ASSESSMENTS AND PROCESS SELECTION

OXYGEN PURITY

Introduction

The purpose of these subtasks is to consider the effects of oxygen purity on the operation and economics of a plant producing medium Btu gas for pipeline delivery to various industrial users. Medium Btu gas is to be produced by the gasification of Kentucky #9 coal under pressure and using oxygen as the gasification medium. Design coal feed rate is 20,000 ST/D with a nominal gas production equivalent to about 300 x 10^9 Btu's/day for a total of four operating modules. Gas heating value is to be at least 285 Btu/SCF on a higher heating value basis.

Figure 1 is a block diagram showing the system considered in these assessments. This scheme applies equally to each of the five gasification processes being considered in the overall study, i.e. BGC/Lurgi slagging gasifier, Lurgi dry ash gasifier, Babcock & Wilcox entrained flow gasifier, Texaco entrained flow gasifier and Koppers Totzek entrained flow gasifier.

Oxygen Production

Total oxygen requirements are in the range of 2250 ST/D to 4500 ST/D per module depending upon the gasification process. Due to the existing technology with the largest present plants having a capacity of 2500 T/D in a single train, the high oxygen demand gasification processes will require two air separation plants per module.

Several air separation unit process schemes are possible. These are summarized in the following table.

		Relative Power	Relative Capital
1.	Low Pressure with Reversing Exchangers	1.0	1.0
2.	Low Pressure with Regenerators	1.05	1.03
3.	Low Pressure with Molecular Sieve	1.06	1.05
4.	Pumped Liquid-split Cycle	1.08	1.00
5.	Pumped Liquid N_2 Cycle	1.08	1.00

While the above data are for an oxygen product delivery pressure of 1200 psig, the conclusions are believed valid for the present conditions. Accordingly, the low pressure cycle with reversing exchangers has been selected for this project.

The air plant will require air at about 85 psig and will deliver a gaseous oxygen stream at about 0.5 psig. This stream must then be compressed to meet gasifier requirements.

Oxygen Purity

Air plants typically produce gaseous oxygen with purities in the range of 95 to 99.8%. Purities of less than 90% can be obtained by blending air with the product oxygen. Minimum power occurs at about 95% purity with slightly greater power usage required to obtain higher purity due to the need for additional fractionation trays. The power requirement increases below 95% due to the greater volume of gas - air plus oxygen - which must be compressed to meet the gasifier requirements.

Variations in oxygen purity are limited by the final product specification of 285 Btu/SCF GHV. The result of this limitation is shown in Figure 2. The results are approximate as gasifier heat and material balances were not generated for various oxygen purities.

Figure 2 shows that all three entrained flow gasification processes are relatively sensitive to oxygen purity in terms of the 285 BTU/SCF minimum heating value for the product gas. The minimum oxygen purity which meets the gas product specification for the B&W and Koppers Totzek gasifiers is around 88%. The Texaco process requires that a

Form No. 130-171

portion of the carbon dioxide be removed from the product gas to meet the minimum heating valve. At the 50% removal level for CO_2, The Texaco gasification process req es a minimum of 93% oxygen purity.

The BGC/Lurgi slagging gasifier, due to the higher methane content of the gas, is less sensitive to oxygen purity and can meet the desired gas specification with oxygen purity as low as 50%. The Lurgi dry ash gasifier is assumed to behave in similar fashion to the slagger.

Overall Effect

Figure 1 shows a block flow diagram for the plant. Variations in oxygen purity have an effect on oxygen compression, gasification, purification, and product gas compression. Although a precise analysis of the effect of oxygen purity was not attempted in this assessment, valid conclusions can be drawn from the data developed during the course of this study.

Compression brake horsepower has been estimated for the oxygen and air compressors. Each of these machines will require additional horsepower as oxygen purity decreases. This increase in shown in Figure 3. Incremental power from product gas compression has been estimated as a function of oxygen purity. Figure 4 shows the effect of oxygen purity on compression requirements. The compressor power requirement is total air + oxygen power plus the incremental power for product gas compression as the oxygen purity declines from the base of 99.8%. This curve bottoms out at about 97 to 98% oxygen purity.

Capital costs have been supplied by Air Product and Lotepro. The Air Products numbers were used for convenience in preparing this report. Figure 5 shows the effect of oxygen purity on the turnkey price of the oxygen plant of 2 units each as a function of purity and tonnage per unit.

Decreased oxygen pruity effects the design of the gasification facility and downstream equipment. This effect results from the increasing quantity of nitrogen flowing through the plant as oxygen purity decreases. Figure 6 was plotted assuming that plant costs vary to the six tenths power. Module costs varyings from 200 to 500 MM$ were assumed. The capital costs shown represents the cost of the Air Plant plus the gasification portion incremental cost.

Supplier Contacts

The following potential suppliers for the air plants were contacted regarding this question:

1. Air Products and Chemicals, Inc.
   Box 358
   Allentown, Pa. 18105
   215-398-8540

2. American Air Liquide, Inc.
   405 Lexington Avenue
   New York, N.Y. 10017
   Mr. Frank Wolff
   212-867-3060

3. Lotepro, Inc.
   1140 Avenue of the Americas
   New York, N.Y. 10036
   212-575-7878

4. Union Carbide Corporation
   Linde Division - 8th Floor
   270 Park Avenue
   New York, N.Y. 10017
   212-551-4293
   Mr. Don Curran

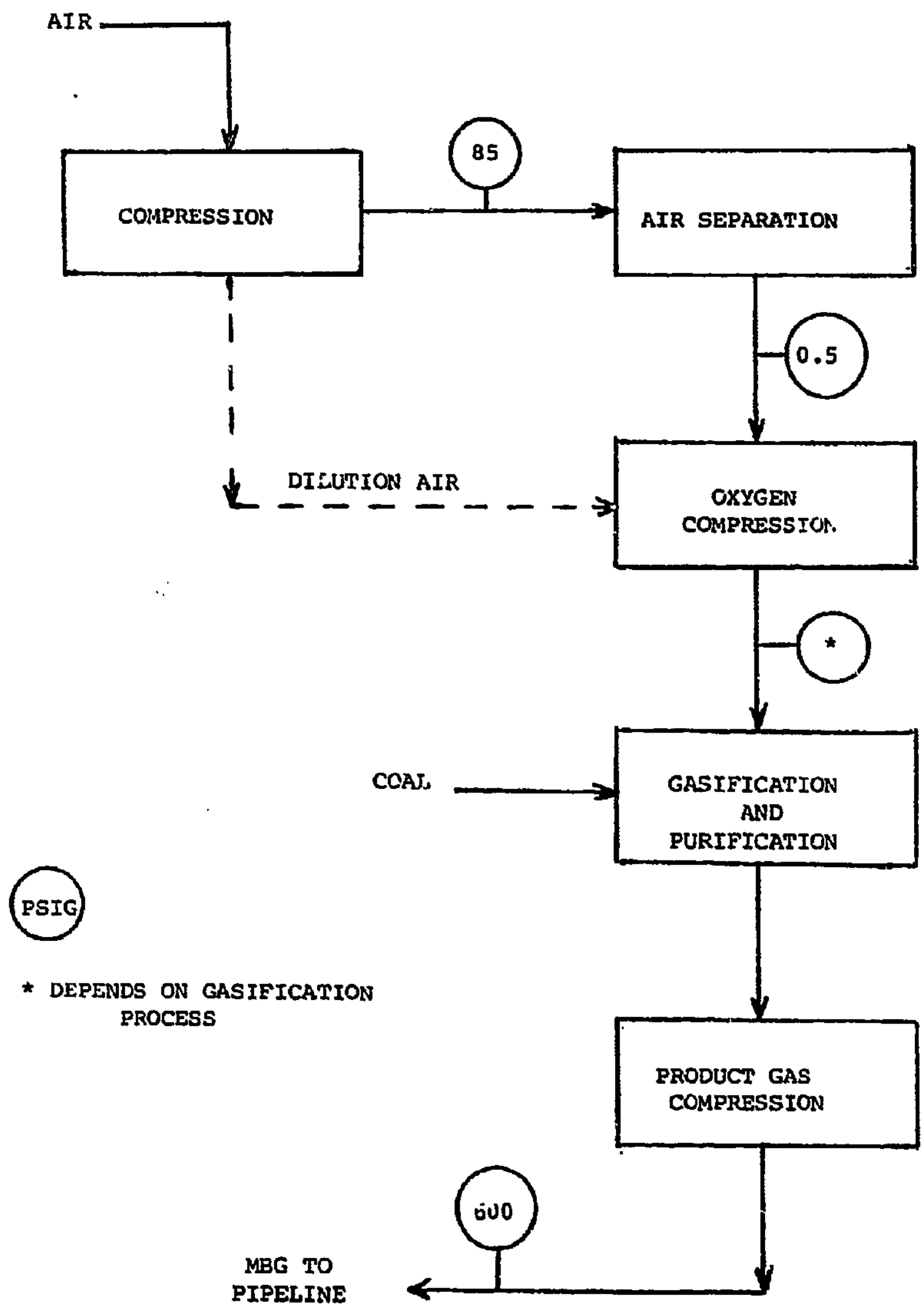
Information was received from each of the suppliers. This information was specific to this project and supported the work done by Foster Wheeler in evaluating the question of oxygen purity.

Conclusions and Recommendations

It is the conclusion of this study that oxygen of high purity be produced and used in the gasifier. Dilution of the product gas with nitrogen will result in an overall increased operating cost. The air plants should be specified to produce 98% oxygen which appears a reasonable value based on the data shown in this assessment.

FIGURE 1

BLOCK FLOW DIAGRAM FOR MGB PRODUCTION



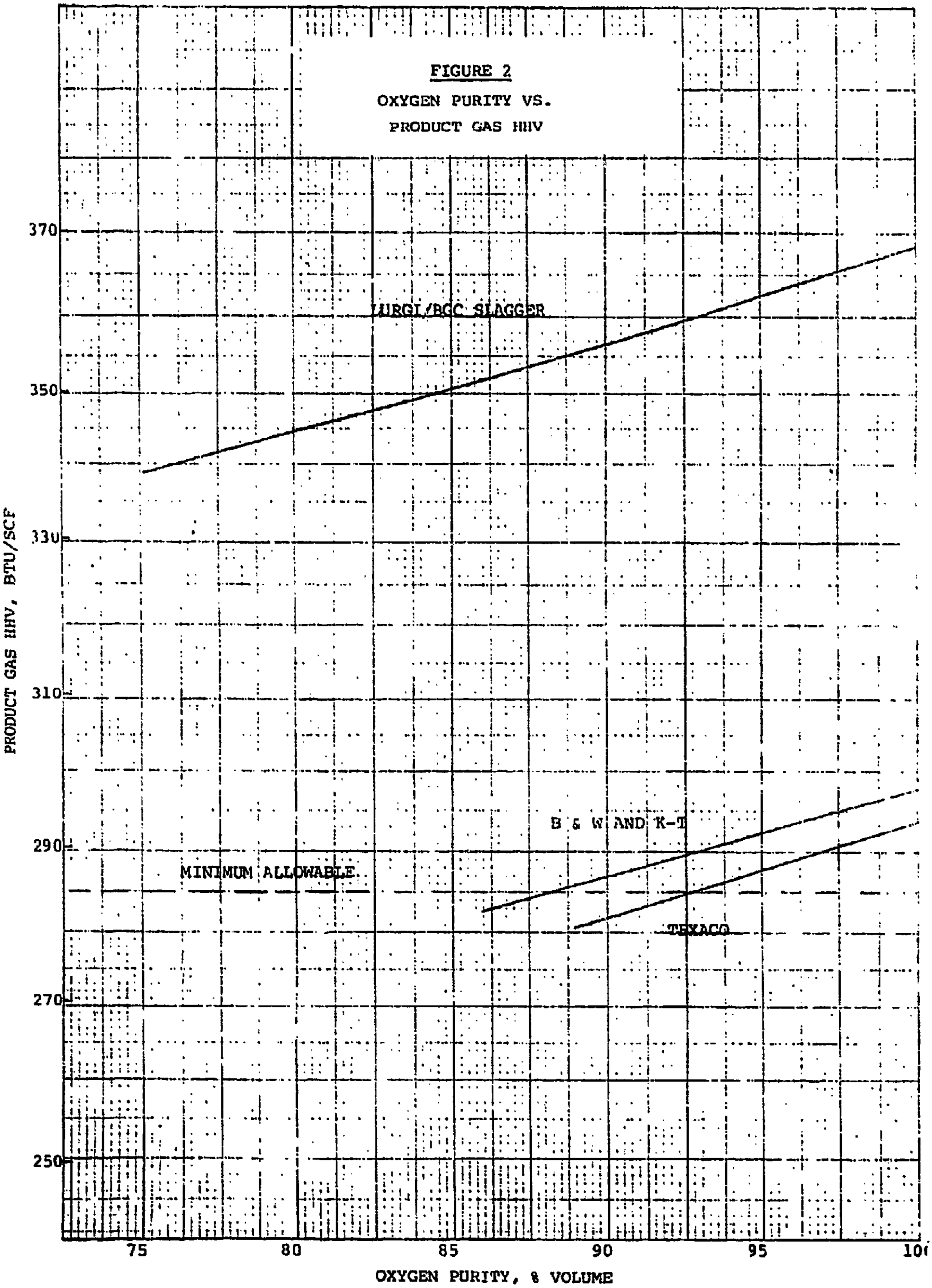


FIGURE 2
OXYGEN PURITY VS.
PRODUCT GAS HHV
LURGI/BGC SLAGGER
B & W AND K-T
MINIMUM ALLOWABLE
TEXACO
370
350
330
310
290
270
250
PRODUCT GAS HHV, BTU/SCF
75
80
85
90
95
100
OXYGEN PURITY, % VOLUME

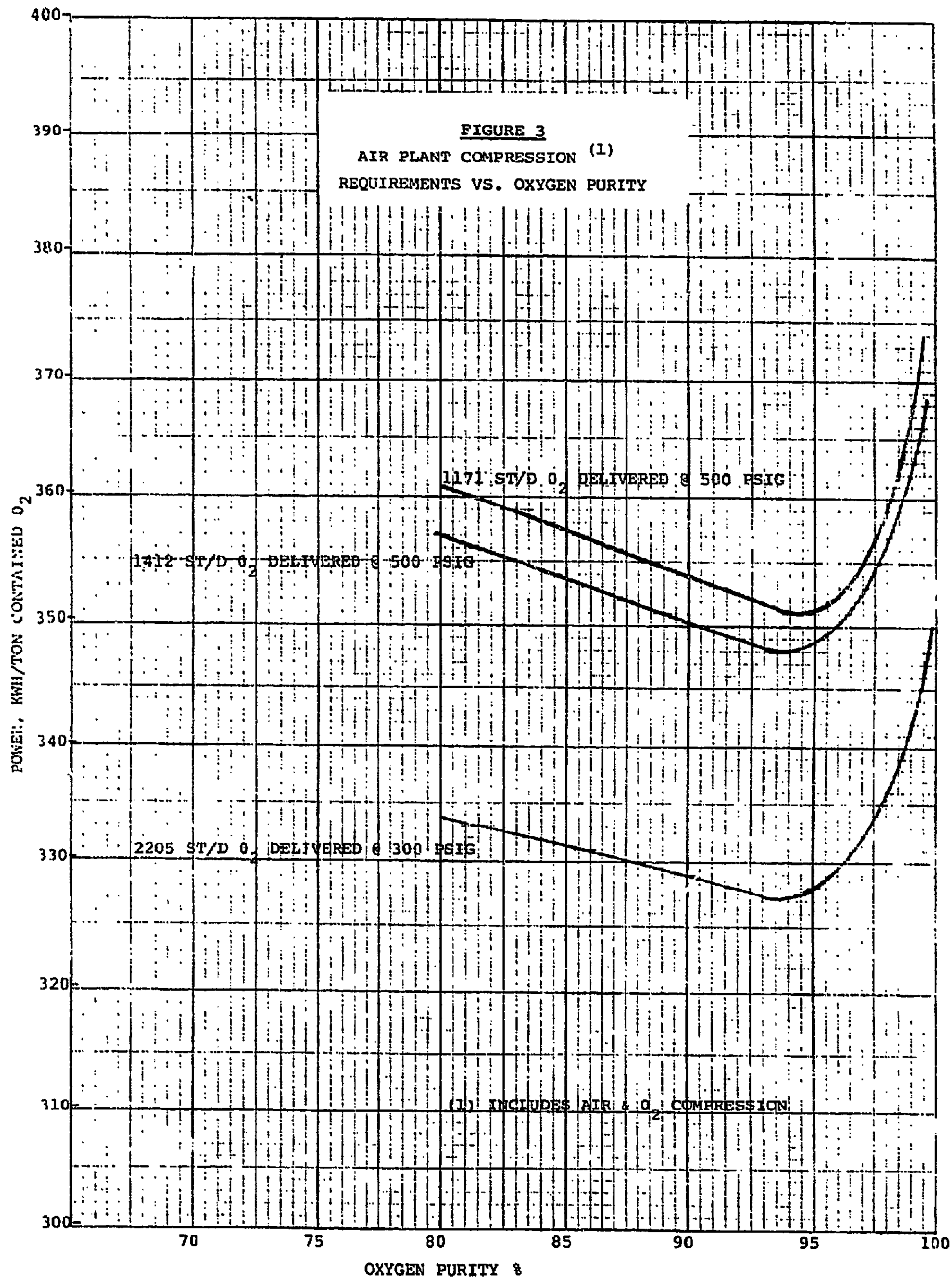


FIGURE 3
AIR PLANT COMPRESSION (1)
REQUIREMENTS VS. OXYGEN PURITY
1171 ST/D O_2 DELIVERED @ 500 PSIG
1412 ST/D O_2 DELIVERED @ 500 PSIG
2205 ST/D O_2 DELIVERED @ 300 PSIG
(1) INCLUDES AIR & O_2 COMPRESSION
POWER, KWH/TON CONTAINED O_2
OXYGEN PURITY %
400
390
380
370
360
350
340
330
320
310
300
70
75
80
85
90
95
100

