A - Syncrude, as designed in the TBD.

Case B - Fuel oil and naphtha.

The TBD is the basis for this study. The major differences between the two cases are in Area 104, Oil Recovery and Treatment, although other process areas are affected to some extent.

For both cases, the raw oil vapors and sour water liquor contained in the stream form the pyrolysis-gasificiation process step are condensed and separated. The collected oil is dehydrated and hydrotreated, and the liquor is sent to Area 113, Ammonia Recovery.

In the syncrude case, the raw oil is first hydrotreated in an ebullating bed hydrotreating stage. Then the low viscosity hydrotreated oil from the first stage, is stripped with steam to remove C_1-C_4's and a naphtha fraction, and is filtered to remove solids. Part of this filtered oil (about 20 percent in the TBD) is sent to in-plant uses which require fuel oil for heating. Most of the oil is then upgraded to syncrude in a fixed-bed hydrotreater. Gas, light hydrocarbons, and liquor produced in the fixed-bed hydrotreater are removed from the syncrude, and the naphtha fraction is added back as part of the syncrude. The final syncrude oil is composed of oil from the phase, fixed-bed hydrotreater and naphtha.

In the fuel oil-naphtha case, the raw oil collected from the pyrolysis-gasification process stream is also dehydrated and then hydrotreated in the ebullating bed hydrotreating stage. Gas, light hydrocarbons, and naphtha are removed and separated from the fuel oil. As in the TBD, a vacuum filter is employed to remove the solids contained in the hydrotreated oil. However, filtered oil, naphtha, and light hydrocarbons are separated as by-products of this process. Off-gas is combined with raw synthesis gas going to the Gas Purification Area, Area 105.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Process Description - continued

The impact of eliminating the fixed-bed hydrotreater in Area 104 on other areas of the process is either enlargement or reduction in sizes of units or equipment related to the decreased hydrogen consumption and off-gas produced in hydrotreatment. The effect on sizing in the other process areas are summarized below:

Area 105, Gas Purification. The major gas feed to this area is raw synthesis gas which is the same in both cases since the same amount of coal input is assumed for each. Off-gas from hydrotreatment is a secondary gas feed to this area. In Case B, off-gas is only about 80.5 percent of Case A because of elimination of the fixed-bed hydrotreater. But total gas feed in Case B is only 0.5 percent less than in Case A. The resulting yield of light hydrocarbons recovered from this gas stream in Case B is 95.1 percent of Case A.

Area 106, Hydrogen Generation. Since there is no fixed-bed hydrotreater in Case B, the hydrogen requirement for hydrotreating is only 84.3 percent of Case A. Only 86 percent as much purified synthesis gas as in Case A is consumed in generating hydrogen gas feed to the ebullating bed hydrotreating stage in Case B.

Area 107, Shift and Methanation. Ther is about 1.5 percent more synthesis gas in Case B for feed to shift and methanation. This produces about 1.1 percent more high Btu pipeline gas in Case B.

Area 108, Bulk CO_2 Removal. Because of an increased amount of shift and methanation reactions, approximately 2.6 percent more CO_2 will be removed in Case B.

Area 109, Gas Compression and Dehydration. The result of increased shift and methanation gives 1.1 percent more enriched pipeline gas in Case B for compression and dehydration.

Area 112, Claus Plant. The elimination of the fixed-bed hydrotreater in Case B reduces H_2S in the off-gas by 1.2 percent; thus, elemental sulfur recovered in Case B is only 99 percent of Case A.

Area 113, Ammonia Recovery. Dry ammonia produced in Case B is 13 percent less than in Case A.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Process Description - continued

The preceding process descriptions are based on the TBD for the syncrude case, and on modifications of it for the fuel oil-naphtha case. The RTBD process design will have significant differences in the hydrotreating area, such as a distillation scheme instead of filtration for removal of solids in the oil. Also, less fuel oil will be consumed by in-plant uses; this currently is expected to be only a third as much as the 20 percent consumed in the TBD. These differences in the TBD process design, however, will have no significant effect on the comparative economics and conclusions of this trade-off study.

Technical Comparison

The fuel oil-naphtha configuration has the advantage of being less complex than the syncrude case because second stage hydrotreating is eliminated.

Thermal yield, defined as higher heating value of products divided by the higher heating value of total coal input, is calculated and listed in Table 3.3.4-2. The results show that the thermal yield of the fuel oil-naphtha case is 0.25 percentage points higher than the thermal yield of the syncrude case.

Economic Comparison

The syncrude case was based on catalyst and chemical costs, equipment costs, and capital investment data in the ICGG Proposal,[1] and on material balances in the TBD.[2] A revised material balance was prepared for the fuel oil-naphtha case, using preliminary hydrotreating data from HRI.[6] This was used as the basis for estimating differences in capital and operating costs, annual revenue, and the effect on required selling price of the pipeline gas product.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Table 3.3.4-2

Process Thermal Yield

	Case A Syncrude		Case B Fuel Oil-Naphtha	
	10^9 Btu/sd	Percent	10^9 Btu/sd	Percent
Coal input (28,733 ton/sd)	633.3 (1)	100.0	633.3 (1)	100.0
Products output:				
Pipeline gas (954 Btu/SCF) (2)	244.2	38.56	246.7	38.95
Syncrude (23,044 bbl/sd)	136.1	21.49	-	-
Fuel Oil (19,395 bbl/sd)	-	-	116.5	18.40
Naphtha (3,700 bbl/sd)	-	-	19.7	3.11
Light hydrocarbons	15.2	2.40	14.2	2.24
Total energy products	395.5	62.45	397.1	62.70

(1) Based on 60°F HHV of feed coal less the HHV of sulfur content.

(2) Including 30 percent of the total light hydrocarbon production blended in to raise the higher heating value to 954 Btu/scf, dry basis.

Case A figures are for the TBD. Production rates are from Table 3.3.4-2.

Notation: sd = stream day

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Economic Comparison - continued

Capital Cost

Equipment costs for the syncrude case (Case A), listed in Table 3.3.4-3A, are taken from the ICGG Proposal.(1) For the fuel oil-naphtha case (Case B), equipment was omitted or resized according to process stream flow rates. Case B equipment costs, shown in Table 3.3.4-3A, were estimated from Case A figures using cost-capacity exponents which range from 0.65 to 0.8,(4) depending on the equipment item or process area. Only plant areas which differ in the two cases are listed. The dominant cost reduction is in Area 104, Oil Recovery and Treatment. There are smaller, but significant, savings in the Hydrogen Plant and Product Storage areas. Differences in other process areas are insignificant, however the figures are presented in Table 3.3.4-3 for illustration. The total differential cost between Case A and Case B is a $24.5MM saving in equipment cost for the alternate Case B.

This advantage in equipment cost yields a plant investment saving in 6/1977 dollars of $61.8MM, including adjustment for 12 percent inflation from year-end 1975 based on the CE Plant Cost Index. Annual capital-related cost saving, at 0.30 times the plant investment saving, is $18.5MM/yr (Table 3.3.4-3B). The 30 percent factor used for total annual capital-related costs includes all financial charges (depreciation, interest on debt, return on equity, income tax, etc.) as well as maintenance, insurance, local taxes, and the cost effects of capital-related working capital, construction loan interest and start-up costs.(5)

Operating Cost

The same coal input, 28,733 tons per stream day, was used in both the syncrude and fuel oil-naphtha cases. Size and cost of most process areas remain the same in the two cases. The major differences are in the hydrotreating step, which then affect other process areas because of differences in hydrogen feed and hydrotreater off-gas rates. Coal and operating labor costs were assumed to be the same in both cases.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Table 3.3.4-3

Capital Costs

A. Equipment Cost (12/1975 Dollars)

	Process Area	Case A Syncrude $MM	Case B Fuel Oil-Naphtha $MM
104.	Oil Recovery & Treatment	75.08	52.12
105.	Gas Purification	44.69	44.43
106.	Hydrogen Plant	8.02	7.23
107.	Shift & Methanation	20.14	20.32
108.	Bulk CO_2 Removal	5.73	5.83
109.	Gas Compression & Dehydration	4.43	4.47
112.	Sulfur Recovery	12.63	12.55
113.	Ammonia Recovery	1.07	0.98
120.	Product Storage	18.13	17.48
		189.92	165.41
	Equipment cost saving for Case B		24.51

B. Annual Capital-Related Cost Saving

Plant investment saving, 6/1977 dollars (1)	$61.8 MM
Annual capital-related cost saving (2)	$18.5 MM/yr

Case A figures from the ICGG Proposal, 12/1975 dollars.

(1) Factored from equipment cost saving, including installation cost (factor = 2.25) and escalation, 12%, from 12/1975.
($24.5 MM) (2.25) (1.12) = $61.8 MM

(2) At $61.8 MM x 0.30.[5]

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Economic Comparison

Operating Cost - continued

Substantial reductions in the consumption and cost of catalysts and chemicals, especially in hydrotreating and hydrogen generation, are realized in the fuel oil case. Raw materials consumption and costs for Case A were taken from the ICGG Proposal;[1] for Case B, they were estimated on the basis of material balance differences from Case A.

The differential cost of raw materials in the affected process areas was found to be in favor of the fuel oil-naphtha case by $1.0MM per year.

Cost Summary

In addition to reduced capital and operating costs, the fuel oil-naphtha design yields increased sales revenue. This is due primarily to higher market prices for fuel oil and naphtha than for syncrude. Product yields and output rates for the two cases are shown in Table 3.3.4-5. Corresponding thermal yields are presented in Table 3.3.4-2, showing an 0.25 percentage point (0.4 percent) advantage for the fuel oil case. Small reductions in sulfur, NH_3, and light hydrocarbon production result from the reduced depth of hydrotreating in Case B.

Comparative sales revenue in Table 3.3.4-5 shows a $14MM/yr advantage for the alternate fuel oil-naphtha case. This is based on unit product prices requested from and supplied by ICGG.[5-9]

The overall economic summary in Table 3.3.4-6 shows a cost advantage of $33.6MM/yr for the fuel oil-naphtha design. This reduces the required selling price of pipeline gas by 41¢/MM Btu.

The sensitivity of this gas price advantage to variations in co-product prices is shown in Table 3.3.4-7. These prices may vary with plant location and with changing market conditions. The sensitivity factors can be used to adjust the gas price differential. The cost advantage of fuel oil-naphtha varies directly as the price spread between fuel oil-naphtha and syncrude. If, for example, the

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Table 3.3.4-4

Saleable Products

	Case A Syncrude	Case B Fuel Oil-Naphtha
Coal input, ton/sd	28,733	28,733
Products:		
Pipeline gas, 10^9 Btu/sd	244.2	246.7
Syncrude, bbl/sd	23,044	-
Fuel oil, bbl/sd	-	19,395
Naphtha, bbl/sd	-	3,700
Light hydrocarbons, lb/sd	698,000	652,000
10^9 Btu/sd	15.2	14.2
Sulfur, LT/sd	831	823
NH_3, ton/sd	69.6	60.5
Sodium sulfate, ton/sd	63.1	63.1

Case A figures are from the TBD,[2] (6/1977).

Notation: sd = stream day
LT = long ton

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Table 3.3.4-5

Incremental Revenue

Products	Unit Price	Annual Revenue Items A Syncrude $MM/yr	B Fuel Oil-Naphtha $MM/yr
Syncrude	$13.80/bbl	104.9	-
Fuel Oil Naphtha	$15.40/bbl	- -	98.6 18.8
Light hydrocarbons	$ 0.08/lb	18.4	17.2
Sulfur	$43/LT	11.8	11.7
Ammonia	$130/ton	3.0	2.6
		138.1	148.9
Pipeline gas, 2.5×10^9 Btu/sd increased output @ $4/MM Btu (1)		Base = 0	3.3
		138.1	152.2
Incremental annual sales revenue			$14.1 MM

Case A figures are for the TBD.(2)

(1) Based on pipeline gas production rates in Table 3.3.4-2; and on assumed, typical price of $4/MM Btu (6/1977).