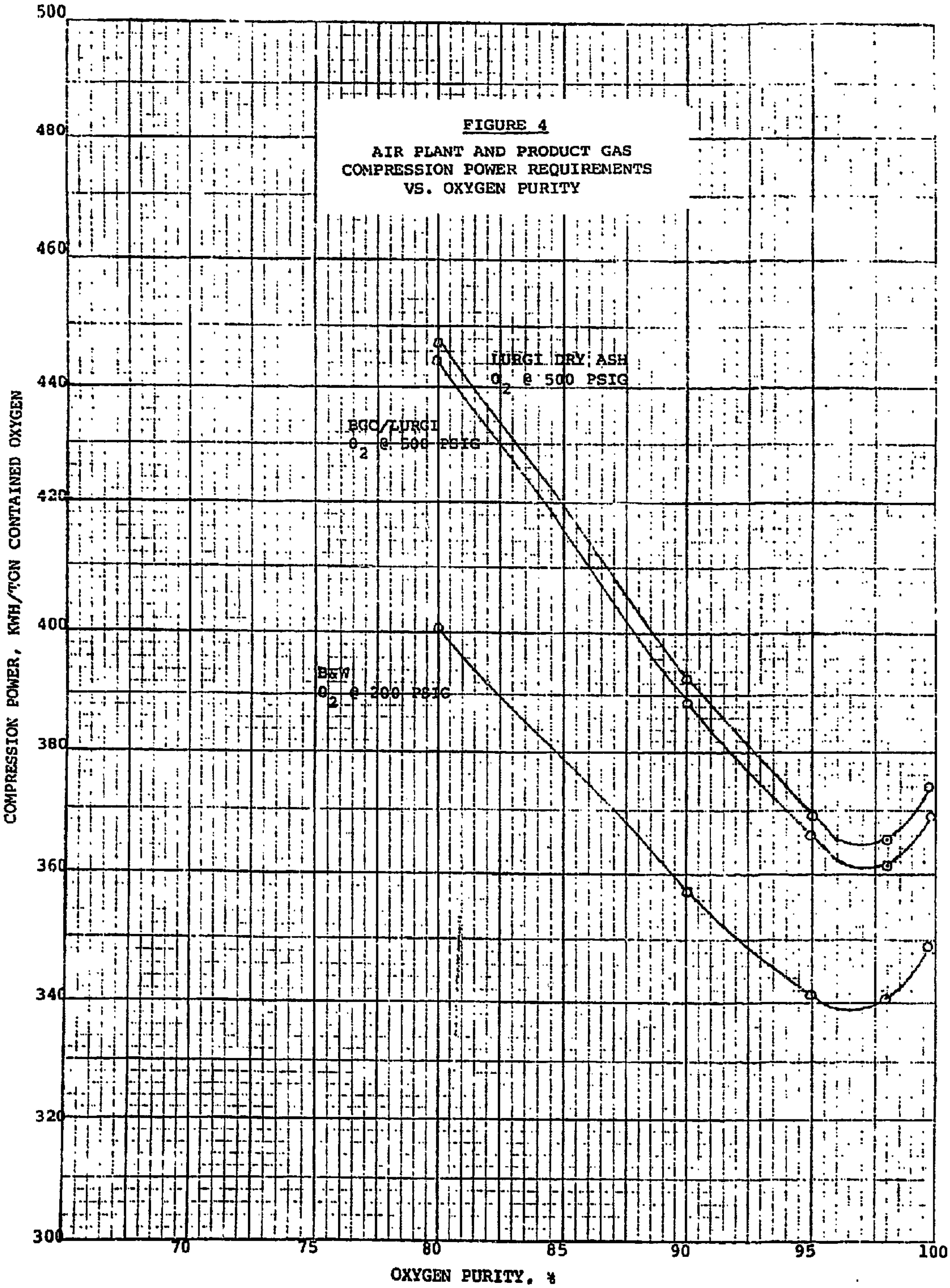


FIGURE 4

AIR PLANT AND PRODUCT GAS
COMPRESSION POWER REQUIREMENTS
VS. OXYGEN PURITY

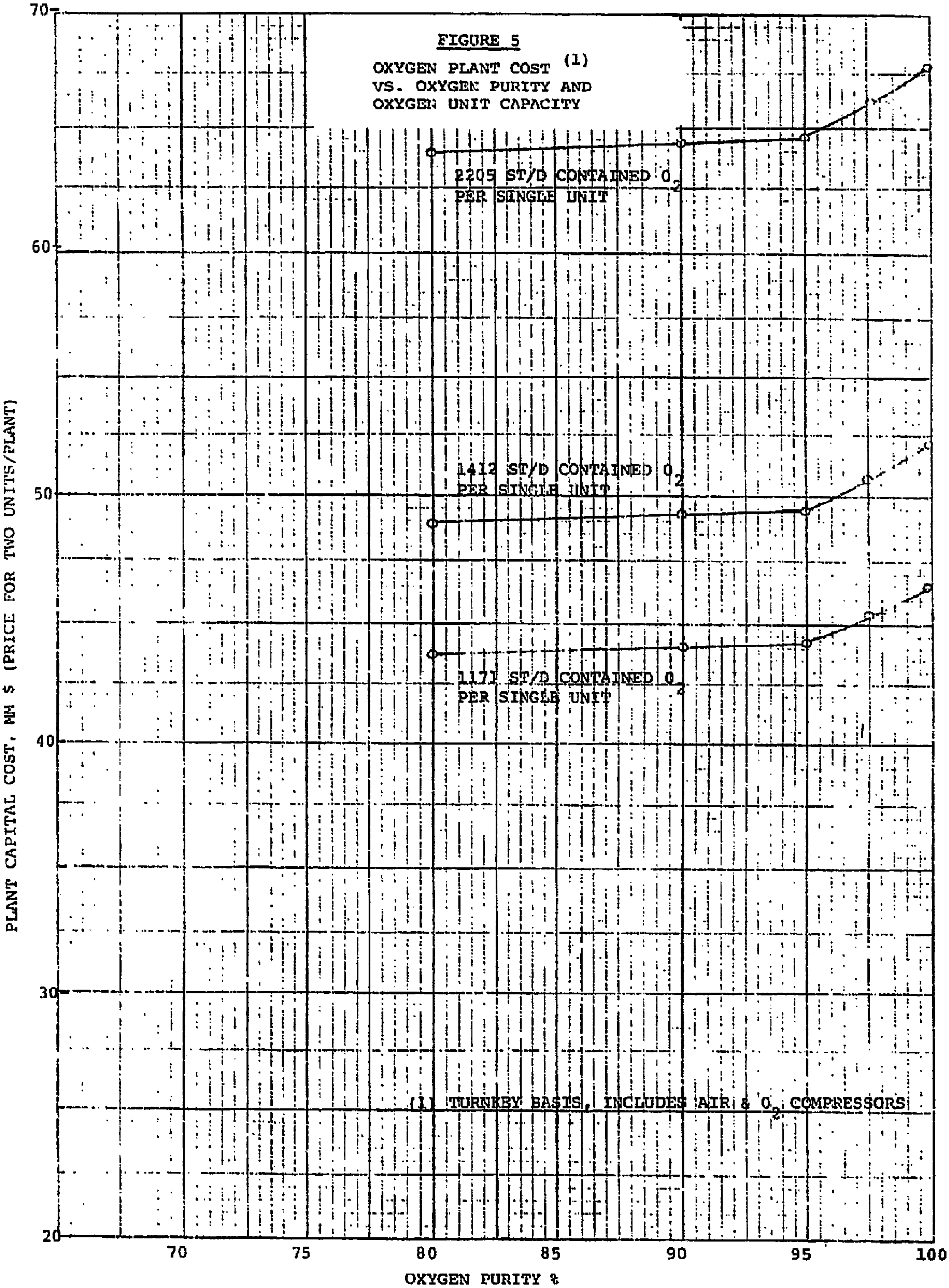


FIGURE 5
OXYGEN PLANT COST (1)
VS. OXYGEN PURITY AND
OXYGEN UNIT CAPACITY
2205 ST/D CONTAINED O_2
PER SINGLE UNIT
1412 ST/D CONTAINED O_2
PER SINGLE UNIT
1171 ST/D CONTAINED O_2
PER SINGLE UNIT
(1) TURNKEY BASIS, INCLUDES AIR & O_2 COMPRESSORS
PLANT CAPITAL COST, MM $ (PRICE FOR TWO UNITS/PLANT)
OXYGEN PURITY %
70
60
50
40
30
20
70
75
80
85
90
95
100

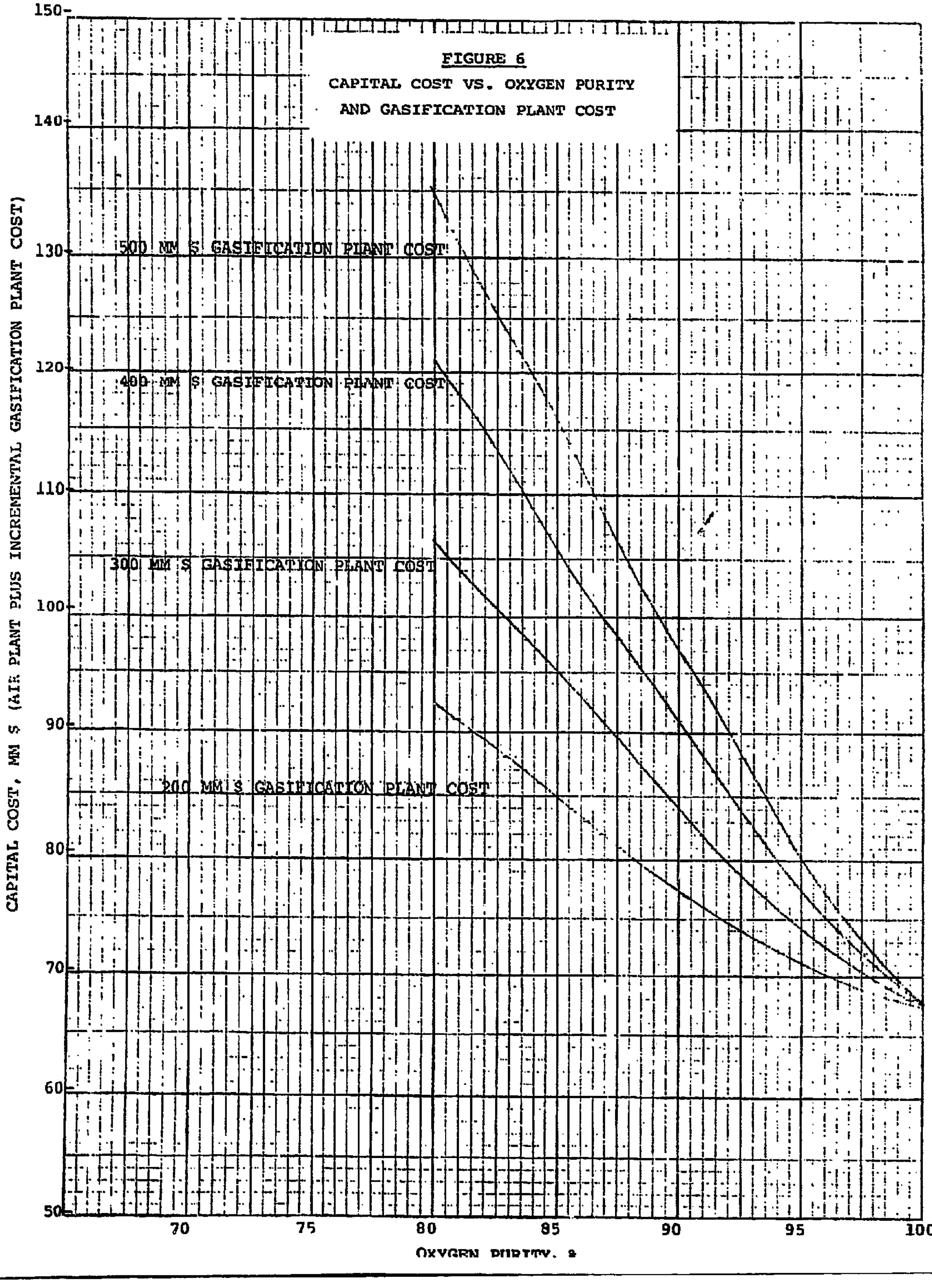


FIGURE 6

CAPITAL COST VS. OXYGEN PURITY AND GASIFICATION PLANT COST

SECTION 3.3

COAL WASHING

ASSESSMENTS AND PROCESS SELECTION

COAL WASHING

Introduction

The objective of these subtasks is to assess benefits and costs associated with coal washing prior to coal gasification. These assessments were made in the context of 20,000 TPD coal gasification plants using Lurgi dry ash and BGC/Lurgi slagging, or B&W, K-T, and Texaco entrained flow gasifiers. This section summarizes results and recommendations regarding coal washing.

Background

The overall process of coal preparation can generally be divided into the following types of unit operations:

1. Comminution
2. Classification
3. Washing
4. Dewatering/drying

Comminution involves reduction in size from run of mine, typically 8" x 0", down to pulverized, typically smaller than 200 mesh. Classification involves separation of the coal particles into various size ranges. Washing generally involves treatment of the various size fractions with a liquid media to separate, preferentially, ash and pyritic sulfur from coal. Dewatering and drying involves removal of surface water from the prepared coal by mechanical or thermal means.

Run of mine coal passes through some of all of these operations before becoming the desired end product. The unit operations can be combined in various ways, with each operation using different types of equipment and technology, depending on the raw coal characteristics and degree of treatment required to produce a specified product. The objective of the overall coal preparation process, as related to a coal gasification plant, is to produce a material suitable as feed to the selected gasification process within economic boundaries of high Btu recovery, low ash content and low sulfur content.

With respect to the gasification processes being studied by Foster Wheeler for TVA, the gasifier coal feed must meet the following requirements:

Gasification Process	Gasifier Feed Properties Size Range	Moisture
Lurgi Dry Ash	2" x ¼"	5 wt. %
BGC/Lurgi Slagger	2" x ¼"	5 wt. %
Entrained Flow Gasifiers (B&W, K-T, Texaco)	-28 mesh/70%-200 mesh	2 wt. %/Slurry(Texaco)

In general, none of these gasification processes place limitations for process reasons on the ash or sulfur content of the coal except in extreme cases, although increased ash or sulfur content can have an adverse economic effect on the overall coal gasification operation. The requirements for gasifier feed dictate the choice and integration of coal preparation unit operations. A generalized diagram illustrating these operations is shown in Figure 1.

The unit operations of comminution, classification, and dewatering/drying have been standardized to a significant degree in current coal preparation technology, based on raw coal and product coal size and moisture requirements. The decision to include a coal washing step and the type of washing step to employ, however, is a matter requiring careful evaluation. In addition, the characteristics of the coal washing step can determine, to a large degree, the nature and extent of other coal preparation unit operations. For purposes of the present study, consideration was given only to coal washing processes using water and water-based media which are currently employed in washing operations for over 95% of the coal that is subjected to such treatment in the United States.

Coal Washing Technology

The following aspects of coal washing technology were examined in the study:

1. The various water and water-based media washing processes that are commercially available and proven.
2. A mild coal washing process incorporating the most adaptable washing process of those identified in item (1) above.
3. A deep coal washing process incorporating the most adaptable washing process of those identified in item (1) above.
4. A coal preparation process that does not include coal washing but which will prepare coal to the size and moisture content required.

The information developed in items 2, 3, and 4 was analyzed and intepreted to provide recommendations regarding inclusion of a coal washing step in TVA's coal gasification plant.

Coal Washing Processes - Water and Water-based Media

General

Coal is a heterogeneous material containing:

- organic combustible matter
- mineral non-combustible matter or impurities which can be broadly divided into
  - ash-forming material (clays, slimes)
  - sulfur containing material consisting of
    a. organic sulfur which is chemically bound to the coal and which is not subject to removal by physical coal cleaning methods.
    b. pyritic sulfur which exists as a separate compound in the heterogeneous coal particle and which can be removed by physical coal cleaning methods.

In order to upgrade Btu value and lower the ash and sulfur content of the raw coal, the washing process must selectively separate mineral matter and pyritic sulfur from organic combustible matter.

Fortunately, there are differences in the specific gravity and the surface wettability (for fine and ultra-fine coal) between the desired coal product and the mineral matter that is to be rejected. These differences are the operating basis of equipment used in the various water and water-based media washing processes.

Comparison of Washing Processes

The various water and water-based media washing processes that are commercially proven and available are listed in Tables 1 and 2. There is considerable published literature describing design and operating modes of these processes (see appended list of references) and these details will not be discussed in this study. Information shown in the Tables was prepared to provide a comparison of each of these washing processes. Based on this information, the following processes were considered most adaptable to the mild and deep coal washing processes considered later in this study:

mild coal washing - jig

deep coal washing - heavy media vessel and cyclone

Each of the above washing processes were then subdivided into two (2) classifications of gasifier designs:

dry ash/slagging (Lurgi/BGC) - coarse 2" x ¼" charge

entrained flow (B&W, K-T, Texaco) - fine mesh charge

It should be noted that although the Texaco design employs a slurry charge, its entrained flow regime is assumed to be similar to the B&W and K-T designs for this coal washing assessment.

Mild Coal Washing - Dry Ash and Slagging Gasifiers

A flow diagram for mild coal washing to produce washed 2" x ¼" product is shown in Figure 2. Coal is first dry screened to separate ¼" x 0 fines and then washed in a jig where clean coal is separated from refuse. The clean coal is dewatered and crushed and screened to produce 2" x ¼" product. Fines from dewatering, dry screening, and crushing are combined into a single ¼" x 0⁻ stream. Refuse from the jig is dewatered and sent to disposal.

Mild Coal Washing - Entrained Flow Gasifier

A flow diagram for mild coal washing to produce a -28 mesh/70%-200 mesh clean coal is shown in Figure 3. Coal is washed in a jig where 3" x 0 clean coal is separated from refuse. The clean coal is dewatered and the 3" x 1" fraction is crushed. Crushed coal is mixed with the 1" x ¼" fraction from dewatering and dewatered fines. The mixture is then pulverized and dried (if required) to produce the fine mesh product. Refuse is sent to disposal.

Deep Coal Washing - Dry Ash and Slagging Gasifier

A flow diagram for deep coal washing to produce a washed 2" x ¼" product is shown in Figure 4. Coal is first wet screened to produce 3" x 1½", 1½" x ¼", and ¼" x 0 fractions. The largest fraction is washed in a heavy media vessel. The intermediate size fraction is washed in a heavy media cyclone. Washed 3" x 1½" coal is crushed and screened and combined with washed 1½" x ¼" coal to produce the 2" x ¼" product. Fines from wet screening are dewatered and combined with fines from crushing. Refuse is dewatered and sent to disposal.

Deep Coal Washing - Entrained Flow Gasifier

A flow diagram for deep coal washing to produce a washed -28 mesh/70%-200 mesh product is shown in Figure 5. Coal is first wet screened to produce 3" x 1", 1" x ¼", and ¼" x 0 fractions. The largest fraction is washed in a heavy media vessel while the intermediate fraction is washed in a heavy media cyclone. The fines fraction is deslimed, thickened, dewatered, and sent to pulverization. Washed 3" x 1" coal is crushed, mixed with washed 1" x 28 mesh coal, and sent to pulverization. Crushed coal and fines are pulverized and dried (if required) to produce the fine mesh product. Refuse is dewatered and sent to disposal.

Coal Preparation Without Washing

Flowsheets for preparation of 2" x ¼" coal and fine mesh coal are shown in Figures 6 and 7. In the former case, coal is crushed to 2" x 0 and screened to separate out ¼" x 0 fines. In the latter case, coal is crushed and then pulverized to produce the fine mesh product for the entrained flow gasifier designs.

Assessment of Washing Processes

Estimated yields of prepared coal are shown in Table 3 for the mild, deep, and no-washing processes described above. Production of 2" x ¼" washed coal for dry ash or slagging gasifiers results in a yield of 63-65% on a Btu basis and 56-58% on a weight basis. The product contains about 5% ash and 3.3% sulfur, equivalent to 66% and 25% reduction in these components, respectively. In contrast, the no-washing case provides approximately 70% yield (Btu or weight basis) of 2" x ¼" product having the same ash and sulfur content as the as-received coal.

Production of fine mesh washed coal for the entrained flow gasifier results in a yield of 86-90% on a Btu basis and 77-80% on a weight basis. The product contains about 5% ash and 3.3% sulfur, equivalent to 66% and 25% reduction in these components. The no-washing case for the entrained flow gasifier provides essentially 100% yield of product having the same ash and sulfur content as the as-received coal.

In all of these washing cases, coal prepared for gasification has significantly lower ash content than the as-received coal. The use of low ash coal in gasification can result in appreciable savings, particularly in ash or slag handling facilities. The washing process, however, results in loss of a significant portion of the thermal value of the coal to washing plant refuse. This results in increased coal [redacted] for the gasification operation. In addition, savings generated by reduced ash or slag handling are likely to be offset by increased cost associated with handling and disposal of washing plant refuse.

Reductions estimated in sulfur content of coal as a result of washing are not large, 22-28%. This level of reduction is not expected to provide any significant savings in the gasification plant compared to the use of unwashed coal.

Recommendations

Based on the above analysis, it appears that washing of coal for TVA's coal gasification plant does not offer any significant economic advantages compared to the use of unwashed coal. In the absence of process requirements for washed coal - which are not evident at the present time - it is recommended that unwashed coal be used for gasification, after crushing, screening, and pulverizing and drying as required, to provide coal of appropriate size range and moisture content.

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6. Coal Age, pp. 66-83, January 1978; pp. 54-59, March 1979; pp. 48-52, August 1977; pp. 76-79, November 1977; pp. 152-153, January 1976; pp. 82-91, October 1978; pp. 72-76, November 1976.

7. Personal Communication with H. Davis, KVS, Danville, PA, January 31, 1980.

8. Buder, M. and K. Clifford, Cost Effects of Coal Cleaning on Utility Power Generation, presented at 2nd International Coal Utilization Conference, Houston, Texas, November 1979.

TABLE 1

COMPARATIVE ANALYSIS OF COAL WASHING EQUIPMENT FOR COARSE COAL

Equipment Characteristics	Jigs	Heavy Media Vessels	Tables	Hydrocyclones	Launders
% of Washing Plants Installed	Very high	High	Low	Low	Low
Relative Costs					
Capital	High	High	Low	High	Low
Operating	Low	High	Low	High	Low
Maintenance	Low	High	Low	High	Low
Feed Size Range	8" to 28m	8" to ½"	Below ½"	To 3/4" x 0	-
Separation					
Specific Gravity Range	Above 1.45	1.30 to 1.80	1.50 to 1.80	1.40 to 1.80	1.60 to 1.80
Sharpness	High	Very high	Low	High	Low
Cutpoint Control	Low	High	Low	Low	Low
Sensitivity to Variations	High	Low	High	High	High
Space	High	High	High	Low	High
Capacity	Up to 1000 TPH	Up to 900 TPH	25 TPH/Deck	75 TPH/Unit	-
Comments	High capacity per unit Largest % of plants	Media Recovery system required	Not recommended on coarse coal	Excellent for primary scalping for tables	Rarely used in new installations
Recommended Unit	X	X			

TABLE 2

COMPARATIVE ANALYSIS OF COAL WASHING EQUIPMENT FOR FINE COAL

Equipment Characteristics	Heavy Media Cyclones	Tables	Flotation	Hydrocyclones	Jigs	Launders
% of Washing Plants Installed	High	High	High	Low	Low	Low
Relative Costs						
Capital	High	Low	High	High	High	Low
Operating	High	Low	Low	High	Low	Low
Maintenance	High	Low	Low	High	Low	Low
Feed Size Range	2" x 28 mesh	3/8" x 0	28 mesh x 0	3/8" by 100 mesh	½" x 0	Not smaller than 100 mesh
Separation						
Specific Gravity Range	1.30 to 1.80	1.50 to 1.80	-	1.40 to 1.80	Above 1.60	1.60 to 1.80
Sharpness	Very high	High	Low	Low	High	Low
Cutpoint Control	High	Low	-	Low	Low	Low
Sensitivity to Variations	Low	High	Low	High	High	High
Space	Low	High	High	Low	High	High
Capacity	100 TPH/Unit	25 TPH/Deck	-	75 TPH/Unit	100 TPH/Unit	-
Comments	Media recovery system required	Very effective in pyrite removal	Reagents required Separation is not S.G. but surface wettability	Excellent for feed thickening prior to dewatering	Requires Feldspar bed	Rarely used
Recommended Unit	X					

TABLE 3

Estimated Yields for Mild, Deep, and No-washing Processes

3" x 0 Coal Feed, 27% less than ¼"

Gasification Process	Dry Ash or Slagger			Entrained Flow		
Washing Process	Mild	Deep	None	Mild	Deep	None
As-received Coal						
Weight, lbs, MF	100.0	100.0	100.0	100.0	100.0	100.0
% Ash	←	15.5	→	←	15.5	→
% S	←	4.5	→	←	4.5	→
HHV, M Btu/lb, MF	←	12.0	→	←	12.0	→
Prepared Coal						
Weight, lbs	55.9	57.4	70.0	76.6	80.3	100.0
% Ash	5.2	4.9	15.5	5.2	4.7	15.5
% S	3.5	3.2	4.5	3.5	3.2	4.5
% Moisture	5.0	5.0	5.0	2.0	2.0	2.0
HHV, M Btu/lb, MF	13.5	13.5	12.0	13.5	13.5	12.0
% Yield	←	2" x ¼"	→	< 28 Mesh/70%	< 200 Mesh	
Btu Basis	62.9	64.6	70.0	86.1	90.3	100.0
Weight Basis	55.9	57.4	70.0	76.6	80.3	100.0
% Reduction in						
Ash	66	68	0	66	68	0
Sulfur	23	28	0	23	28	0