Notation: LT = long ton

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Table 3.3.4-6

Economic Summary

(6/1977 Dollars)

Cost Savings with Fuel Oil-Naphtha Products Instead of Syncrude

Fixed capital reduction $61,800,000

Item	Savings $MM/yr
Sales revenue increase:	
By-products	10.8
Pipeline gas, increased output	3.3
Total added sales revenue	14.1
Capital-related cost reduction	18.5
Raw material cost reduction	1.0
TOTAL	33.6

Pipeline gas price reduction (on 246.7 x 10^9 Btu/sd x 330 days/yr) $0.41/MM Btu

Notation: sd = Stream day

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Table 3.3.4-7

Comparative Gas Price Sensitivity

Effect of By-product Prices

The sensitivity factors below apply only to the gas price advantage of Case B (fuel oil-naphtha) over Case A (syncrude).

An Increase of	In The	Will Change the P/L Gas Price Advantage of Case B over Case A By ¢/MM Btu
\$1.00/bbl	Syncrude Price	-9.4
\$1.00/bbl	Fuel Oil Price	+7.9
\$1.00/bbl	Naphtha Price	+1.5
\$0.01/lb	Light Hydrocarbon Price	-0.19
\$10/ton	Ammonia (NH_3) Price	-0.037
\$10/LT	Sulfur Price	-0.033

Notation: LT = long ton

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

Economic Comparison

Cost Summary - continued

price of fuel oil-naphtha should be \$0.60/bbl higher or lower than in Table 3.3.4-4 (\$14.80 or \$16.00/bbl), the gas price advantage would be 41 $\pm$ (7.9 + 1.5) (\$0.60/bbl) = 41 $\pm$ 6 = 47 or 35¢/MM Btu.

Recommendations

Based on the economic savings, which are substantial, the fuel oil-naphtha process scheme is recommended for the RTBD, replacing the syncrude configuration used in the TBD.

3.3 AREA 104 - OIL RECOVERY AND TREATMENT

3.3.4 Fuel Oil-Naphtha vs. Syncrude Design Study

References

(1) ICGG Proposal for Pipeline Gas Demonstration Plant Program, 21 January 1976; in response to ERDA RFP No. E(49-18)-2012.

(2) ICGG, "Pipeline Gas Demonstration Plant. Phase I. Tentative Baseline Design - Conceptual Commercial Plant," Report No. FE-2012-004, ERDA Contract No. EF-77-C-01-2012, August 5, 1977.

(3) R. Geosits, Bechtel Associates Professional Corporation, "Preliminary Evaluation of Hydrotreating Facilities for Coal Oil from the COGAS Process," June 1975.

(4) Bechtel Associates Professional Corporation, "Scope Definition and Order of Magnitude Estimate for COGAS Process Demonstration Plants," June 1974.

(5) L. H. Dierdorff to Route List, CDC Memo, "Economic Factors for Trade-Off Studies," November 2, 1977. (CDC-G392)

(6) G. Nongbri and J. Gendler, "H-Oil(R) Processing of COED Pyrolysis Oil from West Kentucky Coal for COGAS Development Company," HRI Report No. L-12-58-100-501, January 15, 1978.

(7) L. H. Dierdorff to R. Bloom, Jr., CDC Memo, "Preliminary Cost Estimates - COGAS Industrial Fuel/Moderate Btu Gas Processes," February 10, 1975.

(8) L. H. Dierdorff to R. Bloom, Jr., CDC Memo, "Preliminary COGAS Cost Estimate - 12,000 T/D, 2-Train Plant," September 9, 1975.

(9) D. Field to R. L. McCray, ICGG-ENG-122 Memo, "Current Market Value Analysis," November 1, 1977.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.1 Summary

Introduction

The flue gas from the combustor contains heat in excess of that required to raise steam for the gasification area and to compress air for use in the combustor and flue gas oxidizer. The studies made to determine the most efficient and economical way of recovering the heat and energy from this gas are the subject of this section. An alternate heat and power recovery scheme, where dust is removed from the flue gas before expansion, is evaluated in Subsection 4.7.

Alternates Considered

In the TBD, 1500 psig steam is generated in the flue gas oxidizer, and the flue gas is cooled to 1200°F for power recovery in a dusty gas expander. Part of the 1500 psig steam is reduced to 100 psig in a back pressure turbine and used in the process while the excess steam is used to generate export power. Flue gas, in excess of that required to drive the air compressor, is either bypassed to waste heat recovery or the excess air is compressed and vented.

The following options were available and were evaluated to improve process efficiency and economics:

- Add an electrical generator to the compressor-expander-start-up turbine train to generate and export electricity.
- Replace the start-up turbine with an extraction turbine and use it for both start-up and steam let-down. This scheme eliminates the steam let-down turbine at the flue gas oxidizer.
- Use a single electric generator to recover power from the extraction turbine and the flue gas.
- Work with vendors to determine the best expander design, either one two-stage machine or two single-stage machines as shown in the TBD, to process hot dusty gases.

Evaluation

Economic and technical comparisons were made. The economic evaluation took into account credit for extra electric power generation;

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.1 Summary

Evaluation - continued

changes in the kind and number of equipment items; the cost of the new items; and the installed idle equipment required for start-up. Technical considerations were: the state of development of large induction generators; control of synchronous generators; and the evaluation of single-stage and two-stage expanders. The Expander Trade-Off Study is in Subsection 3.4.2, and the Expander Train Generator Trade-Off Study is in Subsection 3.4.3.

Recommendations

1. Eliminate the flue gas oxidizer power generator system which includes the back pressure turbine and the electric generator.

2. Install an electric generator to recover power from the flue gas in excess of that required to drive the air compressors.

3. Combine all of the power generation and steam reduction into single shaft units. The units will include the flue gas expander, the process air compressor, the electric generator and the extraction turbine.

4. Use two-stage expanders instead of single-stage expanders proposed for the TBD.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Summary

The purpose of this study was to determine if a two-stage expander could be justified in place of the two single-stage expanders proposed in the Commercial Plant Tentative Baseline Design. These expanders recover power from the flue gas under conditions similar to dirty gas expanders in Fluid Catalytic Cracker (FCC) power recovery systems. To date only single-stage expanders have been used in FCC service.

The only domestic suppliers of such expanders are Elliott Company, a Division of Carrier Corporation, and Ingersoll-Rand Company. Both companies were solicited for proposals for both single and two-stage expanders. Both vendors elected to propose two-stage designs, providing justification for their selection and verbal quotes for the additional cost for two single-stage expanders in series.

Power Recovery Corporation, which supplies consulting services for power recovery, was engaged to assist in evaluation of vendor proposals and to make the initial trade-off comparison. The conclusions and recommendations described herein represent the common views of the expander vendors, Power Recovery Corporation, COGAS Development Company (CDC), and Dravo.

Two-stage expanders were previously judged to have two drawbacks which had prevented their selection: (1) The vendors did not have designs comparable to single-stage dusty gas expanders, ready for manufacture, and (2) there was concern that the second-stage (in one casing) would experience more rapid blade erosion than a single-stage machine. The position on both of these points has changed. The vendors have designs in hand which they are now proposing to commercial customers and which have a sound basis in commercially proven expanders. Improvements in blade design and understanding of particulate erosion behavior permit the vendors to state that the blade life of a two-stage expander should be equivalent to a single-stage expander. In addition, it has been recognized that the second expander of a two single-stage combination would see dust concentration effects in the interconnecting ductwork which would have an adverse effect on its erosion life.

With these drawbacks satisfactorily resolved, the following advantages of the two-stage expander clearly favor its selection:

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Summary - continued

1. The additional expander on the common shaft, built for opposite rotation, is avoided. Spare parts requirements are reduced.

2. Large interconnecting ductwork (which adds support and alignment problems, increases pressure loss, and interferes with access to the machinery train) is avoided.

3. Expander power recovery efficiency is increased by five percent.

4. A capital investment cost savings of \$26.5 MM and an annual cost savings of \$11 MM, or 15.6¢/MM Btu of pipeline gas, are realized.

Therefore, it is recommend that two-stage expanders be used.

The purpose of this study was to evaluate the desirability of using an electric generator in the expander train to generate electricity from the excess power available under certain process conditions. The concept of flue gas power recovery is equally viable with or without the electric generator in the expander train.

Based on the Tentative Baseline Design (TBD) material and energy balance for the Commercial Plant, it is estimated that all generators in flue gas expander trains of the Commercial Plant would produce an annual mean electrical power output of 24.6[1] Megawatts. This is equivalent to a savings of \$4.48 MM in annual operating costs at an electricity credit of 2.3¢/KWH.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Summary - continued

Further cost benefits are achieved by combining the 100 psig back-pressure steam turbine and synchronous generator with the start-up turbine and the electric generator in the expander machinery train.

Due to these significant cost benefits, it is recommended that a synchronous generator and an extraction steam turbine be used in the expander train replacing the separate 100 psig steam turbine and synchronous generator.

Performance Specifications

The flue gas leaving the Flue Gas Oxidizers (110-44001-1,3) contains a significant amount of thermal and kinetic energy. The Flue Gas Expanders (110-42001-1,6) are designed to recover this energy from the hot, dusty flue gas and the recovered power is to be used to drive the Air Compressors (110-42002-1,6) which supply air to the Combustors (1003-33006-1,3) and the Flue Gas Oxidizers (110-44001-1,3). The following specifications are applicable to the expander trains in the Commercial Plant producing 256 MM SCFD of pipeline gas:

1. The expander (or expanders) is to expand the hot flue gas to a pressure close to ambient pressure.

2. Since the pressure ratio of the expander is approximately 4.0, the expander is to have two stages so as to expand the flue gas efficiently. The two stages of the expander can be in two separate casings, or in one casing. The flue gas contains sufficiently low loadings of particulates to make the use of FCC type expanders practical.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 11 - Expander Design Study

Performance Specifications - continued

3. The expander (or expanders) is to have a projected blade life of at least one year. Blade life of two or three years is desirable and expected in view of five-year blade life capability in current commercial Fluid Catalytic Cracker (FCC) power recovery systems.

4. Each expander machinery train is to contain an expander (or expanders), an air compressor and a start-up steam turbine.

5. Each expander machinery train is to contain no more than four rotating machinery components on a common shaft for reliability and to stay within proven commercial practice.

6. An electric generator may be used to recover power in excess of air compression requirements, if there are less than four other components in the machinery train.

7. There should be no more than six expander machinery trains in the Commercial Plant.

Process Description

This trade-off study compares the following two cases:

Case 1

This Base Case is the one used in the Commercial Plant TBD. Two singel-stage expanders in series are used to recover power from the flue gas. The two single-stage expanders are positioned at the opposite ends of the common shaft in the expander machinery train. This arrangement is necessary to provide for the axial entry of inlet gas to both the expanders. The expander train also contains a double-ended air compressor and a double-ended start-up steam turbine on the same shaft as the expanders. The expander train is designed to run without a gearbox.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Process Description - continued

Case 2

This alternate case uses a two-stage expander to recover power instead of the two-stage expanders of Case 1. The two-stage expander is at one end of the expander train, which also contains a double-ended generator, a double-ended air compressor and a single-ended start-up turbine. The train is designed to run at 3600 rpm without a gearbox. This case includes the generator in the expander train for the following reasons:

(a) The Case 2 expander extracts more horsepower out of the flue gas compared to the Base Case.

(b) With only one expander, Case 2 can add a generator without exceeding four rotating machines on the common shaft.

Technical Comparison

Early experience with commercial flue gas expanders on FCC units had shown that the expander blade life was dictated by the blade erosion characteristics. It was also found that any concentration gradient of dust within the expander affected the blade life adversely.