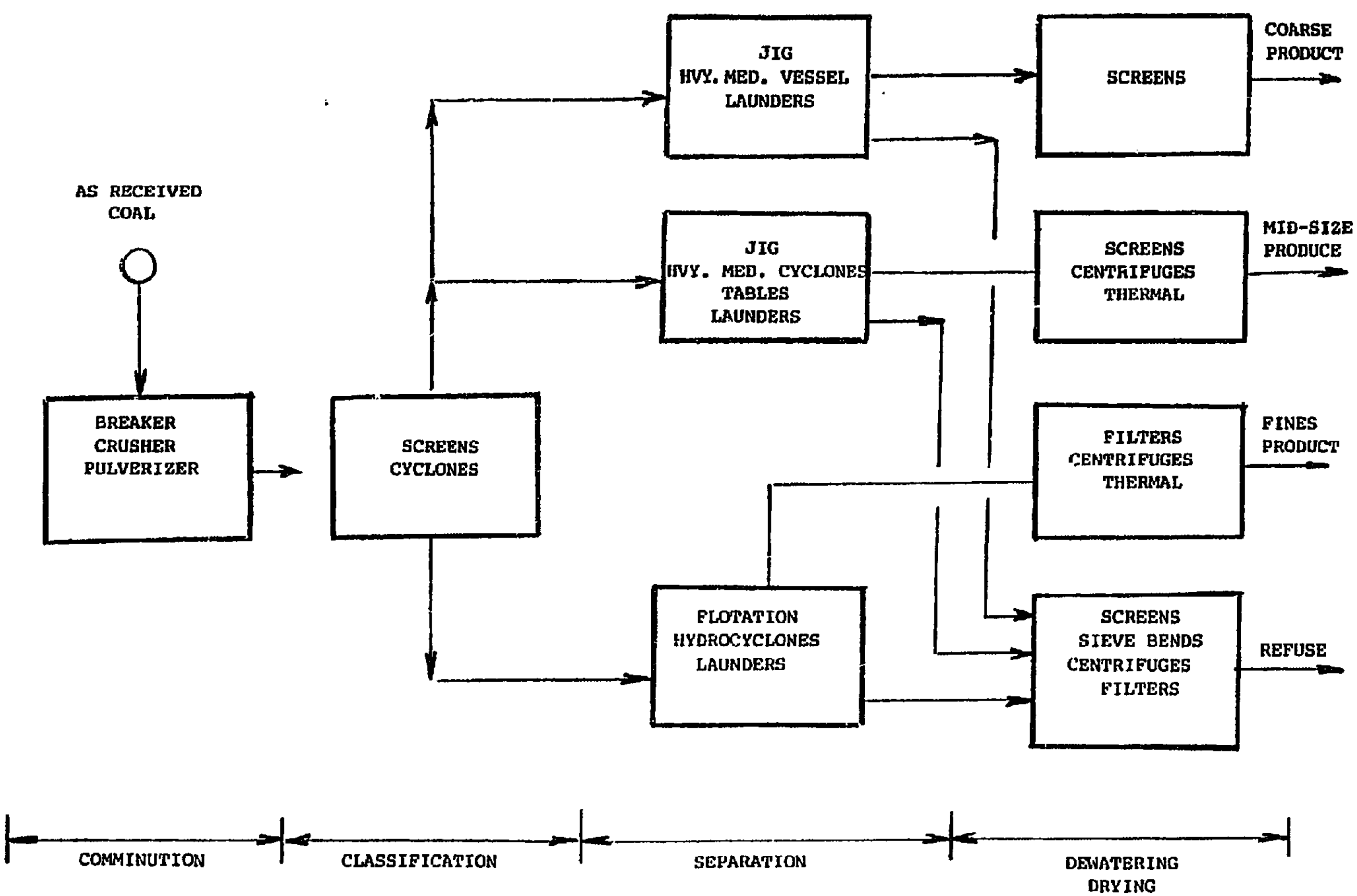


FIGURE 1

COAL WASHING UNIT OPERATIONS

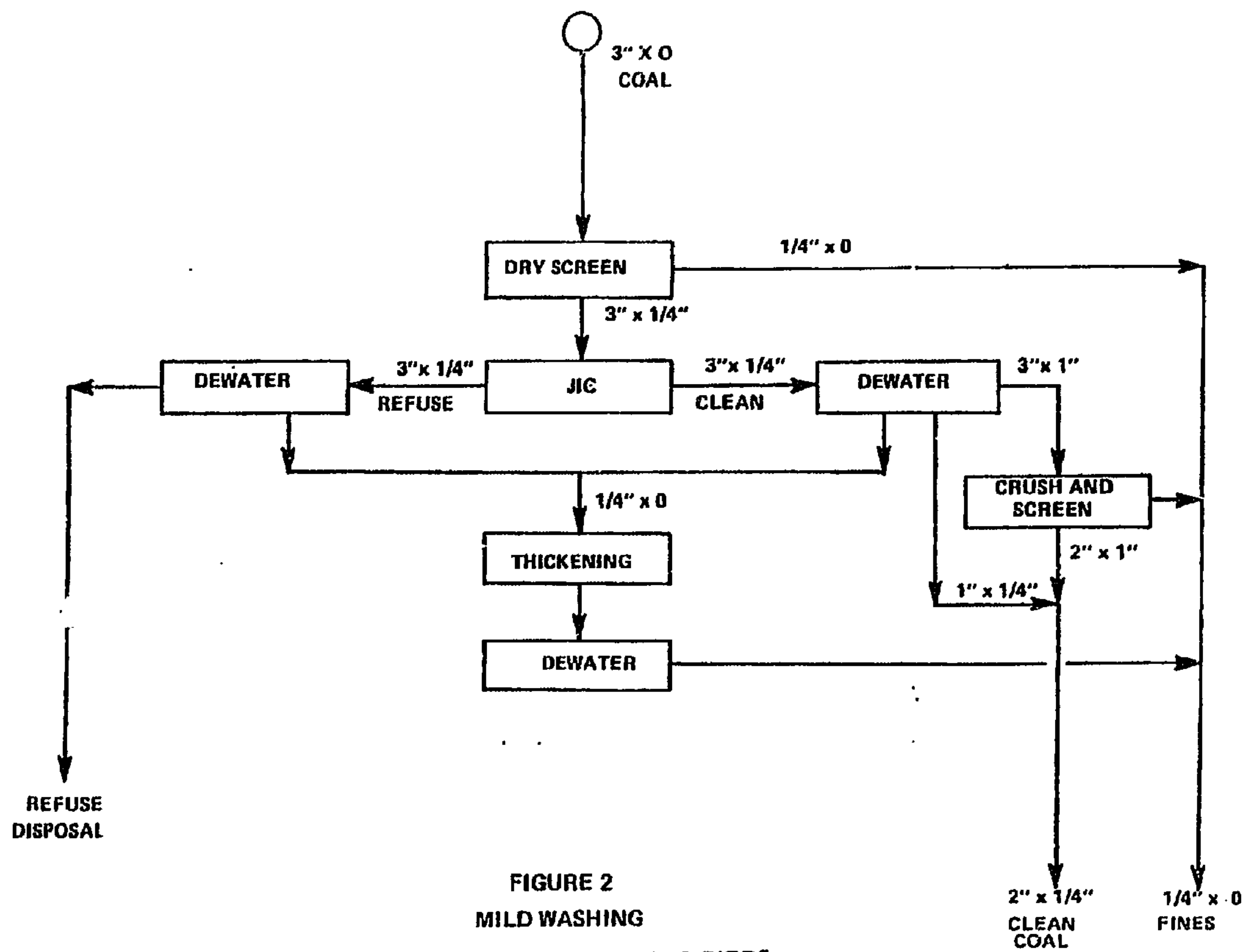


FIGURE 2
MILD WASHING
DRY ASH AND SLAGGING GASIFIERS

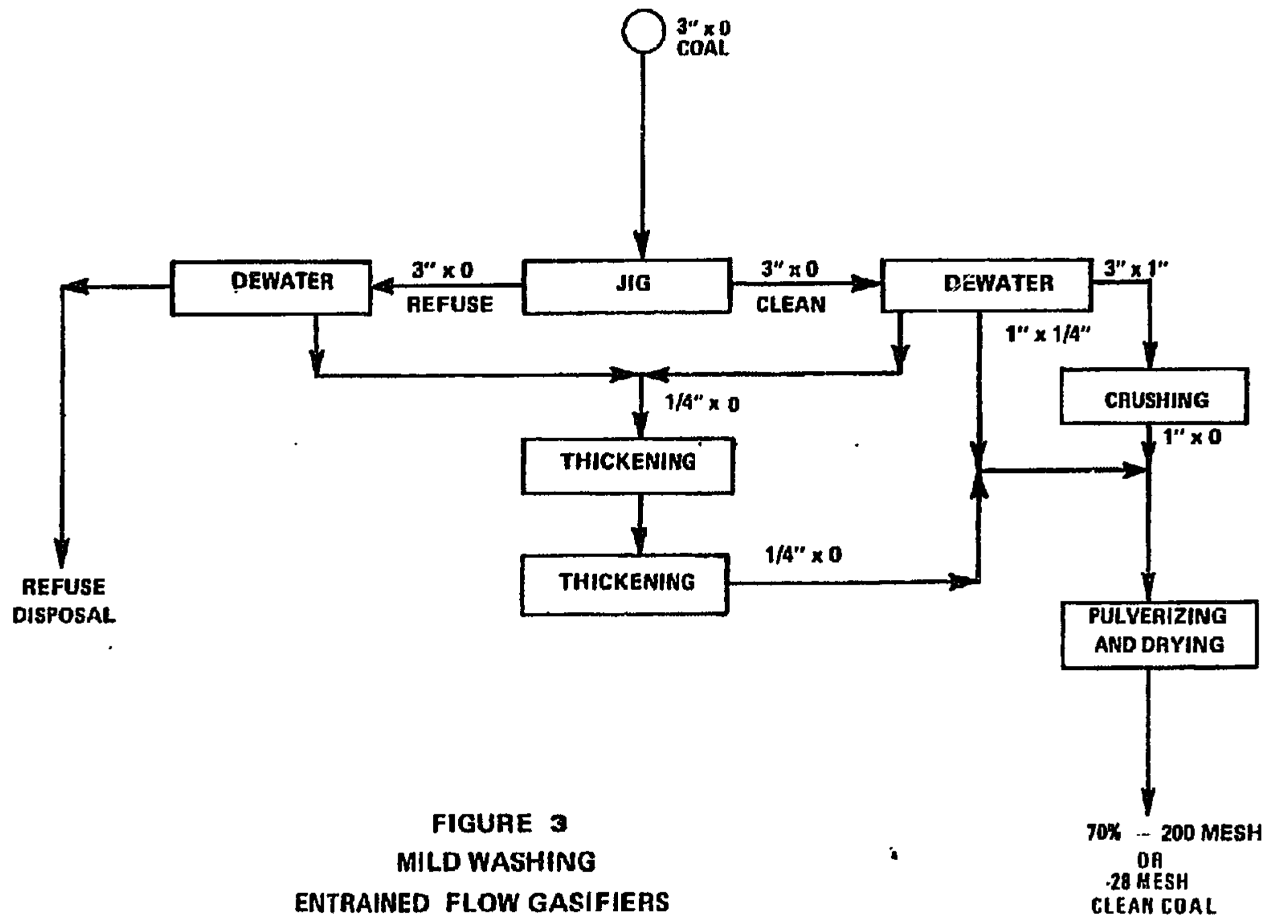


FIGURE 3
MILD WASHING
ENTRAINED FLOW GASIFIERS

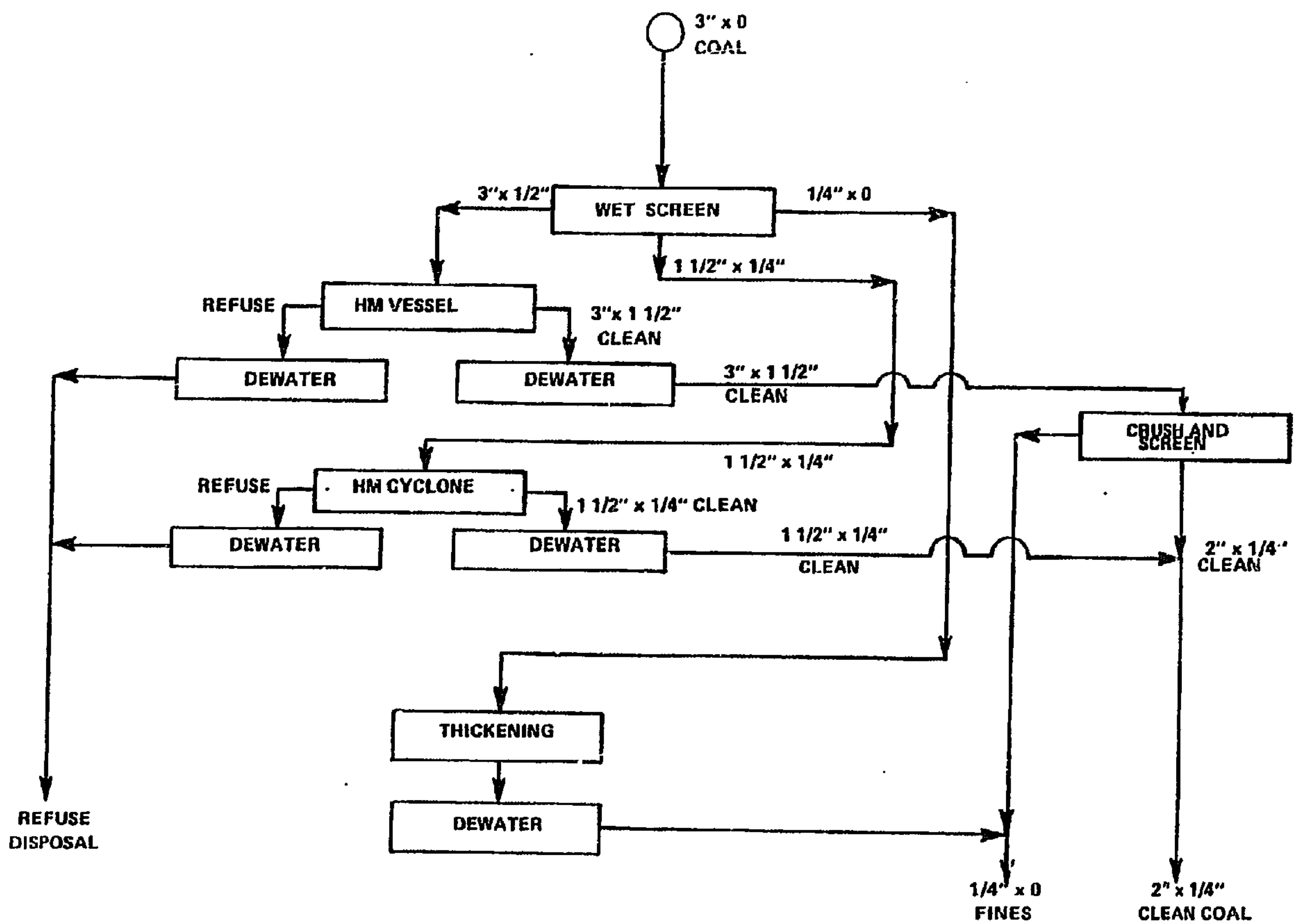


FIGURE 4
DEEP WASHING
DRY ASH AND SLAGGING GASIFIERS

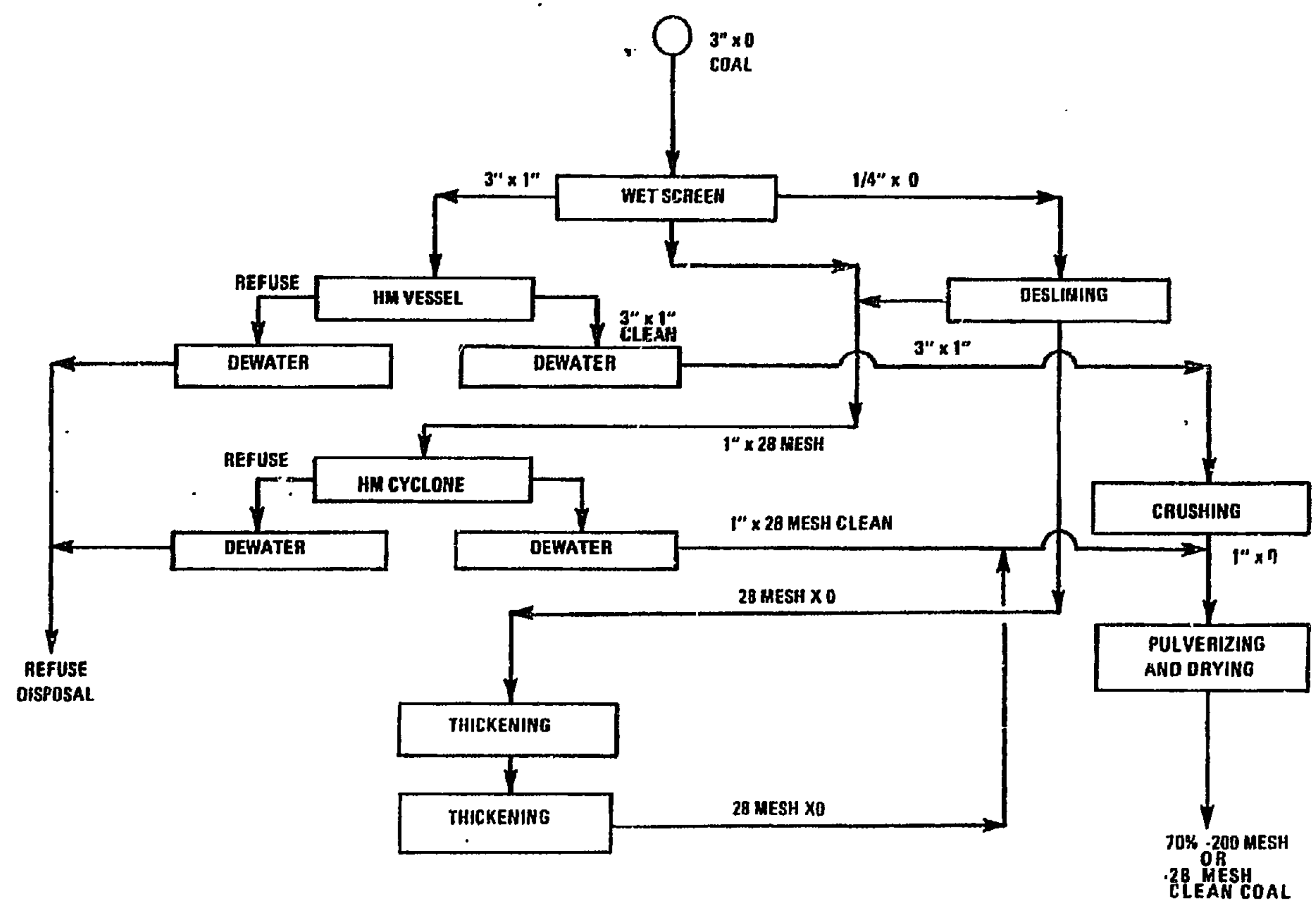


FIG. 5
DEEP WASHING
ENTRAINED FLOW GASIFIERS

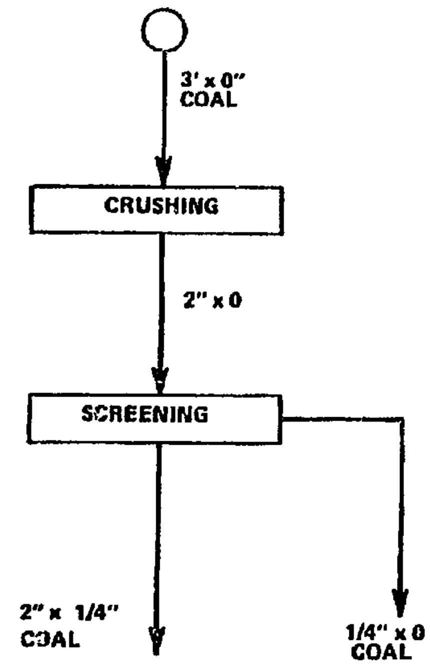


FIGURE 6
NO WASHING
DRY ASH AND SLAGGING GASIFIERS

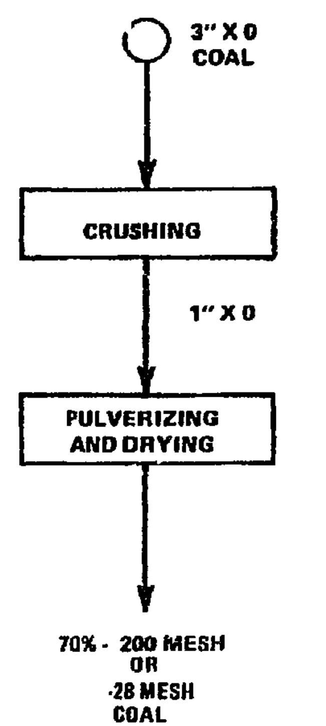


FIGURE 7
NO WASHING
ENTRAINED FLOW GASIFIERS

SECTION 3.4

SULFUR RECOVERY

ASSESSMENTS AND PROCESS SELECTION

SULFUR RECOVERY

Introduction

The objective of these subtasks is to identify the preferred form of recovered sulfur and the preferred process for sulfur recovery. These assessments were made in the context of 20,000 TPD coal gasification plants using Lurgi dry ash, BGC/Lurgi slagging, B&W entrained flow, Texaco entrained flow, and K-T entrained flow gasifiers. This section summarizes the results and recommendations of the assessments.

Background

The coal gasification plants being considered by TVA will process approximately 20,000 TPD of eastern U.S. coal such as Kentucky No. 9 or Illinois No. 6. These coals have high sulfur contents, ranging from about 3.5 to about 5.0%. The process of gasification, using either Lurgi dry ash, BCG/Lurgi slagging or B&W, Texaco, or K-T entrained flow gasifiers, converts 90% or more of the sulfur contained in the coal to volatile sulfur compounds. These compounds, primarily hydrogen sulfide and carbonyl sulfide, must be removed from the raw gas to meet safety, health, and environmental regulations. The volatile sulfur compounds must then be converted to a form which is marketable for industrial use or which can be disposed of in a practical and acceptable manner. Elemental sulfur and sulfuric acid are the two major forms of sulfur which are considered marketable in principle for large scale industrial use. Calcium sulfate is a form of sulfur which might be considered for disposal.

The chemistry of conversion of sulfur contained in volatile sulfur compounds resulting from coal gasification is essentially that of oxidation:

$$H_2S \xrightarrow{\frac{1}{2} O_2} S \xrightarrow{O_2} SO_2 \xrightarrow{\frac{1}{2} O_2} SO_3$$

	H_2S	S	SO_2	SO_3
Oxidation State	-2	0	+4	+6

Sulfur contained in hydrogen sulfide (or carbonyl sulfide) is in its lowest stable oxidation state, -2. Oxidation to the next higher oxidation state (zero) produces elemental sulfur. Further oxidation produces sulfur dioxide (+4), and finally sulfur trioxide (+6 oxidation state). Elemental sulfur is thus the first marketable form of sulfur produced in the oxidation sequence. Further oxidation leads to the other major marketable sulfur compound, sulfuric acid. The major disposable form, calcium sulfate, represents a completely oxidized form of sulfur.

Large quantities of hydrogen sulfide and other volatile sulfur compounds are produced in petroleum refining operations, particularly where high sulfur crude oils are processed in hydrotreating operations. Petroleum refinery practice for many years has been to convery hydrogen sulfide and similar compounds to elemental sulfur for sale to others. This practice thus carries the oxidative conversion of sulfur compounds only to the extent required to produce the first marketable product, elemental sulfur.

Technology for Sulfur Recovery from Hydrogen Sulfide Streams

Recovery as Elemental Sulfur

Claus Process

The predominant commercial method of converting hydrogen sulfide to elemental sulfur is the Claus process. The process, as originally developed about the year 1900, involved oxidation with air in the presence of a bauxite or iron ore catalyst in a reaction chamber. In the early 1940's, a modification was generally adopted in which one-third of the hydrogen sulfide was burned with air to sulfur dioxide in a waste heat boiler. The sulfur dioxide was then reacted with the remaining two-thirds of the hydrogen sulfide in the presence of a bauxite catalyst.

There are four major variations to the Claus process available for use depending on the concentration of hydrogen sulfide in the feed gas to the process. These variations and typical ranges of sulfur recoveries are shown in Figures 3.4-1 to 3.4-4. For gas streams containing high concentrations of hydrogen sulfide (above about 50% H_2S), the total stream is fed to the burner in a furnace or boiler as shown for the straight through variation. Air is fed to the burner to oxidize most hydrogen sulfide to elemental sulfur. The furnace effluent is cooled to condense sulfur. The remaining gas is preheated and fed to a catalytic reactor where hydrogen sulfide reacts with sulfur dioxide to form elemental sulfur and water. The reacted gas is cooled to condense sulfur. Additional reactors and condensers can be included to increase sulfur recovery. Typically, sulfur recoveries up to 95% are obtained with the straight through Claus process.

For feed gases in which the hydrogen sulfide concentration is between about 15 to 50%, the split flow process variation is used. In this variation, up to two-thirds of the gas is bypassed around the furnace and the remainder is burned in the furnace to sulfur dioxide. The furnace effluent is blended with bypass gas and the mixture reacted in one or more catalytic reactors to form elemental sulfur. Preheating of the original feed gas would permit this method to be used where hydrogen sulfide concentration is somewhat less than 15%. Typical sulfur recoveries range from 90 to 93% for split flow operation.

For gases where the hydrogen sulfide concentration is too low to provide stable combustion in the split flow furnace, sulfur recycle or direct oxidation variations can be used. In the former case, product sulfur is burned in the furnace to produce sulfur dioxide. This gas is mixed with feed gas and then reacted in catalytic reactors. In the direct oxidation variation, feed gas is preheated, mixed with air and reacted in catalytic reactors. Typically, suflur recoveries range from 75 to about 90% for these process variations. Technically, there is no lower limit of hydrogen sulfide concentration in feed gas for the sulfur recycle or direct oxidation process variations. Economically, however, the lower limit is usually about 10%.

Special design considerations are required for Claus plant handling feed gases containing appreciable amounts of hydrocarbons or ammonia. In the case of hydrocarbons, special attention is paid to design of the acid gas recovery process. Ammonia can be handled by proper selection of operating conditions in the furnace and downstream equipment.

Tail Gas Cleanup

Sulfur recoveries up to about 95% can be obtained with Claus sulfur recovery plants depending on the process variation and number of catalytic reactors used. In 1978, however, the Federal Environmental Protection Agency established the following limitations on emissions from new, modified and reconstructed petroleum refinery Claus sulfur recovery plants:

In gases discharged into the atmosphere -

SO_2	250 ppmv
H_2S	10 "
Total of H_2S, COS and CS_2 as SO_2	300 "

calculated at zero percent oxygen on a dry basis.

These standards apply to (1) any Claus sulfur recovery plant with sulfur production of more than twenty long tons per day which is associated with a small petroleum refinery, and (2) any size Claus sulfur recovery plant associated with a large petroleum refinery.

If these standards are applied to Claus plants in coal gasification plants, it is necessary to achieve a sulfur recovery of about 99.9%. This level of recovery can be obtained only by the addition of a tail gas treatment process to the Claus plant. Two treatment processes that have been developed for this purpose are the Beavon Sulfur Removal Process and the SCOT Process. The Beavon Process involves contacting the tail gas and a reducing gas with a hydrogenation-hydrolysis catalyst to convert contained sulfur compounds essentially completely to hydrogen sulfide. The reacted gas is fed to a Stretford process unit where hydrogen sulfide is oxidized to elemental sulfur in the presence of an aqueous catalyst. Sulfur is removed from the aqueous phase by froth flotation. The solid is then filtered or centrifuged and melted.

The SCOT Process involves a reduction stage to convert sulfur compounds to hydrogen sulfide and an absorption stage where hydrogen sulfide is scrubbed from the gas and returned to the Claus plant. Both the Beavon and SCOT processes are capable of cleaning Claus plant tail gas to levels required by the above emission standards.

In the SCOT process, the reduction step converts essentially all sulfur containing compounds to H_2S which is then absorbed in an alkanolamine solution. The non-absorbed H_2S, approximately 200-300 ppmv, appears in the final tail gas vent and must be incinerated to meet the H_2S emission standards. Furthermore, a fraction of the CO_2 in the tail gas is co-absorbed with the H_2S and is ultimately recycled to the Claus plant. The recycled gas has little effect on Claus plant operations as long as the CO_2 content is low. As the CO_2 content of the SCOT feed gas increases, the amount of recycled CO_2 becomes a significant diluent. Therefore, the SCOT process may not be a good selection for direct treating of the tail gas from a Claus plant which processes a feed gas having a high CO_2 content.

In view of the potential disadvantages associated with the SCOT process, i.e.

- need to incinerate final vent gas, and
- limitations on CO_2 content in feed gas

Foster Wheeler recommends that the Beavon process be incorporated in the conceptual design studies.

Recovery as Sulfuric Acid

Contact Process

Production of sulfuric acid involves, firstly, the production of sulfur dioxide from sulfur-containing feedstock and, secondly, catalytic oxidation of sulfur dioxide to sulfur trioxide which is reacted with water to form acid. Production of sulfur dioxide is carried out by oxidation of elemental sulfur, metallic sulfides, hydrogen sulfide, or sulfuric acid sludges. Elemental sulfur is the predominant raw material for sulfuric acid plants, accounting for about 90% of the total domestic acid production. Copper and zinc smelter gases and pyrites roaster gases comprise essentially all of the remaining sulfuric acid feedstocks. Only about 1% of total acid production uses hydrogen sulfide as a feedstock for sulfur dioxide production.

The Chamber Process was used extensively until the late 1920's to produce acid from sulfur dioxide. In this process, nitric oxide was used in effect as a catalyst for oxidation. The process, however, had low productivity and could not produce concentrated acid. Currently, the Chamber Process has been essentially completely displaced by the Contact Process.

A simplified flowsheet for a Double Absorption Contact Sulfuric Acid Process is shown in Figure 3.4-5. Molten sulfur is filtered and then burned with air that has been dried in a sulfuric acid drying tower. The burner effluent contains approximately 12% sulfur dioxide which is then diluted with dry air to a concentration of about 9%. The gas flows through vanadium pentoxide catalyst arranged in several beds in a converter vessel. When the conversion level reaches about 88%, the gas is withdrawn from the converter, cooled, and sent to an absorber where sulfur trioxide is absorbed in 98% acid. Gas from the absorber is returned to the converter vessel where it contacts additional catalyst beds to attain the final sulfur dioxide conversion.

Gas from the final catalyst bed is cooled and sent to a second absorber where sulfur trioxide is absorbed in strong acid. The final acid product is made by mixing acid from the drying tower and the absorbers.

Sulfur dioxide and sulfur trioxide mist contained in the tail gas of sulfuric acid plants can be a significant source of atmospheric sulfur emissions. The Federal Environmental Protection Agency has established for new acid plants a limitation of 4 lbs. of SO_2 emission per short ton of sulfuric acid produced. This criterion corresponds to an effective conversion of sulfur dioxide of about 99.7%. The double absorption variation of the contact process is the only feasible means of approaching this level of conversion because of equilibrium limitations on conversion in the absence of intermediate absorption of sulfur trioxide. Even at high conversion, efficient devices for removal of acid mist from the tail gas are required.

The use of hydrogen sulfide as a feedstock for sulfuric acid production requires special consideration if the concentration of hydrogen sulfide in the feed gas is lower than about 80%. With dilute hydrogen sulfide streams, heat generated in combustion of the sulfide may not be sufficient to heat the converter feed gas above the catalyst ignition temperature. This problem can be overcome at the expense of adding elemental sulfur to the burner. A further disadvantage of dilute hydrogen sulfide streams is the increased gas flows resulting from diluents in the feed gas which increase the size of most of the processing equipment.

Marketability Considerations

This assessment of sulfur recovery for TVA's coal gasification plant did not include a detailed market survey. The following items, however, relate to the question of marketability of elemental sulfur and sulfuric acid.

1. Elemental sulfur is a major industrial chemical and raw material. It is produced either from naturally occurring deposits (Frasch sulfur) or from various process gases containing sulfur compounds. Currently, Frasch sulfur accounts for about 60% of the total elemental sulfur produced in the United States. Essentially all (over 90%) of the sulfur recovered from sour gas streams, petroleum refinery streams, and other process streams containing sulfur compounds, is recovered as elemental sulfur. Most of the remainder is recovered as sulfuric acid, primarily in situations where SO_2-containing gases are available as produced

from roasting or smelting of non-ferrous sulfide ores and pyrites. Only about 0.5% of the total sulfur production is made as acid from hydrogen sulfide-containing streams.

2. Elemental sulfur can be shipped long distances in molten form in bulk carriers including truck, rail, barge and ship. Bulk shipments of molten sulfur began in the 1940's and are widely used at the present time. The major advantages of molten sulfur shipment are reduced handling costs, reduced contamination of sulfur, and elimination of remelting operations. Precautions are required, however, in storing, handling and shipping molten sulfur. Elemental sulfur can also be formed into prills if a solid form is required that does not have the problem of remelting massive blocks of sulfur. Solid sulfur is classified as an ORM-c material in the Department of Transportation (Materials Transportation Bureau) Hazardous Materials Table, which is essentially a non-hazardous material category.
3. Sulfuric acid is one of the largest volume industrial chemicals produced in the United States with widespread use in many industrial operations, particularly fertilizer production. Over 85% of the sulfuric acid produced in the United States is made from elemental sulfur. Production of acid from elemental sulfur in the eastern part of the United States is shown in Figure 3.4-6. Production in the State of Florida is about fifteen million tons per year.
4. Commercial grades of sulfuric acid and oleum are shipped in steel drums (returnable), tank trucks, tank cars, and tank barges. Sulfuric acid is classified as a corrosive material in the Department of Transportation Hazardous

Materials Table. Shipments must be made in accordance with DOT specifications and require a DOT label. Dedicated fleets of tank trucks, tank cars, and tank barges are often used. In general, sulfuric acid is more expensive to ship than elemental sulfur on the basis of contained sulfur since the acid contains only about 33% sulfur by weight. Industry practice in general has been to ship elemental sulfur to the locality of end use and to produce sulfuric acid at that locality.

Economic Comparison

A preliminary economic comparison was made of two alternatives for sulfur recovery in TVA's coal gasification - elemental sulfur or sulfuric acid. The following basis was used in making the comparison:

1. The 20,000 TPD coal gasification plant produces an acid gas stream containing about 20% hydrogen sulfide at a rate equivalent of 900 LTPD of recoverable elemental sulfur.
2. Elemental sulfur is recovered from the acid gas stream in a Claus sulfur recovery plant equipped with a tail gas cleanup unit (Case 1). The plant consists of five independent trains, four operating and one spare, serving the four modules of the coal gasification plant.
3. Sulfuric acid is produced from the acid gas stream in a contact sulfuric acid plant (Case 2). The plant consists of five independent trains, four operating and one spare, serving the four modules of the coal gasification plant.

Economic calculations for Case 1 and Case 2 are summarized in Tables 3.4-1, 3.4-2, and 3.4-3. Capital investments for the plants were estimated from Foster Wheeler plant capacity-cost correlations for sulfur recovery plants and licensor information on sulfuric acid plants. Offsites for each battery limit plant were taken as 30% of the battery limit investments. It was assumed that the total fixed investment for each plant was entirely debt capital. Insurance, local taxes, and federal taxes were omitted from the calculations. Shipping costs of 4¢ per ton mile for a distance of 300 miles were included in the calculations.

The calculated return on total fixed investment is 6.3% for elemental sulfur and 24.7% for sulfuric acid assuming $70 and $60 per ton selling price for elemental sulfur and sulfuric acid respectively. If acid could be sold locally, production of sulfuric acid would be economically favored over production of elemental sulfur. If long distance shipping to customers is required, as is likely for relatively large production from TVA's coal gasification plant, production of elemental sulfur would be economically favored.

Elemental Sulfur Form

Elemental sulfur is commonly recovered in its molten form. This requires heated storage and transportation facilities as well as precautions against the release of noxious and potentially explosive fumes. For these reasons, TVA has expressed preference for producing sulfur byproduct in the form of solid prills. At TVA's request, Foster Wheeler obtained evaluation information on two commercial sulfur prilling processes, i.e.:

- Chemsource Sulfur Prilling Process
- Ciech-Intcan Sulfur Air Prilling Process

Basically, these two processes differ in the cooling medium used within the sulfur prilling tower. Chemsource employs a system in which molten sulfur is sprayed into a pool of water where the prills are cooled and solidified. Ciech-Intcan, on the other hand, uses a stream of cool air to form the sulfur prills.

Typical characteristics of the two processes are summarized below:

Product Properties	Chemsource	Ciech-Intcan
Prill shape	Spherical	Spherical
Particle Size, 90%	2-6 mm	1.5 - 5 mm
Angle of Repose	35°	30°
Bulk Density	73 lb/CF	80 lb/CF
Moisture	$<$2%	$<$0.02%

Operating Requirements		
Electric power, KWH/Ton	3.1	4.5 - 7.0
Steam @ 50 per, lb/Ton	15	15
Process water, lb/Ton	210	None
Operators, Men/shift	1	2

Foster Wheeler recommends that the Chemsource process be incorporated into the conceptual design studies because of the following design features:

- Dust pollution is eliminated in forming and also in subsequent handling of sulfur product. Moreoever, the nondusting characteristic is obtained over the range of product type.
- The basic product is closely sized hard surfaced spherical prill as required for industrial use. At the other end of the scale, the Chemsource Process can make a softer, more porous, irregularly shaped product used for agricultural applications. Size also can be varied within limits.
- Operation of the Chemsource process is simple and requires minimal operator attention. The unit is fully instrumented to permit operation from the control panel with the part-time services of the operator.
- Normal precautions on startup and shutdown ensures reliable operation. If the unit has been properly drained on shutdown and the reactor plate and feed system heated on restarting, specification prills will be produced in a matter of minutes, with no appreciable quantity of oversized product collecting on the grizzly of the reactor. Likewise, only a negligible amount of fines will be made

on startup, provided the above precautions have been taken and the feed liquid sulfur is of the quality specified.

- The nondusting character of the prills affords the possibility of economic outdoor storage. While storage and handling facilities of the sulfur product are not included in the battery limits of the unit, they may be included as required. Product with two percent moisture or less is easily handled by standard bulk handling equipment. Further the 1 to 2 % water as surface moisture on the prill contributes to the prevention of dust formation during subsequent handling and shipping.
- The safety control effect of moisture on particulate sulfur is widely recognized. While industrial users of sulfur prefer minimal moisture, very often producers and shippers add water to their particulate sulfur product to control the dust hazard. However, where required, facilities to obtain dry prills may be included.
- Waste waters may be considered as makeup to the cooling tower of the unit. While fresh water is specified, the level and character of the dissolved solids may be compatible with the intended application of the sulfur product. Any dissolved solids would ultimately coat the surface of the prill since part of the makeup water reports as surface moisture which leaves a trace residue on evaporation.

Stack Gas Cleanup

Another aspect of sulfur recovery in TVA's coal gasification plant is that of boiler stack gas cleanup. Production of steam for process use of electric power production will be required in the coal gasification plant. The extent of steam production will depend on the coal gasification process used and decisions made concerning steam versus electric drive of major compressors and pumps. Coal is the obvious choice as fuel for plant boilers since large quantities of coal will be supplied to the plant. In addition, certain fractions of the coal, such as fines, may not be suitable for gasification but could be used as boiler fuel.

Combustion of high sulfur coal in steam boilers produces a significant amount of sulfur dioxide in the boiler flue gas. In principle, this sulfur dioxide could be recovered from the flue gas and recycled to the Sulfur Recovery Plant to produce additional elemental sulfur or sulfuric acid. An alternative is to remove sulfur dioxide from the flue gas as calcium sulfite or calcium sulfate for disposal. Either method provides significant reduction in sulfur dioxide emission from the boilers.

The quantity of coal burned in boilers in the coal gasification plant will depend upon the type of coal gasification process used. In the case of Lurgi dry ash gasifiers, coal used as boiler fuel could range up to 5000 tons per day because of large process steam requirements. Combustion of the high sulfur design coal at this rate would produce the equivalent of about 250 tons per day of sulfur as sulfur dioxide. Capture of this quantity of sulfur dioxide as calcium sulfite or sulfate would result in production of up to 2000 tons per day of solids for disposal. Since this is a relatively large quantity, up to 50% of the coal ash generated in the plant, a process capable of recovering sulfur dioxide from flue gas, such as the Wellman-Lord process, could be considered. For smaller quantities of coal burned in boilers, fluidized bed boilers utilizing limestone to absorb sulfur dioxide could be used since the solids produced for disposal could be dry and would not be excessive in quantity. Wet scrubbing of boiler flue gas produces large quantities of wet sludge which requires large settling/disposal areas.