The two single-stage expanders in series (Case 1) were previously thought to present a more uniform dust distribution to the second stage than possible in a two-stage expander (Case 2). This was attributed to the nosecone and axial entry available to each stage, and the absence of a rotating stage immediately upstream. This presumed uniformity of dust distrubution was the principle reason that two single-stage expanders in series were selected for the Commercial Plant TBD. In additon, expander vendors did not have a two-stage dusty expander design sufficiently developed to be considered seriously.

Because of significant improvements in aerodynamics and blade design, blade erosion no longer is a primary design factor in FCC expanders. Flame-coated blades, have now proven themselves in FCC unit operation with blade lives of three to five years.

The expander vendors are also projecting improved blade erosion sistance for the two-stage design based on their studies of dust particle trajectories and information that indicates particle size reduction in the first-stage.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Technical Comparison - continued

It should also be noted that the early experience on FCC expanders was with Inconel-X blades and AISI-286 disc material. Today's expanders are being offered with Waspalloy discs and/or blades if required, to meet the more stringent stress requirements of temperatures at 1300°F to 1350°F and pressures of 60 psig at the expander inlet. At these inlet conditions, with two-stage design, the blade bending stresses are well within the industry's aerodynamic design experience. Also, the vendors are prepared to offer peripheral steam-cooling at the perimeter to further optimize the blade-to-disc attachment stresses.

It was learned through this study that the two single-stage expanders in series would present a significant potential for concentrating dust because of the changes in the direction of gas flow in the first-stage discharge plenum and in the ductwork between the two expanders. Thus the question of relative erosion resistance and maximum blade life is now at least equal for the two cases, and perhaps even favors Case 2, the two-stage expander design.

Table 3.4.2-1 shows a comparison of the equipment required for each of the two cases. A comparison of the major technical considerations is shown in Table 3.4.2-2. From these tables it can be seen that Case 2 has the following advantages over Case 1:

- Case 2 requires fewer pieces of equipment and spare parts.
- Case 2 does not require the large diameter interconnecting ductwork between the first and second-stage expanders required for Case 1.
- Case 2 eliminates the pressure drop and heat loss associated with the interconnecting ductwork between the two stages required for Case 1. This results in an increase 5 percent in the power recovery for Case 2 as compared with Case 1.
- Case 2 includes generators which improve the plant thermal efficiency by about 1 percent. Generators were not included in Case 1 because of the question of the reliability of

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Technical Comparison - continued

a 5-component machinery train. If the generators were included in Case 1, the electrical output would be 13,200 KW on an annual average basis, as indicated in Table 3.4.2-3, Case 1A.

Economic Comparison

The economic comparison of the two cases is based on mid-1977 equipment costs submitted by the vendors. The equipment costs for the two single-stage expanders in series were obtained as verbal quotes from the vendors for the incremental costs of the two single-stage expanders compared to the two-stage expander. Other costs for the switchgear, ductwork, spare parts and miscellaneous items were estimated by Power Recovery Corporation. The electricity credit rate of 2.3¢ KWH was selected by Dravo/ICGG.

The detailed cost comparison of the two expander alternatives is shown in Table 3.4.2-3. This table also includes a cost comparison of a subcase of Case 1, namely: Case 1A with the generators. The sub-case is provided to show that the addition of generators to Case 1 is not attractive from a cost viewpoint. From this table, it can be seen that Case 2 has the following cost advantages over Case 1, the Base Case used in the Commercial TBD:

- Case 2 yields a reduction of $26.5 MM in the total fixed capital cost of the Commercial Plant.
- Case 2 yields a reduction of 15.6¢/MM Btu in pipeline gas cost at an electricity credit rate of 2.3¢/KWH.

Recommendation

In view of the significant technical and economic advantages, it is recommended that a two-stage expander be used to recover power in the Commercial Plant Revised Tentative Baseline Design and in subsequent plant designs.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Table 3.4.2-1

Equipment Comparison for the Two Expander Alternatives*

Item	Case 1	Case 2
Number of Expanders	Two	One
Number of Compressors	One	One
Number of Steam Turbines	One	One
Number of Generators	None	One
Gimbel & Expansion Joints	2 Gimbel Joints and 4 Expansion Joints	1 Gimbel Joint and 3 Expansion Joints
Control Loops	2 Complete sets for Expanders and 1 Complete set for each Compressor and Turbine	1 Complete set for Expanders, 1 Complete set for Compressor and Turbine, and 1 Complete set for Induction Generator
Lube Oil Consoles	One	One
Sets of Spare Rotors	2 Expanders 1 Compressor 1 Turbine	1 Expander 1 Compressor
Sets of Spare Bearings and Wear Parts	2 Expanders 1 Compressor 1 Turbine	1 Expander 1 Compressor 1 Generator 1 Turbine

*For one expander train only. The Commercial Plant contains a total of six expander trains.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Table 3.4.2-2

A Comparison of Major Technical Considerations for the Two Expander Alternatives

Item	Case 1	Case 2
1. Mechanical Reliability	Less than Case 2 because of increased number of gimbel and expansion joints.	Good
2. Train Arrangement Feasibility	Unproven Arrangement	Proven Arrangement
3. Plot Space Requirement	35 feet x 190 feet	35 feet x 145 feet
4. Maintenance	More than Case 2	Normal
5. Operability	Less than Case 2 because of unproven arrangement and less mechanical reliability.	Good
6. Blade Erosion	Comparable to proven experience in first expander. Not proven for second expander but dust concentration expected in interconnecting ductwork.	Comparable to proven experience in the first-stage. Not proven in the second stage but expected to be equal or less severe than first stage.
7. Equipment Modifications	(a) Increased pressure rating of the inlet casing of first expander. (b) Discharge casing of first expander needs increasing pressure rating. (c) Reverse rotation design required for second expander.	(a) Increased pressure rating of inlet casing of first stage. (b) Expander designed for two stage in one casing.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Table 3.4.2-2 - continued

A Comparison of Major Technical Considerations for the Two Expander Alternatives

	Item	Case 1	Case 2
8.	Energy Recovery	Base	5% more energy recovered with Case 1; Results in a increase of about one percent in Plant Thermal efficiency.
9.	Effect on Gas Cost	Base	Results in about ¢16/MM Btu decrease in SPG cost as compared with Case 1

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.2 Area 110 - Expander Design Study

Table 3.4.2-3

Cost Comparison of the Two Expander Alternates
(Cost in $ MM)

		Case 1 Two Single-Stage Expanders in Series $ MM: 1 (without Gen.)	Case 1 Two Single-Stage Expanders in Series $ MM: 1A (with Gen.)		Case 2 Two-Stage Expander $ MM
Expander[1]	(6-1st Stage)	$ 6.6	$ 6.6	(6-Two-Stage)	$9.0
	(6-2nd Stage)	10.8	10.8		
	Subtotal	$17.4	$17.4		$9.0
Ductwork, gimbel & expansion joints between stages 1 & 2		$ 5.7	$ 5.7		$ -
Compressor[1]	(6 Required)	12.6	12.6	(6 Required)	12.6
Steam Turbine[1]	(6 Required)	8.4	8.4	(6 Required)	6.0
Generator[1]	(None)	-	3.0	(6 Required)	3.0
Switchgear for the Generator[1]		-	3.5	(6 Required)	3.5
Additional spare parts cost for rotors, blades, stators, & all auxiliary equipment		9.6	9.6		7.8
Total Equipment Cost:		$ 53.7	$ 60.2		$41.9
Total Fixed Capital, (2.25 x Total Equipment Cost)		$120.8	$135.5		$94.3
Incremental Annual Operation Cost[5] $MM/yr:					
Fixed Capital Related[3]		0.0[2]	-4.41		-7.95
Electricity Credit[4] @ 2.3 ¢/KWH		0.0[2]	-2.40		-4.48
Total		0.0[2]	+2.01		-12.43
Change in Gas Cost ¢/MM Btu[4]		0.0[2]	+2.5		-15.6

BASIS:

1. There are six machinery trains in the Commercial Plant.
2. Base Case.
3. Using fixed capital related factor of 30 percent.
4. Using 330 operating days per year, and (242×10^9) Btu/d for total pipeline gas output.
5. Mid-1977 costs.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train Generator Design Study

Summary

The purpose of this study was to evaluate the desirability of using an electric generator in the expander train to generate electricity from the excess power available under certain process conditions. The concept of flue gas power recovery is equally viable with or without the electric generator in the expander train.

Based on the Tentative Baseline Design (TBD) material and energy balance for the Commercial Plant, it is estimated that all generators in flue gas expander trains of the Commercial Plant would improve the plant thermal efficiency by about one percent. This is equivalent to a savings of about 6¢/MM Btu in SPG cost.

Further cost benefits are achieved by combining the 100 psig back-pressure steam turbine and synchronous generator with the start-up turbine and the electric generator in the expander machinery train.

Due to these significant cost benefits, it is recommended that a generator and an extraction steam turbine be used in the expander train replacing the separate 100 psig steam turbine and synchronous generator.

Introduction

The Commercial Plant Tentative Baseline Design (TBD) material and energy balances showed that the power available from expanding the flue gas was in excess of the air compression requirements. Excess power in the expander train is normally handled by choosing one of the following alternatives:

- Bypass part of the flue gas around the expander through an orifice chamber for pressure let-down.
- Use a generator in the expander train to convert the excess power into electricity.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train Generator Design Study

Introduction - continued

The choice between the two systems depends on individual project requirements, total plant energy balances and economic factors. Power Recovery Corporation (PRC), power recovery consultants to CDC, were retained to carry out a study of these alternatives and to recommend the better alternative. PRC completed this study and recommended the use of an induction generator in the train. During the course of the PRC study, it became evident that further improvements in the overall power recovery design were possible. These improvements could be achieved by combining the functions of the induction generator and start-up turbine in the expander train and the 100 psig steam backpressure turbine (which operates on 1500 psig steam from the Flue Gas Oxidizer) and its associated synchronous generator. The net effect of this combination would be to reduce the number of steam turbine-generator sets for the Commercial Plant from 9 to 6 and to increase the ratings on the new generators. Due to the unavailability of induction generators in the size range required for the combined load, synchronous generators would have to be used. Synchronous generators, though used commonly in electrical generation systems, have not been used to date in direct connection with flue gas power recovery expanders. The use of synchronous generators may add complexity to the power recovery control system because the synchronous generators need to be synchronized with the power frequency of the Commercial Plant, as opposed to induction generators which derive their field excitation from the plant power and are therefore synchronized with the purchased plant power frequency. The synchronization presents an added control consideration in the power recovery control system design similar to that already required for the turbine/generator sets used in the TBD.

The current RTBD material and energy balances indicate that the required generator size will decrease by 25% for the RTBD. Due to the smaller size of RTBD generator and the preliminary vendor indications that they could increase the induction generator rating from the present maximum value to that for RTBD, it will be possible to use an induction generator in the expander train.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train Generator Design Study

Process Description

This design study compared the following two cases:

Case 1 (TBD Case): This case does not use an electric generator in the expander train. It uses a condensing turbine in the train for start up and as a spare prime mover. It also has a separate 100 psig backpressure steam turbine and its associated synchronous generator. This 100 psig backpressure steam turbine produces all the steam required in Area 103 by direct let down from 1500 psig steam.

Case 2: The case includes a synchronous generator and an extraction steam turbine. The extraction steam turbine produces all the 100 psig steam required in Area 103. The condensing part of the extraction steam turbine is used during start-up and when additional power is required in the train. The synchronous generator absorbs all the power from this steam turbine, and also that produced by the expander in excess of the compressor needs, and converts it to electricity.

Equipment Comparison[2]

There are several major pieces of equipment affected in the comparison between Case 1 and Case 2. Table 3.4.3-1 gives this comparison and equipment capacities for the Commercial Plant producing 256 MM SCFD pipeline gas.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train Generator Design Study

Table 3.4.3-1

Major Equipment Ratings

Equipment	Case 1 (TBD) Quantity Required	Case 1 (TBD) Capacity Rating	Case 2 Quantity Required	Case 2 Capacity Rating
Start-up Turbine	6	Base	0	Not Required
100 psig Backpressure Steam Turbine	3	Base	0	Not Required
Extraction Turbine	0	Not Required	6	Same as start-up turbine for Case 1
Synchronous Generator	3	Base	6	Each unit 25% smaller than single Case 1 unit
Surface Condenser	6	Base	6	35% smaller than Case 1

The start-up turbines used in Case 1 (See Figure 3.4.3-1) and the condensing section of the Case 2 (See Figure 3.4.3-2) extraction turbine are not used during normal steady-state plant operation, and add idling capacity to the following utility systems:

- 1500 psig, 850°F steam system
- Cooling water system
- Boiler Feed Water (BFW) and condensate return system

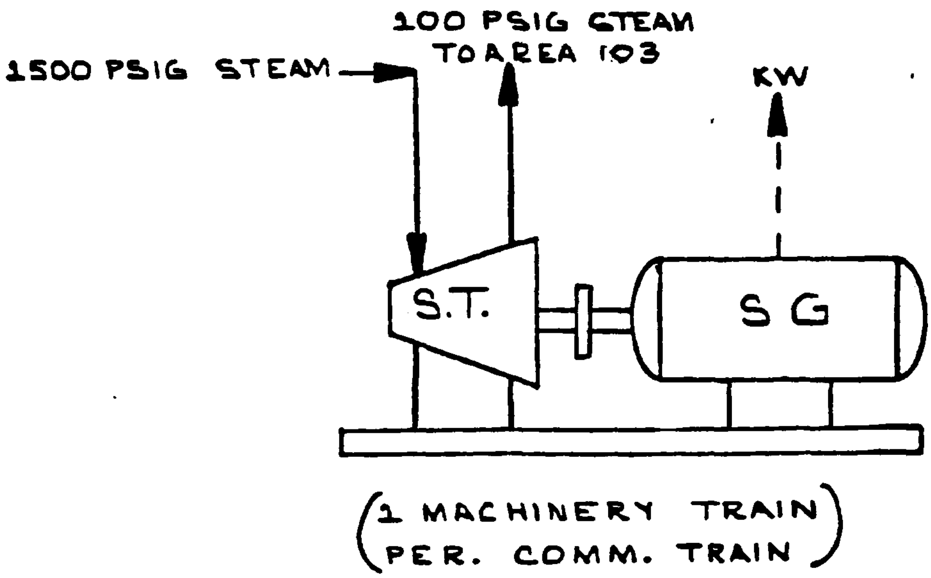
These idle utility capacities related to the steam turbine usage are estimated for the cases and are included in Table 3.4.3-2.

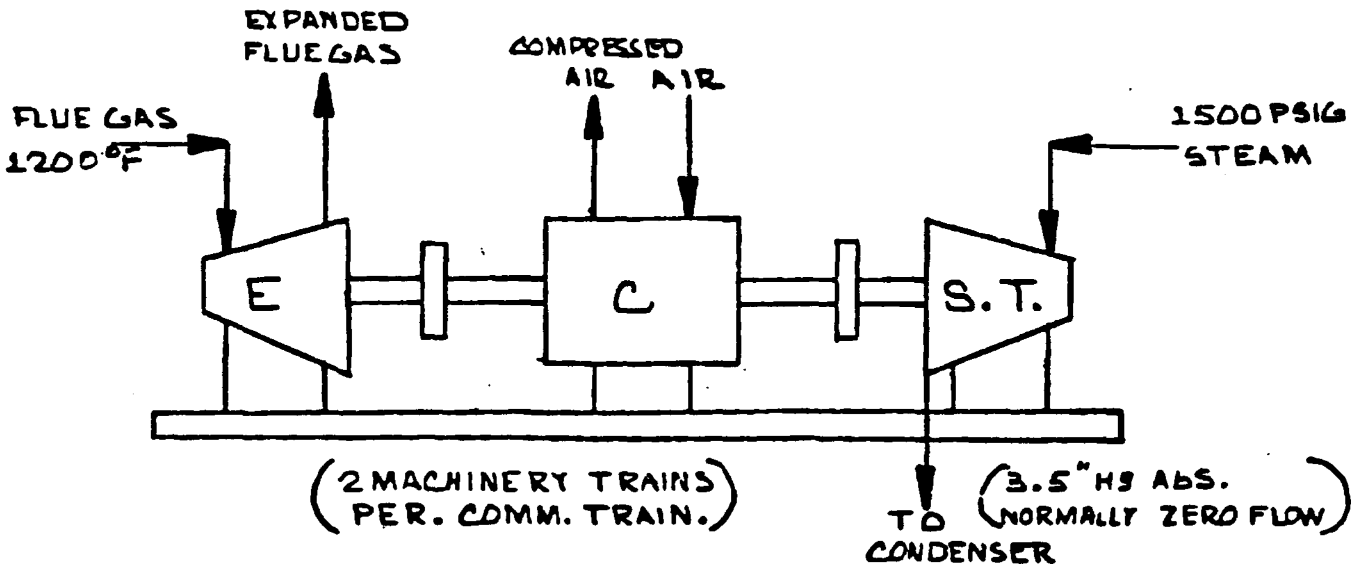
3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train - Generator Design Study

Figure 3.4.3-1

Schematic Diagram For An Expander Train Without A Generator (Case 1, TBD)





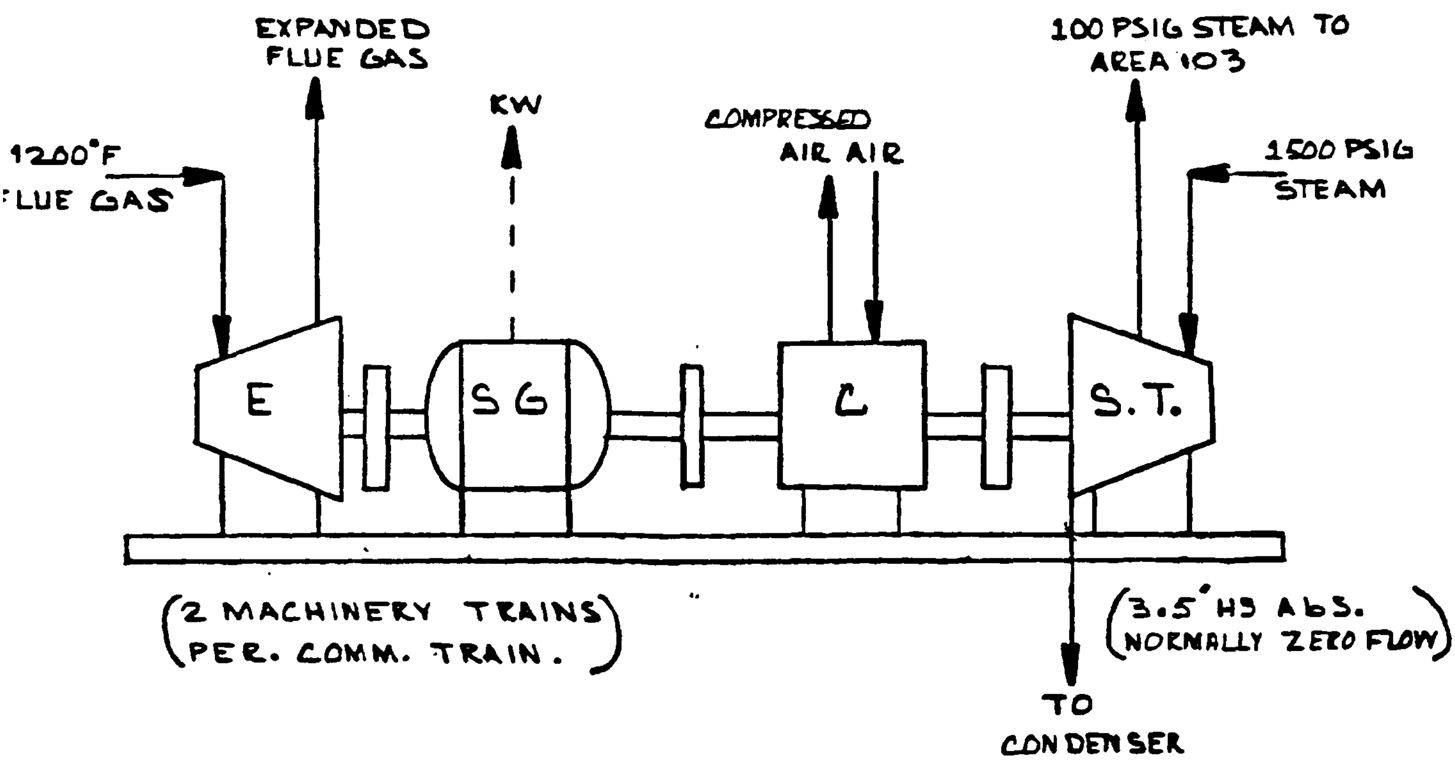
3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train-Generator Design Study

Figure 3.4.3-2

Schematic Diagram For An Expander Train With Synchronous Generator (Case 2)

SCHEMATIC DIAGRAM FOR AN EXPANDER TRAIN WITH SYNCHRONOUS GENERATOR. (CASE 2)



3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train Generator Design Study

Table 3.4.3-2

Utility Capacities Required For Area 110 Start-Up Turbines

Equipment	Case 1 (TBD)	Case 2
Idle Steam Turbine Capacity	Base	50% less than Case 1.
Related Utilities:	Total for Three Process Trains	
BFW Pumps	Base	35% less than Case 1.
Cooling Water Pumps	Base	40% less than Case 1.
Cooling Tower Fans	Base	40% less than Case 1.
Start-Up Steam Boiler Capacity	Base	35% less than Case 1.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train Generator Design Study

Economic Comparison

When Case 2 is compared with Case 1, the TBD case, the following advantages are evident:

- Case 2 eliminates three large steam turbines, each costing about $1MM in installed equipment costs.
- Case 2 reduces the idle surface exchanger capacity requirements by 35 percent.
- Case 2 will reduce the idle utility capacities by 50 percent.
- Case 2 will improve the plant thermal efficiency by about 1 percent as compared with Case 1. This will result in about 6¢/MM Btu reduction in SPG cost.

Case 2 shows the following disadvantages as compared with Case 1.

- Case 2 requires six smaller synchronous generators as compared with three large synchronous generators for Case 1. Thus the equipment cost reduction achieved by reducing the number of steam turbines will be partially offset by the cost increase due to these generators.

Technical Comparison

The introduction of a synchronous generator adds a slight complexity to the expander train and its control system. The combination of the 100 psig backpressure turbine with the steam turbine in the expander train may at first appear to limit the flexibility of the 100 psig steam system, the Flue Gas Oxidizer, and the flue gas expander train. However, a closer look at each of these systems indicated that the flexibility is not significantly affected since the normal design features accommodate either turbine/generator system.

3.4 AREA 110 - FLUE GAS POWER RECOVERY

3.4.3 Expander Train Generator Design Study

Recommendations

1. It is recommended that Case 2, with the synchronous generator and an extraction turbine, be used in the expander train for future Commercial Plant designs due to the expected savings in capital costs and significant savings in operating costs.

2. It is also recommended that an economic study be completed to quantify the savings in annual operating cost.

3. It is recommended that the substitution of the synchronous generator with the proven induction generator be investigated further to establish whether an induction generator of sufficient capacity can be used.

3.5 AREA 114 - THERMAL OXIDIZER AND FLARE

The purpose of this section is to describe the functions of the Thermal Oxidizer and the Flare. Both items are industrial standards for the services they provide and neither has been altered from the design shown in the Tentative Baseline Design (TBD) Report.

Thermal Oxidizer

The Thermal Oxidizer provides for the disposal of plant wastes other than the raw oil-containing streams produced in the plant.