A. Fluidized Bed Boiler

Technology

The fluidized bed boiler is a fuel combustion-steam generation system in which fuel is burned in a fluidized bed of solids and steam is generated in boiler tubes immersed in the fluidized bed. High rates of heat transfer to the boiler tubes and capture of sulfur oxides released during combustion are two of the many advantages of the fluid bed boiler.

Particulate solids are fluidized when gas flows upward through a mass of the solids at velocities up to about 15 feet per second. Under these conditions, the solids behave, in many respects, as a highly agitated liquid. When a fuel such as coal is introduced into a hot mass of solids fluidized by air, combustion occurs rapidly. Heat generated by combustion is absorbed by boiler tubes immersed in the fluidized bed, producing high pressure steam. Rapid agitation of the particles increases the heat transfer coefficient to the submerged tubes by a factor of about 10 compared to convective head transfer. This provides an important cost savings compared to conventional boilers.

Present environmental regulations require some form of control of sulfur oxides emissions when high sulfur fuels are burned in boilers. This control can be accomplished in fluidized bed boilers by the use of crushed limestone as the fluidized bed material. The temperature of the solids is controlled at about 1550-1600°F which is optimum for sulfur oxide capture. Sulfur oxides react with the calcined limestone to form calcium sulfate. Spent limestone is removed from the boiler for disposal in dry form. This is an important cost and operating advantage over wet scrubbing systems.

Another important advantage of the fluidized bed boiler compared to conventional boilers is the capability of burning ash-containing fuels at temperatures below the ash melting point. Problems of ash deposition and corrosion that occur in conventional boilers are avoided.

In addition to capture of sulfur oxides, fluidized bed boilers produce lower NO_x emissions than suspension or stoker fired boilers firing the same fuel. This is due to the relatively low combustion temperature used.

Based on extensive pilot plant testing and engineering work, Foster Wheeler has developed designs of industrial fluidized bed boilers, typically as shown in Figure 3.4-7. One feature of this design is the concept of individual cells. Each cell of the steam generator can be operated independently, allowing a form of load control beyond the range set by limits of fluidization velocity. Separate operation of cells also simplifies start-up procedures involving superheat and provides excellent superheater temperature control over the load range.

Application

The U.S. Department of Energy and the Electric Power Research Institute are both sponsoring research and demonstration projects concerning fluidized bed combustion. Foster Wheeler has been extensively involved in the development of fluidized bed combustion technology as indicated by the following summary of projects.

Pilot Plant

A fluidized bed combustion pilot plant was installed and operated at the DOE pilot plant facility at Alexandria, Virginia. This facility provided basic test data on combustion and heat transfer.

Rivesville Fluidized Bed Boiler

Foster Wheeler designed and installed a multicell fluidized bed boiler at Rivesville, West Virginia. This facility, designed for generation of 300,000 lb/hr of superheated steam, provided additional test data and operating experience.

Georgetown Fluidized Bed Boiler

Foster Wheeler designed and installed a natural circulation fluidized bed boiler at Georgetown University in Washington, D.C. This unit is designed to generate 100,000 lb/hr of saturated steam at pressures between 275 and 625 psig. The boiler contains two independently operated fired fluidized beds each 5 ft, 6 inches deep by 19 feet, 4 inches wide with a plan area of 106 ft^2. At full load conditions, the fluidized beds operate at a temperature of 1600°F with a fluidizing velocity of about 8 feet per second. Approximately 50% of the heat transfer occurs within the fluidized bed. Flue gas from the fluidized beds is cooled by a conventional boiler bank in which the cold end tubes act as downcomers and the remaining tubes are steam generating risers. The two fluidized bed arrangement provides a 4 to 1 turndown.

During operation, approximately 95% of the fluidized bed material consists of calcined and reacted limestone; the remaining 5% is fuel. Coal sized to minus 1.25 inches is injected into each bed with overbed spreader feeders of the same type used in spreader stoker boilers. These feeders do not require dry coal. Limestone is fed by gravity at the surface of each fluidized bed.

The fluidized bed boiler at Georgetown University has been in operation since July of 1979 and as has accumulated over 2000 hours on line. The unit has operated successfully with both one and two fluidized beds in service, generating up to 80,000 lb/hr of steam. Full load capability of a single fluidized bed has been shown to exceed design rates. Emissions of sulfur dioxide and nitrogen oxides from the Georgetown boiler were below Federal and District of Columbia standards when measured in December 1979 while fire 2.3% sulfur coal.

Fluidized bed boilers provide an advantageous new method of generating steam in an environmentally acceptable manner. These boilers can accept a wide range of fuels while simultaneously controlling emissions to levels within standards. Although commercial application of fluidized bed boilers is relatively new, industrial boilers having capacities up to about 600,000 lb/hr of steam can be obtained on a commercial basis.

Form No. 130-171

Recommendations

Consideration was given to various aspects of technology, marketability, and economics of sulfur recovery in TVA's proposed coal gasification plant. The following recommendations are made concerning Foster Wheeler's conceptual design of the coal gasification plants using Lurgi dry ash, BGC/Lurgi slagging, Texaco, Koppers Totzek, and B&W entrained flow gasifiers:

1. Sulfur should be recovered from acid gas streams in the form of elemental sulfur, preferably as solid prills.
2. Sulfur recovery should be carried out in a Claus plant equipped with a Beavon tail gas cleanup unit.
3. Sulfur dioxide should be recovered from boiler flue gases using a Wellman-Lord stack gas scrubbing process when the quantity of coal fired in boilers is large. For small quantitites of coal fired in boilers, fluidized bed boilers should be used.

PRODUCTION COST AND SELLING PRICE ESTIMATE TABLE 3.4-1

PLANT CLAUS PLANT + TAIL GAS CLEANUP			CASE 1	
CAPACITY 5 @ 225 LTPD = 1125	TYPE INVESTMENT ESTIMATE		CURVE	
LOCATION TVA REGION	INVESTMENT - B.L.		60.0	MM$
OPERATING DAYS/YEAR 330	INVESTMENT - OFFSITES @ 30%		18.0	
	INTEREST DURING CONSTR.@ 11%		8.6	
ELEMENTAL SULFUR PRODUCED	TOTAL FIXED INVESTMENT		86.6	
PER YEAR = 297,000 LT	WORKING CAPITAL			
	EQUITY			
	LONG TERM DEBT (INITIAL)		86.6	MM$

COST ITEMS	UNIT	UNIT PRICE	UNIT CONSUMPTION	COST PER UNIT PRODUCT
RAW MATERIALS				
ACID GAS STREAM				NO COST
TOTAL RAW MATERIALS				0.00
UTILITIES				
SEE ATTACHED PAGE				
TOTAL UTILITIES 0.84 MM$ CREDIT/YR				(2.83)
DIRECT OPERATING COST				
DIRECT OPERATING LABOR 2 OPERATORS/SHIFT @ $17,000/YR				0.50
OPERATING SUPERVISION @ 15% OF OPERATING LABOR				0.08
SUBTOTAL				0.58
MAINTENANCE 3% OF B.L. INVEST. + 1% OF OFFSITES INVEST.				6.56
TOTAL DIRECT OPERATING COST				7.14
INDIRECT OPERATING COST				
PLANT OVERHEAD @ 60% OF TOTAL DIRECT OPER. COST				4.65
DEPRECIATION @ 5% OF TOTAL FIXED INVEST.				14.58
INTEREST ON LONG TERM DEBT @ 11%				16.03
TOTAL INDIRECT OPERATING COST				35.26
TOTAL PRODUCTION COST				39.57
CREDITS: SHIPPING COST FOR 300 MILES @ 4¢/TON-MILE				12.00
NET PRODUCTION COST + SHIPPING				51.57
RETURN ON TOTAL FIXED INVESTMENT 6.3%				18.43
ESTIMATED SELLING PRICE				70.0

PRODUCTION COST AND SELLING PRICE ESTIMATE TABLE 3.4-2

PLANT SULFURIC ACID PLANT				CASE 2
CAPACITY 5 @ 750 TPD H_2SO_4 = 3750	TYPE INVESTMENT ESTIMATE		CURVE	
LOCATION TVA REGION	INVESTMENT - B.L.		85.0	MM$
OPERATING DAYS/YEAR 330	INVESTMENT - OFFSITES @ 30%		25.5	
	INTEREST DURING CONSTR.@11%		12.2	
SULFURIC ACID PRODUCED	TOTAL FIXED INVESTMENT		122.7	
PER YEAR = 990,000	WORKING CAPITAL		-	
	EQUITY			
	LONG TERM DEBT (INITIAL)		122.7	MM$

COST ITEMS	UNIT	UNIT PRICE	UNIT CONSUMPTION	COST PER UNIT PRODUCT
RAW MATERIALS				
ACID GAS STREAM				NO COST
TOTAL RAW MATERIALS				0.00
UTILITIES				
SEE ATTACHED PAGE				
TOTAL UTILITIES 1.20 MM$ CREDIT/YR				(1.21)
DIRECT OPERATING COST				
DIRECT OPERATING LABOR 7 OPERATORS/SHIFT @ $17,000/YR				0.52
OPERATING SUPERVISION @ 15% OF OPERATING LABOR				0.08
SUBTOTAL				0.60
MAINTENANCE 3 % OF B.L. INVEST. + 1 % OF OFFSITES INVEST.				2.79
TOTAL DIRECT OPERATING COST				3.39
INDIRECT OPERATING COST				
PLANT OVERHEAD @ 60% OF TOTAL DIRECT OPER. COST				2.18
DEPRECIATION @ 5 % OF TOTAL FIXED INVEST.				6.20
INTEREST ON LONG TERM DEBT @ 11 %				6.82
TOTAL INDIRECT OPERATING COST				15.20
TOTAL PRODUCTION COST				17.38
CREDITS: SHIPPING COST FOR 300 MILES @ 4¢ /TON-MILE				12.00
NET PRODUCTION COST + SHIPPING				29.38
RETURN ON TOTAL FIXED INVESTMENT 24.7%				30.62
ESTIMATED SELLING PRICE				60.00

TABLE 3.4-3

ECONOMIC CALCULATIONS

UTILITY COSTS

Item	Unit	Unit Price	Consumption	MM $/YR
	Case 1			
Steam				
400 psig suphtd	M lb	$2.00		
400 " satd	"	1.80		
50 " satd	"	1.50	(252)	(2.99)
BFW	M lb	.25	284/hr	0.56
Cooling Water	M gal	.02	5.2/min	0.05
Power	KWH	.02.7	4560/hr	0.98
Fuel Gas	MM Btu	1.25	36/hr	0.36
Cat. & Chem.	M $/yr			0.20
			TOTAL	(0.84)
	Case 2			
Steam				
400 psig suphted	M lb	$2.00	(382/hr)	(6.05)
400 " satd	"	1.80		
50 " satd	"	1.50		
BFW	M lb	.25	528/hr	1.05
Cooling Water	M gal	.02	52/min	0.49
Power	KWH	.02.7	13600/hr	2.91
Fuel Gas	MM Btu	1.25		
Cat. & Chem.	M $/yr			0.40
			Total	(1.2)