The expected wastes are water-quenched oils from Area 104 which normally are processed continuously unless problems aris in downstream equipment. If problems occur, the oils are stored in the Raw Oil Storage Tank until they can be recycled to the process. Oils from contaminated wastewater streams, such as those found in process sewers, are recovered in the Oil-Water Separator and recycled to the process. Residual oil in the water phase is digested in the biotreatment plant. These are the main sources of contaminated oil, neither of which is disposed of in the Thermal Oxidizer.

Oil spills, in which the oil becomes contaminated with dirt or other foreign substances, are disposed of in the Thermal Oxidizer. Off-spec process streams, which are determined by operating personnel to be of insufficient quality to warrant their recycle into the process, will also be combusted in the Thermal Oxidizer.

One thermal oxidizer is provided for each train of the Commercial Plant.

Flare

The Flare combusts atmospheric contaminants in process streams and vents. These streams are produced during start-up and shutdown, or as a result of process upsets (i.e., off-spec SPG) and plant emergencies. The Flare has sufficient capacity to oxidize any process stream produced in a single train of the Commercial Plant.

3.6 AREA 116 - WATER SUPPLY

3.6.1 Summary

Approximately 12,000 GPM of raw water will be withdrawn from the Mississippi River to satisfy the requirements of the plant. The water supply system consists of an intake structure and an overland pipeline from the Mississippi River to the plant site. In this section, two types of intake systems currently being considered are discussed. However, until more specific site information is available and discussions with controlling agencies on the various alternatives are held, a specific recommendation cannot be made. A plan for obtaining the required information from informed sources to permit the recommendation of a specific water intake method is briefly outlined.

Finally, the basis for the selection of the intake structure for the baseline concept is presented.

3.6.2 Performance Requirements

The water supply system must obtain and transport approximately 12,000 GPM of raw water to the plant from the Mississippi River near Chester, Illinois. The system selected shall have the capability of maintaining the supply of water to the plant throughout the year, in the required quantities, and as free from suspended matter as is economically feasible.

3.6.3 Design Data - Mississippi River

The following data for the Mississippi River were received from the Corps of Engineers, St. Louis, MO:

Elevations at the intake structure:

Low water approximately	El.	340.0'
Normal water approximately	El.	356.0'
High water approximately	El.	387.0'

Annual flows from the Mississippi River at Chester, Illinois:

Maximum	873,000 cfs
Minimum	40,000 cfs
Mean	200,000 cfs

3.6 AREA 116 - WATER SUPPLY

3.6.3 Design Data - Mississippi River - continued

The plant usage of approximately 12,000 gpm represents less than 0.1% of the minimum river water flow. The analysis of the raw water is provided in Subsection 5.3 - Area 117, Water Treatment Systems.

3.6.4 River Water Intake Systems

The withdrawal of water from the river can be accomplished in at least two ways: by well or by intake structure. The decision as to which of the two is the more favorable for this application must be determined from a study of the nature of the source, its geological environment, and the physical character of the area in which it is to be utilized.

The site currently proposed for the water intake has a relatively large bottom area as opposed to the original site. Whether this is suitable for the installation of wells is unknown, however. In view of the availability of this bottom area, the use of wells cannot be eliminated until more information becomes available.

For the baseline concept the location proposed for the water intake system was north of the site currently being considered. At this location the shore of the river is a steep bluff with virtually no bottom area. Since the suitability of a well installation is dependent on the availability of sufficient bottom area to permit the drilling of the wells into the sand and gravel stratum, the terrain precluded the use of this method and dictated the incorporation of an intake structure into the baseline design.

The intake structure would be located on the shore with the low water elevation setting the elevation of the bottom of the structure and the height being controlled by the high flood water elevation. There are four methods which can be considered for transferring the water from the river into the pump pit. These are as follows:

1. Direct flow from river through conventional vertical traveling screens and trash racks.

2. Infiltration beds.

3. Perforated pipes placed in the river channel above the stream bed.

4. Perforated pipes located in an off-stream channel above the channel bed.

3.6 AREA 116 - WATER SUPPLY

3.6.4 River Water Intake Systems - continued

Details of these four methods can be found in "Intake System Assessment For Central Columbia River," as published in the ASCE Journal of the Power Division, December, 1974. Brief descriptions of each follow.

Base Case - Screens and Racks

The Base Case system consists essentially of sloping trash racks, fish escape slots, traveling screens and pumps. The sloping trash racks are mechanically cleaned to remove the debris. The traveling screens are usually cleaned using high pressure sprays near the top, whereby debris is dislodged into a trough from where it can be collected for disposal. Adequate cross sectional area for the flow of water into the screens would be provided to limit the velocity to 0.5 ft/s.

A detailed process description can be found in Section 5.0, Baseline Process Concept, sub-section 5.3, Process Description. Process Flow Diagram No. 116-100-1, in Subsection 5.4, depicts the Baseline Concept.

Alternate 1 - Infiltration Beds

System No. 2 is used commercially for water purification by municipalities. The filter consists of a layer of sand or crushed coal supported on a bed of gravel. Perforated pipes are used as filter underdrains. Suspended particles and bacteria adhere to the sand grains when water is passed through the filter. Water passages are thus reduced, and a straining action results. As more material is trapped in the sand bed, the pores become clogged and it is necessary to backwash the filter. To reduce the suspended solids reaching the filter bed, a diversion channel is required. This acts as a stilling basin and is divided to allow for switching and removal of silt accumulations periodically.

Alternate 2 - Perforated Pipes in the River Channel

The third system consists of perforated pipes in the river channel. The number and size of the perforations used in the pipe determine the velocity through the perforations and the velocity

3.6 AREA 116 - WATER SUPPLY

3.6.4 River Water Intake Systems

Alternate 2 - Perforated Pipes in the River Channel - continued

approaching the perforations. Low approach velocities can be attained without significantly increasing the complexity and cost of the system. In general, the longer the pipe is, the more non-uniform the velocity distribution. Similarly, the velocity distribution becomes more non-uniform as the diameter of the pipe decreases. The pipes must be located far enough from the shore to be at least 2 feet off the bottom and at least 2 feet below the low water elevation. The pipes would be located outside the river channel and thus would not constitute a navigational problem to commercial navigation.

During periods of high pollutant content in the river, a water backwash in conjunction with the river current should keep the perforations open. If biological growth should tend to close the perforations, divers could scrub the pipes in place or remove entire sections for cleaning. The reliability of the perforated pipe intake is very high, based on reported experience. Since there are no moving parts, operation and maintenance would be fairly simple.

The pressure-head differential between the water level in the pump pit and the river must be monitored to insure that clogging has not occurred. Problems of winter icing should be minimal with this system.

Alternate 3 - Perforated Pipes in Off-Stream Channel

The fourth system is similar to the third system except the perforated pipes are located in an off-stream channel. A canal and stilling basin would have to be dredged so that water could be diverted from the river to the pipe intakes.

3.6 AREA 116 - WATER SUPPLY

3.6.4 River Water Intake Systems

Conclusions

Because of the lack of specific site information a decision concerning the type of water intake system to be employed cannot be made. In the near future, it is planned to consult with knowledgeable individuals employed by companies and municipalities presently utilizing the Mississippi River in the vicinity of Chester, Illinois as a source of water supply. Based upon the information acquired during these consultations, a recommendation as to the type of system to be installed will be made.

3.6.5 Screening

There are periods during the year when the suspended solids in the Mississippi River are relatively high. An evaluation of this condition is planned to determine the desirability of providing in-line screening ahead of the booster pumps. Such filtration may significantly reduce both pump and transmission line erosion as well as reduce potential operational problems at the softener-clarifier associated with significant fluctuations of the solids level in the water to be treated.

The stream containing the removed solids will require thickening and/or dewatering. The dewatered solids can be disposed of with the screenings.

3.7 AREA 117 - WATER TREATMENT

3.7.1 Summary

Of the three design variations studied for treatment of the raw and recycled water necessary to satisfy the water requirements of the gasification facilities, Alternate 2, (zeolite softeners and moving bed exchangers) appears to be the most desirable. The slightly higher operating cost, over that of the Base Case, is more than offset by the lower capital cost, the increased operational flexibility, and the greater assurance of uniformly treated water quality.

Table 3.7.1-1 provides a comparative summary of capital requirements, total annualized cost, technical evaluation, and the total evaluation.

Table 3.7.1-1
Evaluation Summary
(In Year End 1977 Dollars)

Process Alternative	Capital Requirement (MM $)	Total Annualized Cost (MM $/yr)	Technical Evaluation	Total Evaluation
Base Case	0	0	0	0
Gaco-Sternson	+5.99	+1.08	+1	-2
Best Tech., Inc.	-0.47	-0.65	0	+7

The technical evaluation column above is derived from the Technical Evaluation Matrix, Table 3.7.10-3. It indicates the deviation from the Base Case when each alternate was evaluated on a basis of 40 points.

The total evaluation column above is derived from the Process Selection Matrix, Table 3.7.10-4. This column reflects the difference between the Base Case and the alternates based on the total numerical evaluation achieved by each alternate on an overall scale of 100 points.

3.7 AREA 117 - WATER TREATMENT

3.7.1 Summary - continued

It should be recognized that this evaluation and these costs are based upon the waste treatment process and resulting effluent characteristics of the Base Case waste treatment only. Alternate waste treatment processes and alternative processes selected in other plant areas may affect, to some degree, the water treatment process selected as well as the capital requirements and operating costs of water treatment. The technical capability, reliability and risk factors, however, will remain unchanged.

3.7.2 Raw Water Character

The raw water source is the Mississippi River near Chester. Illinois. The analysis of this water appears in the table entitled "Raw Water Analysis," in Section 5.0, Baseline Process Concept, Subsection 5.3, Process Description, this report.

3.7.3 Water Use Treatment Requirements

The water use treatment requirements are shown in the table titled "Water Treatment Requirement for Each Water Use Category," in Section 5.0, Subsection 5.3, Process Description, this report.

3.7.4 Water Use Tolerance Levels

The ion and/or ion combination limitations determining the degree of water treatment necessary are as shown on Table 3.7.4-1.

3.7.5 Performance Specifications

Each of the unit processes must perform its water treatment function with consistent reliability in such a manner as to provide the required quantity of treated water at the required purity, with a minimum production of wastes which can be treated in the most cost effective manner. Each of the vendors contacted was made aware of the objectives and the necessity of balancing the economics of water treatment against the costs of waste treatment, so generated in order to achieve the goal of zero liquid waste discharge.

Table 3.7.4-1

Water Use Tolerance Levels

	Raw Water	Cooling Tower Make-up	Cooling Tower Tolerance	Low Pres. Boiler Make-up	High Pres. Boiler Make-up
Total Hard. (as $CaCO_3$) mg/l (4)	254			0.2	0.0
Calcium (as $CaCO_3$) mg/l (4)	172	(1)	(1)	---	---
Magnesium (as $CaCO_3$) mg/l (4)	82	(2)	(2)	---	---
P. Alk.	0			(Total Alk.)(5)	0
M. Alk.	154		150 max.	(-400)	
SiO_2	15	(2)	(2)	30	1
SO_4	101			---	---
Cl	29.4		3,000	---	---
Na	32.5			---	---
TDS	379		5-6,000	4000(6)	100 (6)
Sus. Sdids mg/l	200(3)		300		
Ca x CO_3 (as $CaCO_3$)			6,000		
Carbonates, mg/l			5		
Bicarbonates, mg/l			300		
(2) Mg x SiO_2			36,500		
(1) Ca x SO_4			1.5×10^6		
PO_4			20		
NH_3			20		
pH			7.0		

(1) and (2) Cooling tower make-up value determines the cooling tower circulation concentration according to: Mgx cycles x SiO_2 x cycles. This value should not exceed 36,500 ($CaSO_4$ by similar calculation). Note: It is possible that at excess of 150 mg/l the silica may ppt. as SiO_2 even though the Mg SiO_2 limit has not been reached.

(3) Max. 4-500

(4) Potable and process water - 100

(5) Limit to 10% of specific

(6) Specific conductance

3.7 AREA 117 - WATER TREATMENT

3.7.6 Processes Considered

In general, there are three classes of water quality required for use by the various areas of the coal gasification process. Potable and process water require the least treatment; cooling tower make-up, in order to minimize blowdown, and low pressure boiler make-up require a relatively high degree of treatment; and the make-up water for the high pressure boilers requires complete demineralization and silica removal.

The water treatment systems considered to achieve these treatment objectives were composed of various unit processes as follows:

Lime-soda softening of cooling tower make-up with deionization and demineralization of boiler make-up.

Excess lime-soda softening with weak acid cation exchange for cooling tower and low pressure boiler make-up, and demineralization for high pressure boiler make-up.

Clarification, filtration, and zeolite softening for cooling tower and low pressure boiler water make-up, and demineralization for high pressure boiler water make-up.

The Base Case provides lime-soda softened water for cooling tower make up and potable and process water uses, and pretreated water for low pressure boiler make-up and for the high pressure boiler water demineralizer. Carbon filtration is provided ahead of the demineralizer to prevent fouling and/or deterioration of the resin beds by low level suspended solids, soluble organics and any possible chlorine residual resulting from chlorinization of raw water at the river water pumping station.

An alternate system, employs excess lime-soda softening, recarbonation, clarification, and filtration as the treatment provided for potable and process water. The softening also provides pretreatment for a weak acid cation exchanger, producing an effluent to be used for low pressure boiler water make up and blending on a 50-50 basis with softened water for cooling tower make-up. The weak acid cation exchanger also provides pretreatment for the demineralization of high pressure boiler water make-up.

3.7 Area 117 - WATER TREATMENT

3.7.6 Processes Considered - continued

A third system is one in which cooling tower make-up, low pressure boiler make-up, process water, and potable water are treated with zeolite softeners. The necessary freshwater for high pressure boiler make-up is treated by moving bed deionizers, combined with the condensate return and evaporator distillate, and then is passed through carbon filters and mixed bed condensate polishers to provide the total high pressure boiler make-up requirements.

3.7.7 Base Case

The Base Case process description is in Section 5.0, Baseline Process Concept, Subsection 5.3.

The Base Case Process Flow Diagrams are shown in Section 5.0, Baseline Process Concept, Subsection 5.4.

Utility requirements for the Base Case are shown in Section 5.0, Baseline Process Concept, Subsection 5.5.

The energy balance for the Base Case appears in Section 5.0, Baseline Process Concept, Subsection 5.6.

3.7.8 Alternate 1

Introduction

This water treatment system which consists of a number of unit processes has as its primary objective water treatment unit processes based upon the minimizing of the volume of high dissolved solids wastes which must be evaporated. To achieve this objective the amount of softening provided for cooling tower make-up was improved to permit 15 cycles through the tower and a blowdown of 300 GPM.

The process employs conventional excess lime-soda softening followed by recarbonation. This softened water is satisfactory for potable and process water use. The low pressure boiler make-up water requirement, 50% of the cooling tower fresh water make-up, and 50% of the demineralizer feed requirement are further softened by weak acid cation exchange. The demineralizer feed is pretreated by carbon filtration to remove any possible chlorine carry-through from the chlorine added at the river water pumping station. The demineralized water satisfies the fresh water high pressure boiler make-up requirements.

3.7 AREA 117 - WATER TREATMENT

3.7.8. Alternate 1

The balance of the high pressure boiler make-up is provided by carbon filtered distillate, resulting from evaporation of the high dissolved solids waste streams (cooling tower blowdown, and spent exchanger and demineralizer regenerant), and by condensate return, which is provided with carbon filtration and condensate polishing before reuse in the high pressure boiler system.

The process, and the relationship of treatment units, is shown in Block Flow Diagram No. 117-120-1A1.

Process Description

Lime-Soda Softening

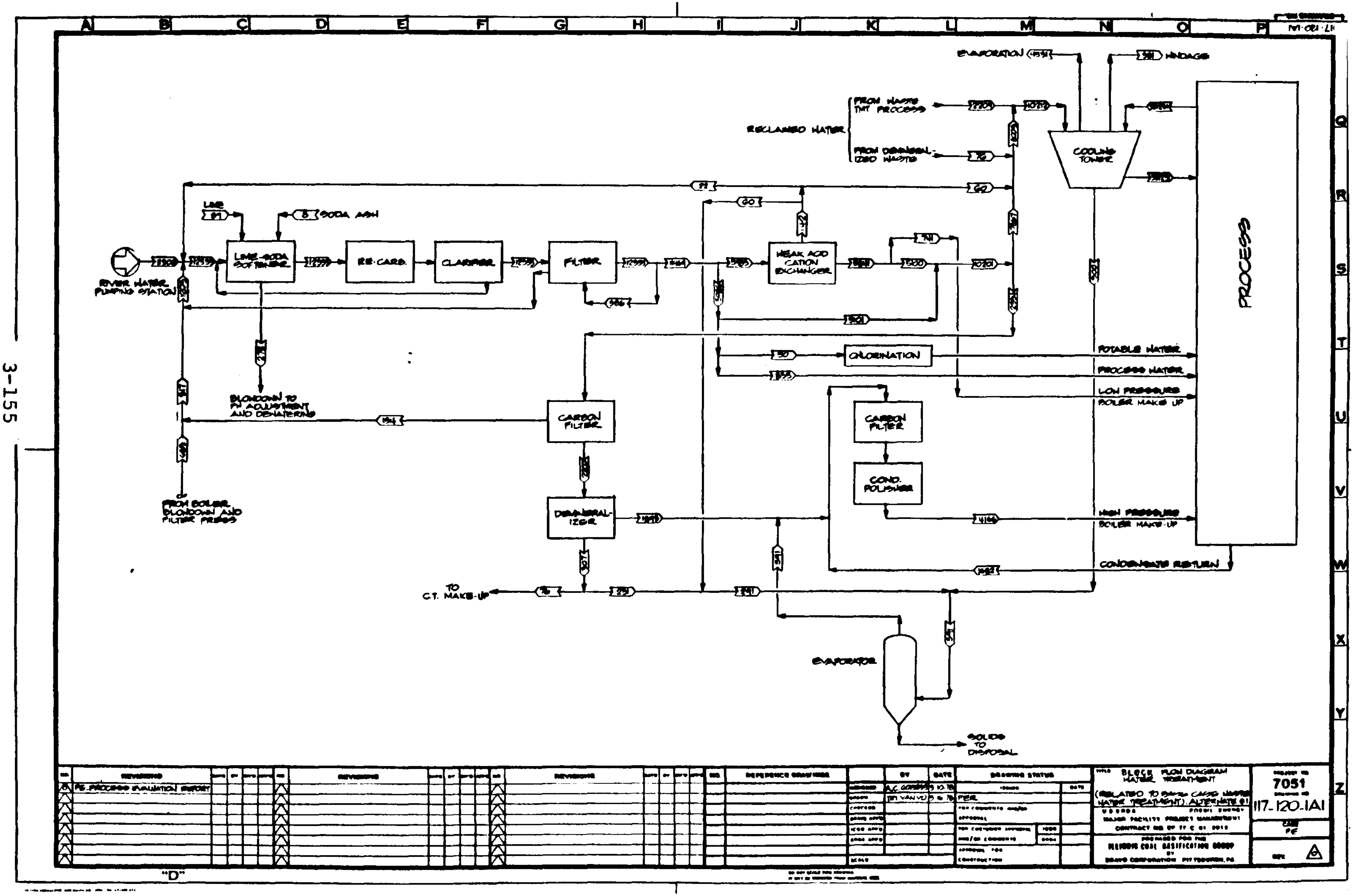
The lime-soda softening process is similar to that described for the Base Case in Section 5.3. Softening takes place in accordance with the following chemical reactions:

$$CO_2 + Ca(OH)_2 = CaCO_3\downarrow + H_2O$$
$$Ca(HCO_3)_2 + Ca(OH)_2 = 2CaCO_3\downarrow + 2H_2O$$
$$Mg(HCO_3)_2 + Ca(OH)_2 = CaCO_3\downarrow + MgCO_3 + 2H_2O$$
$$MgCO_3 + Ca(OH)_2 = CaCO_3\downarrow + Mg(OH)_2$$
$$2NaHCO_3 + Ca(OH)_2 = CaCO_3\downarrow + Na_2CO_3 + 2H_2O$$
$$MgSO_4 + Ca(OH)_2 = Mg(OH)_2\downarrow + CaSO_4$$
$$CaSO_4 + Na_2CO_3 = CaCO_3\downarrow + Na_2SO_4$$

However, with the theoretical amount of lime required, the magnesium removal is limited by the resulting hydroxide alkalinity. To maximize magnesium removal the alkalinity must be increased. This is accomplished by adding an excess of lime. The resulting excess causticity is then precipitated with carbon dioxide according to the reaction: $Ca(OH)_2 + CO_2 \rightarrow CaCO_3\downarrow + H_2O$. To prevent overloading the filter, the precipitate is then settled out in the secondary clarifier.

Filtration

To assure removal of all suspended solids, some of which are extremely slow settling hydroxides, the secondary clarifier effluent is filtered.



EVAPORATION
WINDAGE
FROM WASTE TMT PROCESS
RECLAIMED WATER
FROM DEMINERAL-IZER WASTE
COOLING TOWER
PROCESS
LIME
SODA ASH
RIVER WATER PUMPING STATION
LIME-SODA SOFTENER
RE-CARB.
CLARIFIER
FILTER
WEAK ACID CATION EXCHANGER
BLOWDOWN TO PH ADJUSTMENT AND DEWATERING
CHLORINATION
POTABLE WATER
PROCESS WATER
LOW PRESSURE BOILER MAKE UP
CARBON FILTER
CARBON FILTER
COND. POLISHER
HIGH PRESSURE BOILER MAKE-UP
FROM BOILER BLOWDOWN AND FILTER PRESS
DEMINERAL-IZER
CONDENSATE RETURN
TO C.T. MAKE-UP
EVAPORATOR
SOLIDS TO DISPOSAL
BLOCK FLOW DIAGRAM WATER TREATMENT
7051
117-120-1A1
"D"

3.7 AREA 117 - WATER TREATMENT

3.7.8 Alternate 1

Process Description

pH Adjustment

The excess lime-soda treatment results in a high alkalinity water which may be corrosive in nature. It is necessary, therefore, to adjust the pH to a pH-alkalinity relationship which results in a stable water. This adjustment is accomplished by the addition of more CO_2 in sufficient quantity to neutralize the excess alkalinity after the magnesium and calcium precipitations have been completed and the resulting solids settled out. However, even with this treatment accomplished at a reasonably economic lime addition, complete precipitation of the magnesium and calcium cannot be accomplished. A silica reductión from 15 to 10 mg/l will occur as a result of adsorption on the magnesium hydroxide.

Weak Acid Cationic Exchange

With the above degree of treatment, the cooling tower could operate at 12 cycles, limited by $MgSiO_2$, and resulting in a blowdown of 485 GPM. Process water requirements and, with chlorination, potable water demand can be satisfied with this water. Additional softening, to safisfy the requirement for low pressure boiler water make-up and cooling tower make-up, compatible with a 15 cycle operation, is accomplished by weak acid cation exchange which reduces the calcium and magnesium to zero.

This reduction of calcium and magnesium in the cooling tower make-up will permit additional cycles, reduce the blowdown, and require less evaporator capacity. With calcium and magnesium no longer a limiting factor, the cycles through the cooling tower are limited by SiO_2 to 15. Split treatment, with 50% of the filtered, pH adjusted effluent being treated by the weak acid cation exchange units, provides sufficient reduction in calcium and magnesium to permit the 15 cycles.