FIGURE 3.4-1

CLAUS PROCESS VARIATIONS
SPLIT FLOW
90-93+% SULFUR RECOVERY

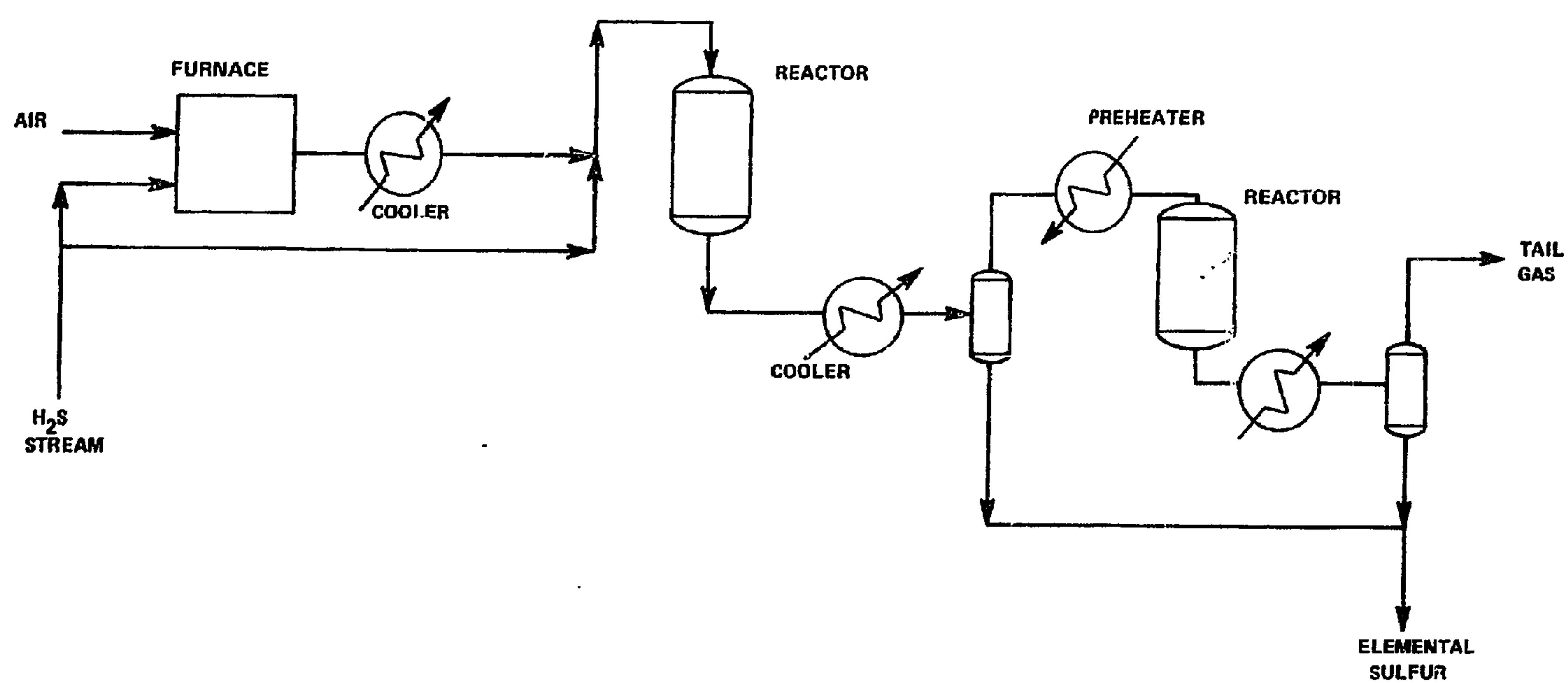


FIGURE 3.4-2

CLAUS PROCESS VARIATIONS
STRAIGHT THROUGH
93- 95+% SULFUR RECOVERY

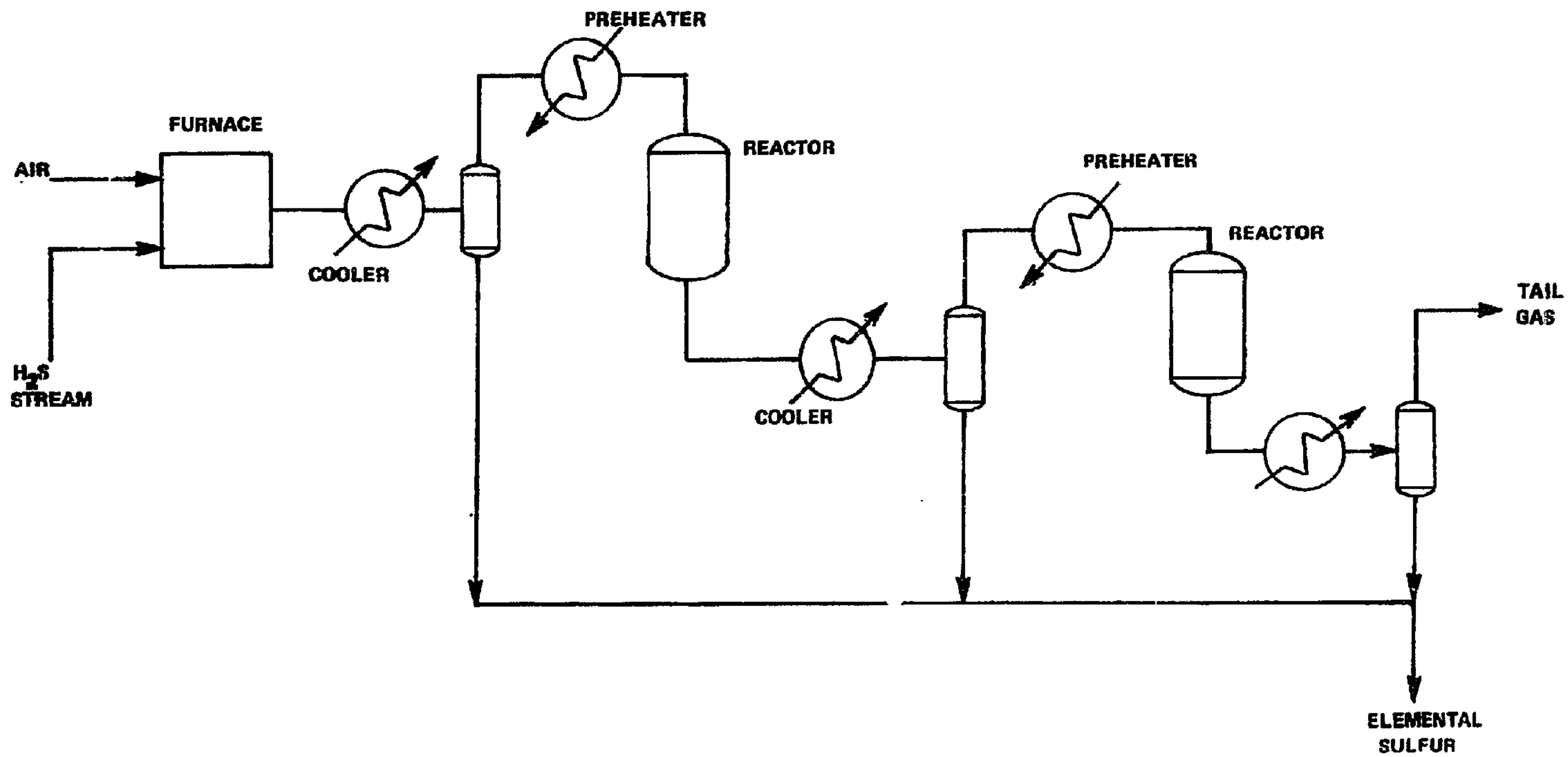


FIGURE 3.4-3

CLAUS PROCESS VARIATIONS
SULFUR RECYCLE
70- 90+% SULFUR RECOVERY

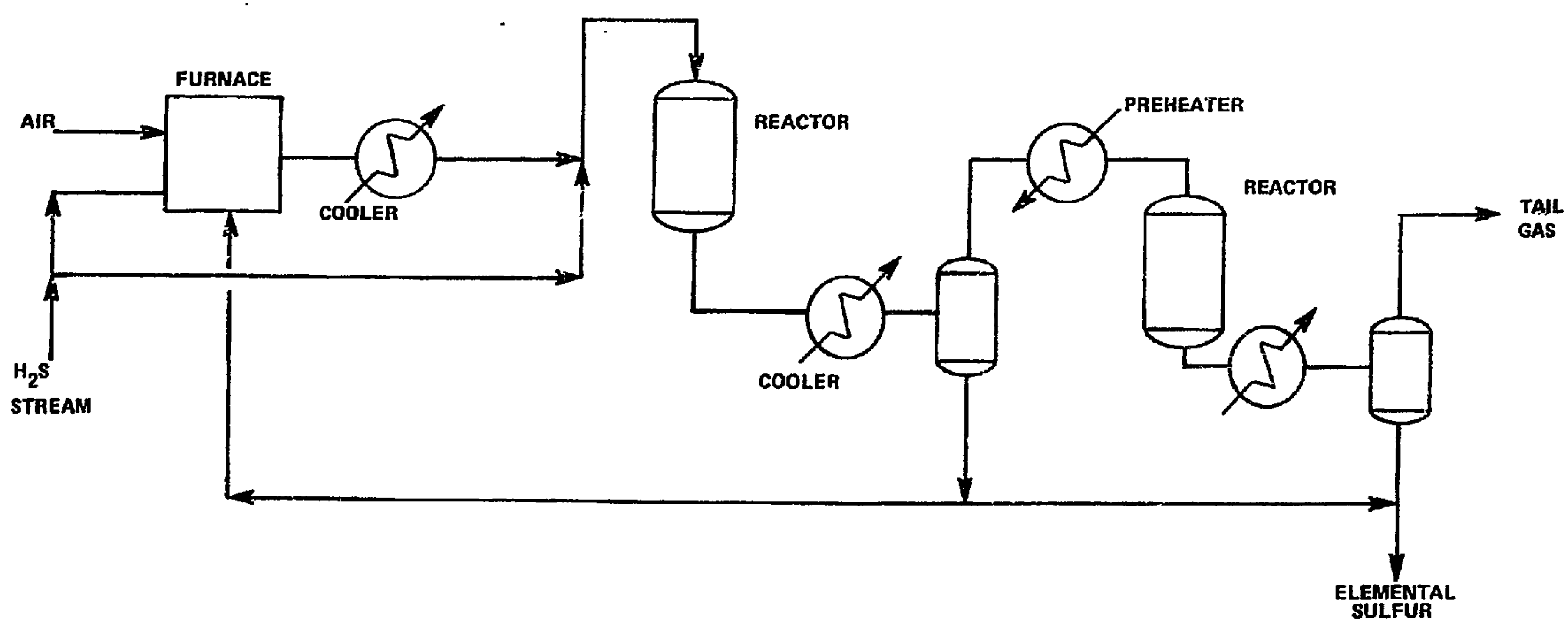


FIGURE 3.4-4
CLAUS PROCESS VARIATIONS

DIRECT OXIDATION
70-90+% SULFUR RECOVERY

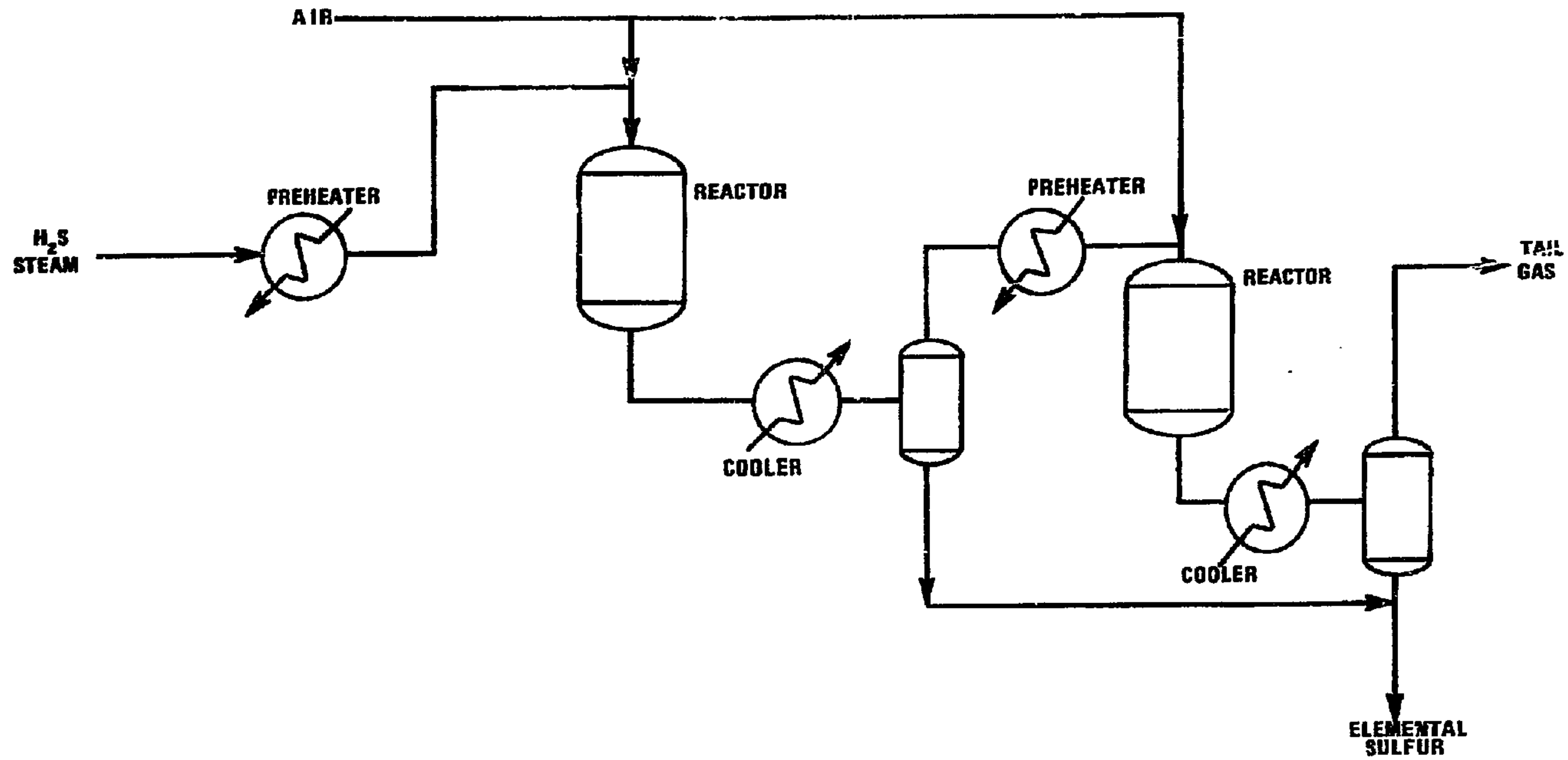


FIGURE 3.4-5

DOUBLE ABSORPTION CONTACT SULFURIC ACID PROCESS

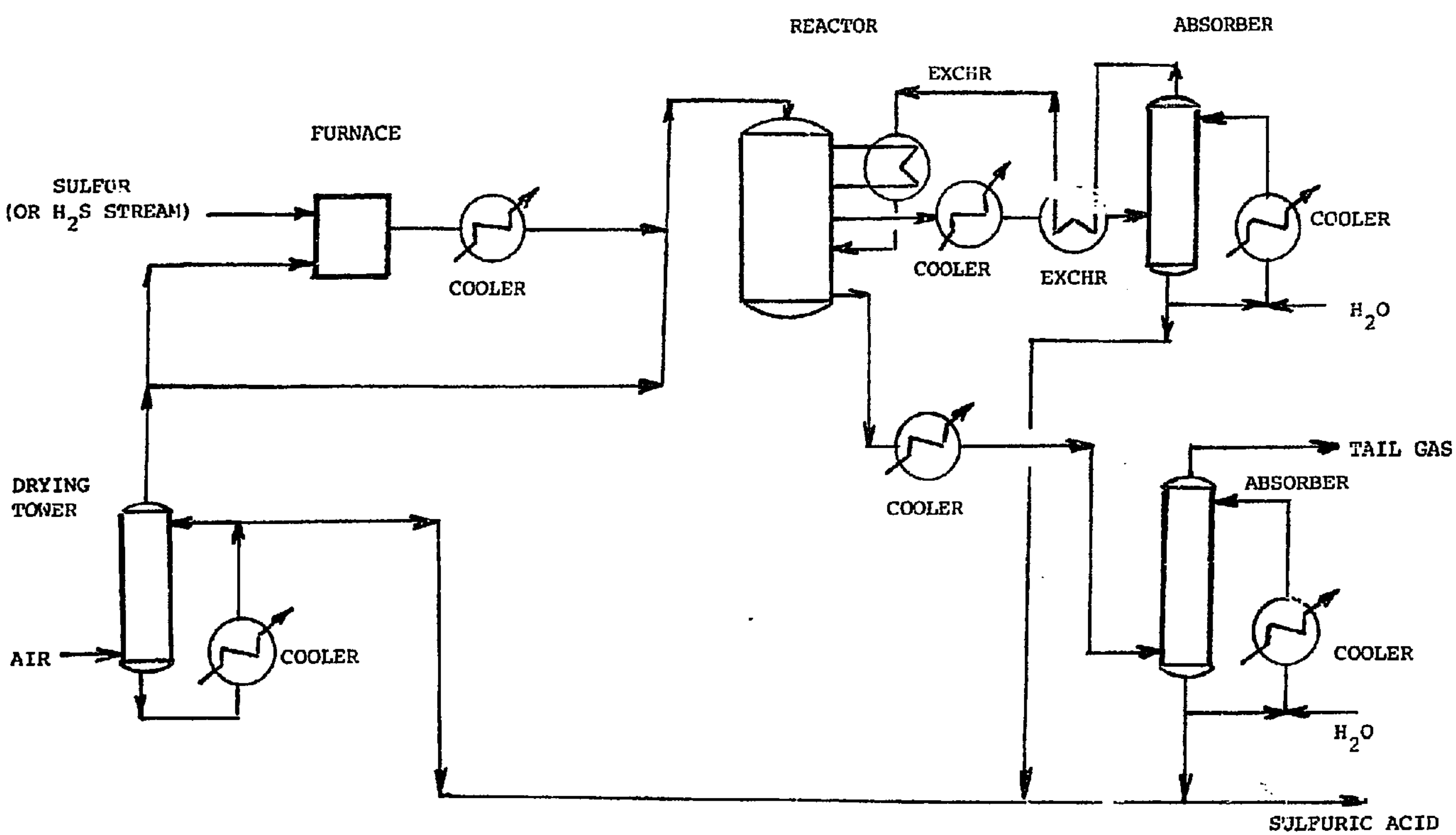
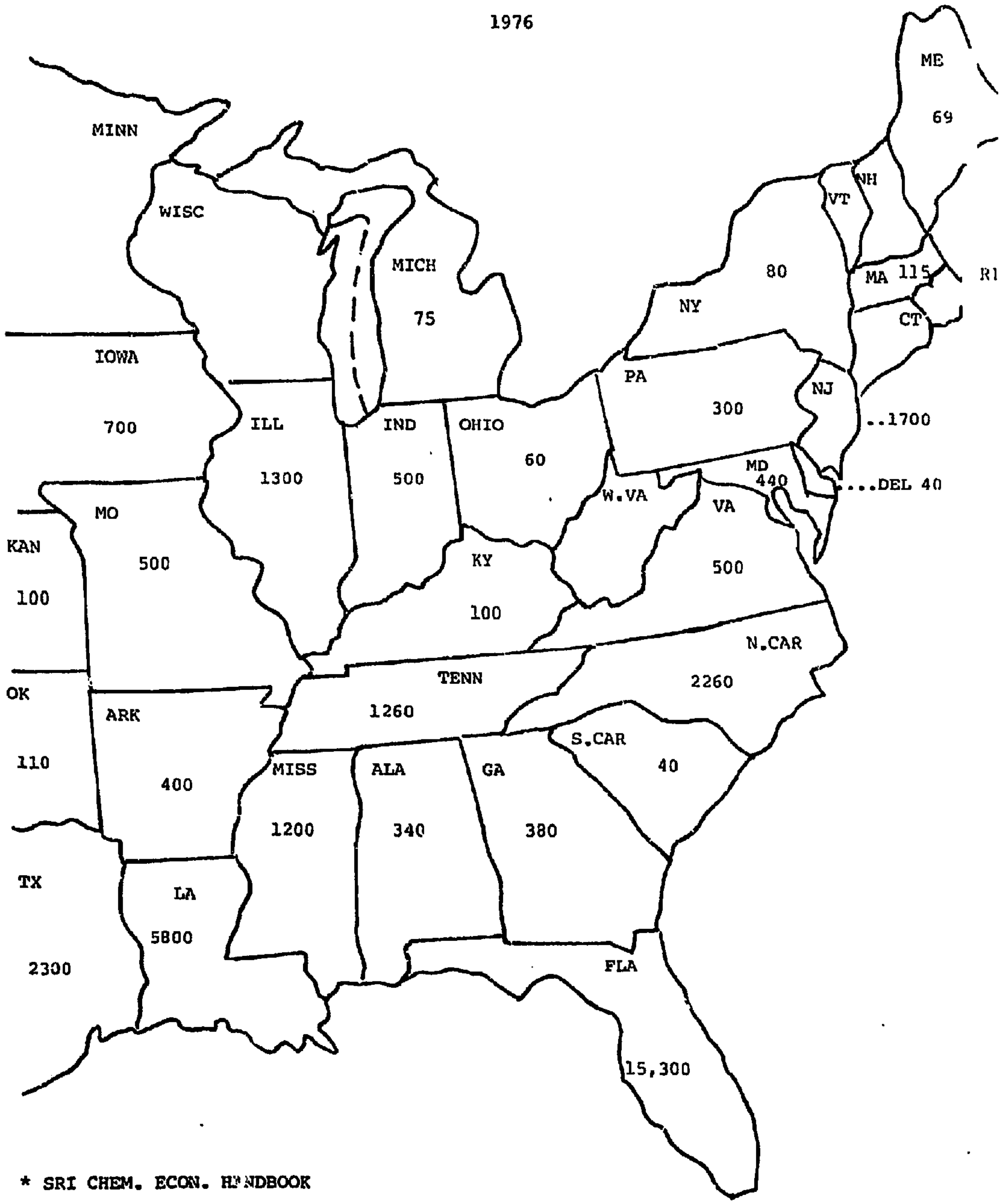