A sufficient amount of water undergoes this split treatment to provide cooling tower make-up and feed for the high pressure boiler make-up demineralizers.

3.7 AREA 117 - WATER TREATMENT

3.7.8 Alternate 1

Process Description - continued

Carbon Filtration

Carbon filtration is provided ahead of the demineralizers to protect them from possible carry-through of chlorine from chlorination of the raw water, or from solids resulting from post-precipitation of the excess lime-softened water. Additional carbon filters, ahead of the condensate polishers, protect them from possible organic contaminants.

Demineralization

The demineralizer system consists of cation exchange units, forced draft degasifiers, and anion exchange units which provide additional treatment to produce the quality required for high pressure boiler make-up water.

Condensate Polishing

Mixed bed condensate polishing units remove any mineral contaminants which may be carried by the condensate return. As an added safety precaution for the high pressure boiler make-up, the distillate from the inorganic dissolved solids evaporator is also run through carbon filters and the condensate polishers.

System Capability

Table 3.7.8-1 provides a tabulation of the water analysis expected from each of the unit treatment processes and of the 15 cycle cooling tower water composition.

3.7.9 Alternate 2

Introduction

This basis of this alternate for the treatment requirements for the various water uses is by use of zeolite softeners and moving bed ion exchange units. The moving bed system greatly

TABLE 3.7.8-1

System Operating Results External Treatment

Demineralization (mg/l) — Cooling Tower 15X

	AFTER RE-CO_2 + FILT	WAC	50-50 WAC-LS	CAT	DEGAS	ANION	MB	BA	AA
Ca	45	--	22	--	--	--	--	330	330
Mg	25	--	13	--	--	--	--	195	195
Na	183	166	175	0-1	0-1	0-1	TR	2625	2625
H	--	--	--	166	166	0	--	--	--
	253	166	210	166	166	0-1	TR	3150	3150
HCO_3	82	0	44	0	0	--	--	660	42
CO_3	5	0	0	0	0	--	--	--	--
OH	0	0	0	0	0	0-1	TR	--	--
Cl	42	42	42	42	42	--	--	630	630
SO_4	105	105	105	105	105	--	--	1575	2193
NO_3	19	19	19	19	19	--	--	285	285
	253	166	210	166	166	0-1	TR	3150	3150
SiO_2	10	10	10	10	10	TR	TR	150	150
S.S.	1	--	--					74*	74*
TDS	374	240	310						
CO_2	0	74	37	78	10	0	0	--	--
pH	8.7		6.4					9.2	7.0

		BA	AA
Ca x SO_2	(10^3)	520	724
Mg x SiO_2	(10^3)	29	29

Notes:
WAC Weak acid cation exchange
LS Lime soda softened
MB Mixed bed (Condensate Polishing)
BA Before acidification
AA After acidification
* Includes SS pick-up from air through cooling tower

3.7 AREA 117 - WATER TREATMENT

3.7.9 Alternate 2

Introduction - continued

reduces the regeneration waste stream volume. The proposed system has eliminated the lime-soda softening unit process with its attendant sludge creation and disposal problems. Block Flow Diagram No. 117-120-1A2 presents the general concept of this process.

Process Description

Clarification and Filtration

The raw river water is treated with polymer coagulants and/or floculants to aid in the removal of suspended solids. The detention time provided by the primary clarifiers is such that the greater portion of the suspended solids will be removed. To protect the carbon filters and ion exchange units from solids build-up and excessive backwash requirements, the clarifier effluent is filtered.

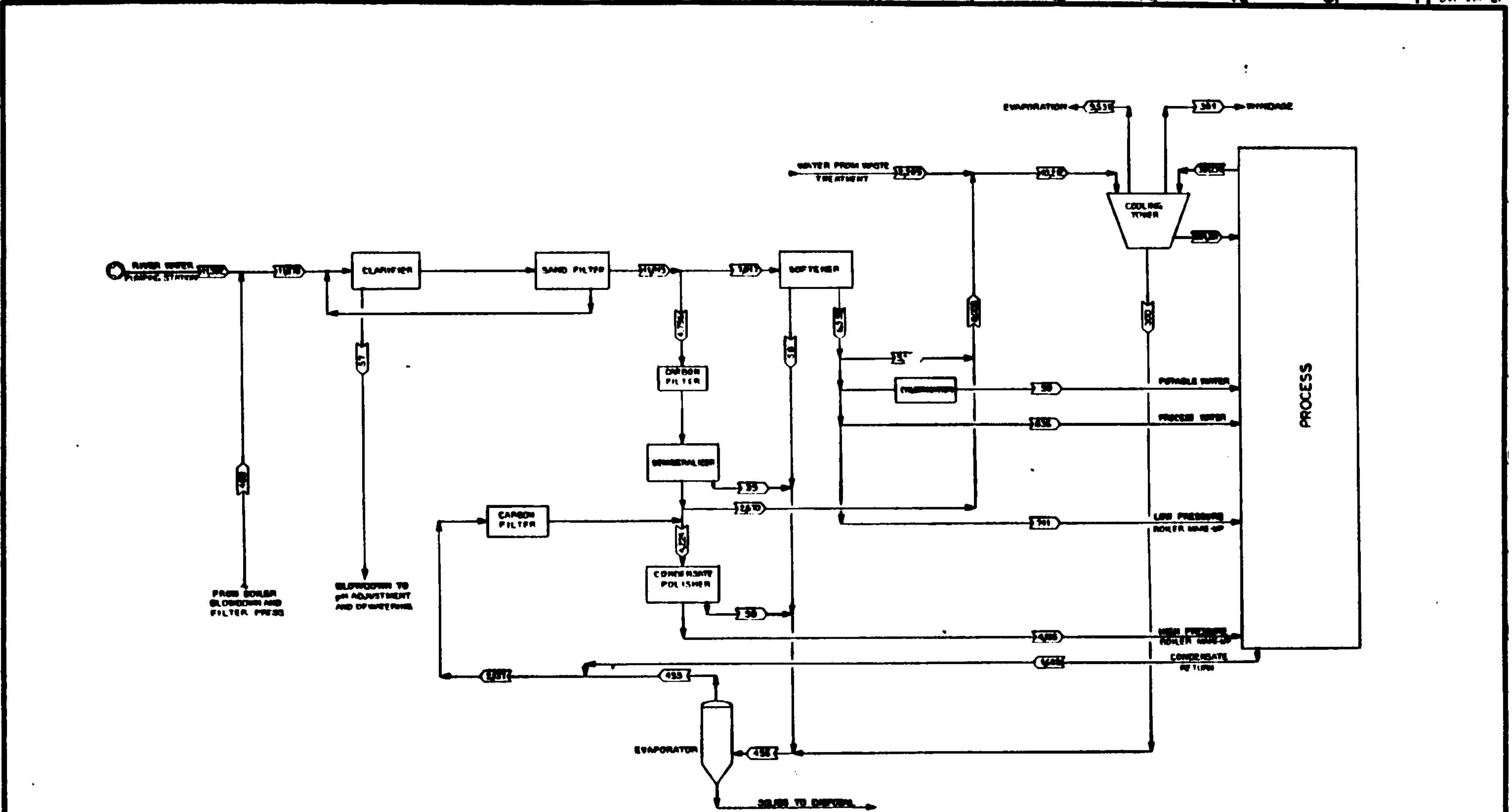
Softening

Zeolite softening units, regenerated with sodium chloride solution, soften to the degree required for providing potable, process and low pressure boiler make-up water; and are sufficient for providing the cooling tower with make-up permitting 15 cycles (again SiO_2 limited).

Carbon Filtration

The fresh water portion of the high pressure boiler water make-up is provided by a filtered water stream separate from that feeding the softeners. This second stream is treated by carbon filtration to remove any chlorine which may remain from the chlorination of the raw water at the River Water Pumping Station.

A second set of carbon filters is provided for the evaporator distillate and condensate return to remove possible organic contamination.



NOTE: CARBON FILTER BACKWASH WILL CONSIST OF INTERMITTENT FLOWS AND WILL BE DISCHARGED TO BIOSYSTEM

BLOCK FLOW DIAGRAM
WATER TREATMENT
(RELATED TO BASE CASE WASTEWATER TREATMENT ALTERNATE - 2)

PROJECT NO 7061

117-120-1A2

3.7 AREA 117 - WATER TREATMENT

3.7.9 Alternate 2

Process Description - continued

Demineralization

The fresh water portion of the high pressure boiler water make-up is processed by anion-cation demineralization units. A sufficient portion of the cooling tower make-up is also handled by by the anion-cation demineralization units to reduce silica to a level compatible with 15 cooling tower cycles.

Condensate Polishing

The carbon filtered condensate return and evaporator distillate, after carbon filtration, are routed through a mixed bed condensate polisher to assure high pressure boiler make-up water quality.

3.7.10 Technical and Economic Evaluation

Evaluation Criteria

Technical and economic evaluations of the three processes under consideration are weighted in accordance with the following criteria:

Total Capital Requirement - 20%

Total Annualized Cost - 40%

Technical Evaluation Matrix - 40%

Total capital requirements, including installed cost, working capital, and fees, are estimated for each process on a consistent basis. The least capital intensive process is awarded 20 points The bottom score is obtained by dividing the lowest capital requirement by the highest and multiplying by 20. The system falling within the range between maximum and minimum is awarded points by interpolation. Annualized costs are estimated and evaluated in the same fashion as capital requirements. Lowest annualized costs receive 40 points, etc.

3.7 AREA 117 - WATER TREATMENT

3.7.10 Technical and Economic Evaluation

Total Capital Requirements

The total estimated capital requirement for the Base Case and the two (2) alternatives under consideration is summarized in Table 3.7.10-1.

The installed cost for each system was developed from equipment costs and installation factors on a consistent basis. It includes all material, labor, and indirect costs necessary to install the unit on a battery limits basis. Inter-area costs such as pipe racks, utility generating facilities, etc. are not included.

Working capital normally consists of the sum of inventories, accounts receivable, cash equivalents, and net of accounts payable. The working capital for the Process Design Study consists of the sum of inventories only and includes estimated costs for initial charge of catalyst and/or chemicals, sixty (60) days supply of catalyst or chemicals added on a continuous basis, sixty (60) days of operating supplies and thirty (30) days of maintenance supplies. Accounts receivable, cash equivalents, and net of accounts payable are NOT included in the working capital for the Process Design Studies; however, these cash amounts are included in the working capital requirement for the baseline economic evaluation of the Pipeline Gas Commercial Plant.

Table 3.7.10-1

Total Capital Requirement
($ in Thousands)

System	Change from Base Case	Installed Cost $	Working Capital $	Total
Base Case	0	16,102	157	16,259
Alternate 1	+5992	22,056	195	22,251
Alternate 2	-474	15,553	232	15,785

3.7 AREA 117 - WATER TREATMENT

3.7.10 Technical and Economic Evaluation

Annualized Costs

Total annualized costs for the Base Case and the two (2) alternatives under consideration are summarized in Table 3.7.10-2. Consumption rates for reagents and utilities, together with labor requirements, are based on licensor data reported in each alternate's sub-section. Unit costs for reagents, utilities, etc. and the basis for capital intensive annualized costs are outlined below.

Raw Materials

Soda Ash	\$55/ton
Lime	\$37/ton
Sulfuric Acid	\$52/ton
Sodium Hydroxide	\$100/ton
Anionic Polymer	\$4,850/ton

Utilities (See Section ^.0, sub-section 6.6)

Electricity	\$0.023/KWH
Raw Water	\$0.43/M gal
Instrument Air	\$0.04/M SCF

Labor

Operating	\$60,000/yr/Operator Position
Maintenance	1.5% of Area Installed Cost
Supervision	15% of Operating & Maintenance Labor
Administration & Overhead	60% of Total Labor

Waste Disposal — \$2.50/ton for trucking to landfill and level.

Supplies

Operating	30% of Operating Labor
Maintenance	100% of Maintenance Labor

Indirect Operating Costs

General Expense	0.1% of Area Installed Cost
Local Taxes & Insurance	0.5% of Area Installed Cost

3.7 Area 117 - WATER TREATMENT

3.7.10 Technical and Economic Evaluation

Annualized Costs - continued

Capital Charges

Depreciation	5% of Area Installed Cost
Average Capital Charge	11% of Area Installed Cost

Capital Charge Credit

Alternate I results in reduced cooling tower blowdown, which in turn requires less evaporator capacity and reduced capital costs for Area 118, Waste Treatment and Disposal. This reduction in Area 118 capital costs was credited to Area 117. Likewise, Alternate 2 effects a reduction in both the evaporator size and the number of sludge presses required in Area 118. This capital cost reduction was credited to the Best Technology water treatment process alternate.

3.7 AREA 117 - WATER TREATMENT

3.7.10 Technical and Economic Evaluation

Table 3.7.10-2

Annualized Cost Summary

Annual Cost $	Base Case	Alternate 1	Alternate 2
Direct Operating Costs			
Raw Materials	$ 634,700	$ 854,600	$1,121,000
Utilities	2,438,300	2,440,000	2,455,100
Labor	775,600	939,900	650,100
Supplies	295,500	384,800	269,300
Sub-Total Direct Costs	4,144,100	4,619,300	4,495,500
Indirect Operating Costs	96,600	132,300	93,300
Total Net Operating Cost	4,240,700	4,751,600	4,588,800
Capital Charges	2,576,300	3,529,000	2,488,500
Capital Charges Credit	-------	386,100	914,300
Net Capital Charges	2,576,300	3,142,900	1,574,200
Total Annualized Costs	$6,817,000	$7,894,500	$6,163,000
Equivalent Gas Cost Contribution, $/MM Btu SPG	0.084	0.098	0.077

3.7 AREA 117 - WATER TREATING

3.7.10 Technical and Economic Evaluation

Technical Evaluation

Technical evaluation of the three (3) processes is based on five (5) criteria, namely: state of development; process flexibility/operability; reliability; design package; and the impact the process has on waste treatment requirements. For evaluation purposes, each criterion is assigned points and presented in matrix form in Table 3.7.10-3. The process having the highest rating in a given criterion is given 8 points in that category. Each of the other processes is given a point rating, relative to the maximum 8 points, for each of the categories. The process with the highest technical rating is awarded 40 evaluation points. The lowest technical rating is divided by the highest rating and multiplied by 40 to obtain the bottom score. Other scores are determined by linear interpolation between highest and lowest in the same manner as the economic evaluation.

Table 3.7.10-3

Technical Evaluation Matrix

Criteria	Base Case	Alternate 1	Alternate 2
State of Development	8	8	7
Flexibility/Operability	7	7	8
Reliability	8	8	8
Design Package	8	8	5
Waste Treatment Impact	5	6	8
Total	36	37	36
Total on 40 pt. basis	39	40	39

3.7 AREA 117 - WATER TREATMENT

3.7.10 Technical and Economic Evaluation

Process Technology Summary

Base Case

The lime-soda ash softening process is a widely used treatment process, especially in the municipal water treatment field. Softening, however, is limited to 40-60 mg/l hardness as calcium carbonate. In this case, this hardness (which is primarily magnesium hardness) results in limiting cooling tower cycles to 7-10. The process requires frequent routine analysis to maintain the proper chemical dosages and pH value; and frequently raw water variations require trial and error adjustments of the amount and kind of chemical additives to maintain a uniform, optimum treated water quality.

The additional treatment necessary for high pressure boiler water make-up is provided by a cation deionization unit, followed by anion units, to completely remove calcium, magnesium, and silica. Dual-media mixed bed condensate polishers provide a safeguard for reuse of possibly contaminated condensate in the high pressure boiler system.

The demineralization and mixed bed polishers are described in detail in Subsection 3.7.8.

Alternate 1

The basis of the process is its use of established unit processes in a unique combination to permit increasing the cooling-tower cycles, thus reducing the blowdown and the required evaporation capacity for blowdown treatment. All units are well-developed, reliable treatment devices. The excess lime-soda softening suffers from the same operational problems stated for the Base Case. The softened water, after additional softening by the weak acid cation exchanger, is suitable for cooling tower and low pressure boiler water make-up. Demineralizers, as provided in the Base Case, are used in treating the softened water for high pressure boiler make-up. Condensate is treated by the same system as that used in the Base Case.

3.7 AREA 117 - WATER TREATMENT

3.7.10 Technical and Economic Evaluation

Process Technology Summary - continued

Alternate 2

This approach replaces the lime-soda softening step with zeolite softening, thus removing some of the operational problems of the other two processes. The state of the art for the softeners is as well advanced as that of lime-soda softening. Available demineralization and deionization units reduce waste production through the use of a moving bed, regeneration technique. This approach was developed some time ago but lost favor because of mechanical problems. The equipment supplier believes it has solved these problems, and has developed equipment to provide deionization in a manner resulting in a considerable reduction in waste treatment requirements.

Operating Considerations

Operation of both the Base Case and Alternate 1 requires rather voluminous analytical work and close attention to both the lime softening and clarifier operations to maintain a high quality effluent. Alternate 2 eliminates the lime-soda operation and lends itself more readily to automation.

The moving-bed demineralizers technology is quite well developed. However, in its early stages, there were some problems with resin transfer and regeneration. The equipment supplier has stated it has solved these problems and minimized the volume of spent regenerent.

Recommendations

Alternate 2 is the recommended process on the basis of lowest installed cost as well as lowest annulaized cost. Table 3.7.10-1 indicates capital savings of $549,000 over the next least expensive process. The Annualized Cost Summary, Table 3.7.10-2, shows considerable savings in annualized costs, resulting in a 0.15% savings in projected gas cost. The relative ratings of the processes, evaluated in accordance with the previously described evaluation criteria, is shown by Table 3.7.10-4.

3.7 AREA 117 - WATER TREATMENT

3.7.10 Technical and Economic Evaluation

Recommendations - continued

Table 3.7.10-4

Process Selection Matrix

	Total Capital Requirement	Total Annualized Cost	Technical Evaluation	Total Points
Base Case	19	34	39	92
Alternate 1	14	36	40	90
Alternate 2	20	40	39	99

3.8 AREA 119 - FIRE PROTECTION SYSTEMS

3.8.1 Summary

The Fire Protection System is a complete and operable process as described in Section 5.0, Baseline Process Concept, Subsection 5.3 - Process Description. The system, by its nature, is not suitable to process variations; therfore, no alternate processes could be considered. System cost is based upon size alone, and the estimated cost used in overall cost estimate is based upon the plant site layout as shown by Drawing No. 100-200-1.

Costs remain as defined in Section 6.0, Baseline Economic Evaluation, sub-section 6 3 - Capital Requirement, Figure 6.3-1.

Preliminary information concerning soil conditions indicates it may be necessary to construct the firewater storage pond in such a manner as to assure its being impervious. Seepage into the soil could be a water supply problem of considerable magnitude were an impervious basin not provided.

This problem will be given serious consideration once the nature of the soil at the pond location has been defined. Several methods will be investigated for comparison of installed cost, maintenance, reliability and life expectancy. Among the methods considered will be various types of membranes, relatively rigid linings (i.e. Gulf Seal and asphalt paving), and colloidal clays which become an integral part of the earth material used in forming the basin. Baroid, Dowell, and Volclay are in this latter category. Soil cement and soil asphalt may provide additional alternatives depending on the soil character.

3.9 AREA 120 - FACILITIES

General

The facilities which comprise Area 120 provide services or functions essential to the successful, continuous operation of the plant. Table 3.9-1 itemizes the facilities included as part of the Commercial Plant and identifies the functions and/or services rendered by each.

Physically, Area 120, unlike the process areas, does not consist of a neatly clustered arrangement of tanks, towers and pumps. Instead, the structures, systems, and components which comprise the facilities appear to be arranged in an almost random pattern throughout the plant. This arrangement is mandated by the particular service or function which the unit is providing and the need for it to be located near the plant area it is serving.

Many of the facilities can be found on the plot plans contained in Subsection 5.9. Brief descriptions of the facilities and their functions are given in Subsection 5.3, Area 120 - Facilities.

Additional particulars on storage tanks, buildings, and sewers are described in the following paragraphs.

Tanks

Fixed roof tanks are presently planned for use in storing product oils. A study was initiated to determine if fixed roof storage will permit compliance with governing safety and environmental regulations.

Certain physical data are required to make this determination. The flash point must be known, for if it is too low, flame arrestors must be provided to protect the product oils from external sources of fire. The vapor pressure at storage temperature is the criteria used to determine if floating roof storage tanks or some other vapor conservation method, such as pressure storage vessels or a vapor chamber, must replace or be used in conjunction with the fixed roof tanks. The accurate determination of storage loss rates, required for the economic analysis of competing storage systems, depends on the slope of the distillation curve. These physical data

3.9 AREA 120 - FACILITIES

Facility \ Services Provided	Storage	Transportation	Waste Collection	Distribution	Security	Safety	Reliability	Operability	Maintainability	Administration	Environmental	Quality Control	Employee Benefit
Synthetic Gas Pipeline		X											
Administration Bldg.										X			X
Maintenance Shop	X					X	X	X	X				X
Laboratory						X	X	X			X	X	X
Personnel Facilities						X							X
Railroad Facilities	X	X						X					
Plant Roads		X	X	X	X			X	X	X			
Parking Areas					X					X			X
Distribution Piping		X		X				X					
Plant Fence and Guardhouses					X	X				X			
Lighting					X	X		X	X	X			X
Storage Areas	X		X	X	X	X	X	X	X	X	X	X	
Potable Water System	X			X		X							X
Sanitary Sewer System			X			X					X		X
Storm Sewers			X		X	X	X	X	X		X		
Chemical Sewers			X		X	X					X		X

Table 3.9-1 Commercial Plant Facilities/Required Services or Functions

3.9 AREA 120 - FACILITIES

Tanks - continued

are not yet available; however, they will be available for the Revised Tentative Baseline Design (RTBD). At that time, the oil storage facilities will be reviewed and redesigned if necessary.

By-products and chemicals are stored in standard industrial facilities. Vapor losses from the ammonia and light hydrocarbon storage vessels are minimized by the use of vapor recovery compression-refrigeration systems.

All storage tanks will be diked as required to satisfy safety and environmental standards.

Buildings

The administration building, which is to accommodate 150 to 200 personnel, is a brick structure housing the general plant offices, data processing center, classrooms/meeting rooms, visitors' reception area, plant cafeteria, and first aid station.

A prefab maintenance building serves as the plant maintenance center. Freeboard is provided in this building for the operation of a mobile crane.

Spare parts and equipment, along with personnel supplies, are stored in the warehouse, a prefab structure. The warehouse is also equipped with mobile lifting equipment for the handling of stored material.

The plant laboratory is a concrete block structure which contains the instrumentation and equipment necessary to ensure proper plant operation and product quality. The fire station is a prefab structure, and houses plant fire-fighting personnel and equipment. The gatehouse is a brick structure and serves as the center for matters pertaining to plant security.

Sewers

Sewers are provided to collect sanitary wastes, contaminated plant runoff, and chemical spillage or process wastes. The sanitary sewer system collects wastes from all toilet facilities, lunch rooms, locker rooms and shower rooms and directs them to the sanitary treatment plant.

3.9 AREA 120 - FACILITIES

Sewers - continued

Contaminated plant runoff due to rainfall and runoff from areas of high potential for contamination are collected in the stormwater impoundment basin. From there, it is pumped to Area 118 for treatment. A chemical sewer is provided to collect spillage and/or overflows from storage areas, process areas, and loading/unloading areas, and divert them to Area 118 for treatment.