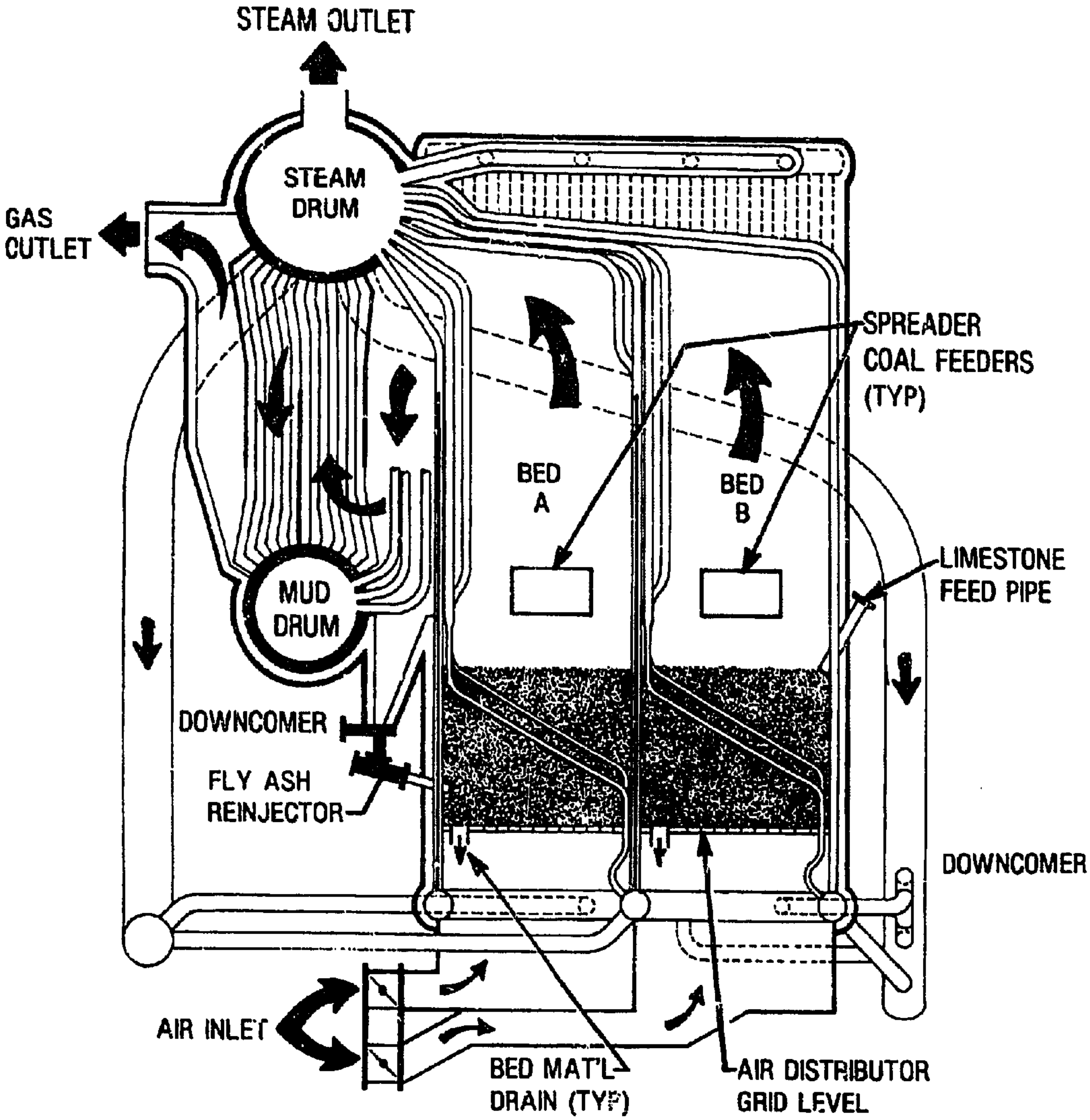
FIGURE 3.4-6

SULFURIC ACID PRODUCTION IN EASTERN USA *
ELEMENTAL SULFUR FEEDSTOCK
THOUSANDS OF SHORT TONS PER YEAR

1976



* SRI CHEM. ECON. HANDBOOK



STEAM GENERATOR ARRANGEMENT. - FLUIDIZED BED

FIGURE 3.4-7

SECTION 3.5

EQUIPMENT DRIVES

ASSESSMENTS AND PROCESS SELECTIONS

EQUIPMENT DRIVES

Introduction

The objective of this assessment is to identify the major equipment drivers required by coal gasification plants employing the following type of gasifier:

- Babcock and Wilcox entrained flow
- Texaco entrained flow
- Lurgi dry ash
- BGC/Lurgi slagger
- Koppers Totzek entrained flow

In addition, the candidate methods for supplying the driver power requirements are described and the rationale for selecting the preferred method, as incorporated in the baseline design studies, is discussed.

Major Equipment Drives

The major prime movers for rotating equipment incorporated in baseline gasification plant designs are summarized in Table 1.

Common to all five types of gasification plants is the large power usage needed to drive the compressors associated with the air separation section. In general, the air compressors require the largest single prime movers in the plant, the exception being the product gas compressors for those designs employing low pressure gasification processes. These compressors are rated at 20,000 to 40,000 BHP. The oxygen compressors, which are also large capacity machines, fall into the 10,000 - 20,000 BHP range.

With the exception of the Texaco gasification process, all the plants require product gas compressors to deliver gas at 600 psig to the plant gate. These machines vary in size from 4,000 to 30,000 BHP depending on the gasifier operating pressure and, in the case of K-T gasifier, the degree of intermediate gas compression.

Within the processing area, other large single drivers are found in the acid gas removal section of each plant. These are associated with solution circulation pumps, recycle gas compressors, and refrigeration compressors. The K-T based plant is unique in that the raw gas, generated at essentially atmospheric pressure, undergoes intermediate stages of compression in preparation for acid gas removal.

Large single drivers, over 1000 BHP, are generally required in the cooling water system and water treating sections of the support facilities. These are associated with the cooling water circulation pumps and boiler feed water supply pumps.

Driver Selection

Candidate drivers for supplying power for the major rotating machinery include the pressure reducing or back pressure steam turbine, the condensing steam turbine, and the electric motor. Gas turbines are excluded from consideration due to their high operating costs, i.e. the cost of natural gas or product medium BTU fuel gas is prohibitive.

The selection of electric motor or steam turbine drives for a gasification plant primarily depends on the plant steam system. In general, this type of plant cannot justify extremely high pressure steam, as do utility power plants. A typical system uses a variety of steam pressure levels in order to provide proper temperature required by the process scheme. When process users require intermediate or low pressure steam levels, such as 200 psig and 50 psig, this provides a good opportunity using pressure reducing steam turbines as drivers.

A pressure reducing steam turbine consumes high pressure steam delivered from the boilers and exhausts it at a lower pressure for use in the process. The pressure reducing turbine, in conjunction with a process user, offers the distinct advantage over other equipment drives of giving, by far, the lowest operating cost. The key to the low operating cost is the fact that while the turbine converts only a fraction of the available steam energy into shaft work, the remaining heat energy in the exhaust steam, particularly the latent heat, is useable for process heating purposes.

A condensing turbine drive also uses high pressure boiler steam, but exhausts steam into a condenser operating at a pressure of a few inches of mercury. This turbine uses a greater portion of the system energy than does the non-condensing type. However, the temperature and pressure of the exhaust steam is too low for process use, and consequently the latent heat of steam is lost by being transferred cooling media. As a result, the condensing turbine has the highest operating cost of the three alternatives for small drives and its cost approaches that of electric drives only for very large horsepower drives.

Electric motors are the most common equipment drives employed in process plants. Their reliability is good and their efficiencies range from 75% for small motors to 95% for 1000 horsepower and greater units. The operating cost of electric motors depends on the plant charge for electricity, but in general, it falls between the operating cost of the pressure reducing steam turbine and the condensing turbine. When the choice is producing steam for a condensing turbine drive or purchasing electric power for a motor drive, the balance favors electric drives when fuel costs for the power utility and process plant are equal.

A comparison of estimated operating costs of the three types of equipment drives is given below. This study was conducted by Foster Wheeler for a conceptual design of a commercial industrial fuel gas plant, based on coal, and the results are generally confirmed by independent literature sources

<u>Operating Cost of Delivered Power</u> *

	$ / MMBTU
Pressure Reducing Turbine	1.50 - 1.70
Electric Motor	6.10 - 6.50
Condensing Turbine	6.30 - 7.00

* Based on fuel at $1.25/MMBTU and electric power at $0.02/KWH for large drives (>1000 HP).

Based on this analysis, the recommended driver scheme for the baseline gasification plant designs is to maximize the use of back pressure steam turbines and to provide electric motor drives for the balance of the rotating equipment. Of course the use of topping turbines is limited by the need for the resulting low or medium pressure exhaust steam in the particular plant design. In instances where high pressure steam is generated via process waste heat boilers, condensing turbines are employed on large driving requirements to the extent that topping turbines cannot be fully utilized.

Steam and Power Supply

Of the five gasification plant concepts under consideration, only the Lurgi dry ash and BGC/Lurgi slagging gasifier cases do not generate significant amounts of waste process steam, by virtue of their relatively low gasifier outlet temperature (900-1000°F) and the need to quench the off-gases for tar removal. Consequently, the bulk of the process steam requirements has to be generated via coal fired boilers at elevated pressure and let down to process conditions via topping steam turbine drives.

The other three gasification cases (B&W, Texaco, and Koppers Totzek) employ relatively high temperature gasifiers which permit heat recovery from the off-gas via generation of high pressure steam. In these situations, the need for separate coal fired steam generators is minimized since the steam from process waste heat boilers is generally sufficient to satisfy all process steam requirements. Excess steam can be utilized to provide driving power in condensing turbines.

Therefore, in all of the gasification plants considered for this study, the in-plant steam generation capacity, either from coal fired boilers or waste heat boilers, is preferably designed to satisfy only the processing requirements, i.e. no substantial steam generation capacity is provided solely for equipment drive purposes. Those equipment drives which cannot be accommodated by the plant steam balance are delegated to electric motor drives.

This recommendation for selecting the type of plant drivers is based on the following criteria:

- no substantial difference in investment cost exists between electric motor drives and steam turbine drives, when all associated ancillary systems are included e.g. motor controls, electric switch gear, steam piping, etc.
- when equal fuel cost is considered for both steam generation, driver operating costs favor electric motors over condensing steam turbines.

Following the above recommendation to employ electric motor drives in preference to additional in-plant steam generation for turbine drives, the other option that should be considered is on-site electric power generation versus purchased electric power to satisfy the balance of the plant power requirements. The choice of electric power supply is influenced by the following factors:

- availability of purchased power
- relative costs of purchased and on-site generated power
- environmental constraints for the gasification plant complex

Under the proposed design philosophy for satisfying the gasification plant steam balances, the plants' electric power requirements depend on the type of gasifier selected, e.g.

Gasifier	Approximate Power, MW
B&W	250
Lurgi Dry Ash	100
BGC Slagger	80
Texaco	100
K-T	450

Assuming that purchased power is available to satisfy the above requirements from TVA's regional grid system, it is unlikely that on-site power generators would compete with TVA's relatively low cost power. This is apparent from the relative costs of coal fired electric power plants (1978 cost basis):

Capacity, M.W.	Plant Cost, $/KW
100	1150
450	600
1000	450

For equivalent coal cost, an on-site power plant with capacity of 100 to 400 MW will not be able to produce power at a cost comparable to that of TVA's large central power stations.

Furthermore, on-site power generation would represent additional emissions of sulfur dioxide, NO_x, and particulates associated with the gasification complex, as well as increased coal and ash handling facilities. Therefore, unless sufficient purchased electric power is unavailable, the recommendation is to exclude on-site power generation from the scope of the gasification plants.

Fluidized bed steam generators have been recommended for this project because of their capability to satisfy steam demands while simultaneously controlling SO_2, NO_x and particulate emissions to levels within the Federal and most state and local emission limits. Additionally, they can be designed to simultaneously burn uncleaned low BTU gas produced in the coal gasification process of this project. Furthermore, high heat stream may be ducted to these boilers utilizing them for the fluidizing media while capturing their sensible heat. Consequently, costly emission problems are eliminated by combining these streams and handling their difficult emission problems in a central facility offering a high efficiency energy recovery. They can also be designed to handle surplus coal fines produced by the coal handling facilities which is another advantage not without significant importance for this facility.

Fluidized bed steam generators are offered commercially in sizes to 600,000lb/hr., with superheated steam at desired pressure and temperature. An industrial unit supplying 100,000 lb/hr of saturated steam is operating satisfactorily in commercial service. Also a utility system supplying 300,000 lb/hr. of super-heated steam has shown that the fluidized bed technique can be integrated into plants requiring higher pressure superheated steam to drive steam turbines. Numerous installations are in various stages of design, construction and start-up throughout the world.

Form No. 130-171

TABLE 1

SUMMARY OF MAJOR PLANT DRIVERS (1)

Service	Babcock & Wilcox No.	Babcock & Wilcox BHP/Driver	Texaco No.	Texaco BHP/Driver	Lurgi Dry Ash No.	Lurgi Dry Ash BHP/Driver	BGC Slagger No.	BGC Slagger BHP/Driver	Koppers Totzek No.	Koppers Totzek BHP/Driver
Air Separation Plant										
Air Compressors	16	19,500	8	40,000	8	20,600	8	20,200		
Oxygen Compressors	8	11,700	8	20,000	4	7,100	4	6,900	8	1,500
Nitrogen Compressors	16	5,400			4	650				
Coal Gasification										
HP Water Pumps	-	-	8	1,200	-	-	-	-		
Lock Gas Compressors	-	-	-	-	4	1,700	4	1,000		
Acid Gas Removal										
Raw Gas Compressors	-	-	-	-	-	-	-	-	8	42,500
Lean Solution Pumps	8	1,700	8	2,200	4	?	4			
Recycle Gas Compressors	4	200	4	1,150	4	?	4			
Refrigeration Compressors	4	4,500	4	9,000	4	?	4	8,600		
Gas Compression										
Product Gas Compressors	4	29,000	-	-	4	7,000	4	3,500	4	15,800
Support Systems										
C.W. Circulation Pumps	12	?	12	3,800	16	900	8	850		
BFW Pumps	8	?	12	900	4	1,700	4	1,350		

(1) For 4-module plant

SECTION 3.6

PLANT RELIABILITY, SPARES & GAS STORAGE

ASSESSMENTS AND PROCESS SELECTION

PLANT RELIABILITY, SPARES, AND GAS STORAGE

Introduction

Several of the tasks required under TVA contract number TV-53453A involve an estimate of plant availability. An ideal data source for such an estimate would be the existence of identical plants with four or five years of operating experience. Since such a data base does not exist, indirect methods must be used to estimate the availability of the proposed designs.

The purpose of this report is to define these areas where major equipment redundancies are necessary to meet the required availability. Also, this study should identify those areas where parallel lines of equipment should be used. Finally, the merits of storage will be developed in terms of plant availability.

Basic Approach

The approach used was to consider process blocks and attempt to estimate the availability of the assembly of these blocks in the plant modules. Blocks were subdivided when necessary to estimate the need for redundant equipment trains to achieve the overall design availability.

The Process Block Flow Diagrams were reviewed to establish the number of units critical to the establishment of overall module availability. A typical count of critical units is as follows:

B & W Gasifier	11 critical units
Lurgi Dry Bottom Gasifier	12 critical units

Critical units are defined as those which could shut down an entire plant module due to unscheduled outages. Units which can be designed so as to allow their shutdown without causing an immediate plant shutdown are considered non-critical. Availability estimates were made for each of the critical blocks to develop overall plant module availability.

Many of the critical blocks were treated as single units. These units were considered to have a high degree of availability as indicated by past experience.

Several units were treated as consisting of several parallel trains. This was done where redundancy was required to increase module availability or to justify storage for the same purpose.

Three types of equipment and/or block arrangements are possible. These are as follows:

1. Equal or unequal units arranged in series.
2. Unequal units arranged in parallel.
3. Equal units arranged in parallel.

The following formulas were used to estimate the availability of the arrangements given above:

Series

$$A = A_a A_b \cdot \cdot \cdot A_n \tag{1}$$

Unequal Parallel

$$A = 1 - (1-A_a)(1-A_b) \cdot \cdot \cdot (1-A_n) \tag{2}$$

Equal Parallel

$$A \left\{ \leqq x \right\} = \sum_{x=0}^{x} \frac{N!}{(N-x)!x!} a^{(N-x)} (1-a)^{x} \tag{3}$$

N = Total number of units

x = Number of units out of service

a = Availability of a single unit

A = Availability of the toal block

Equations 1 and 2 are straightforward and commony used in reliability studies. Equation 3 has also been used in reliability studies. This last equation is a method of approximating the probabilities for one of two possible events.

Data Sources

A limited amount of published data exists to assist in this type of analysis. Information used for this present analysis includes:

a. Oxygen Plant Operating Data from several papers by Air Products.

b. Lurgi Dry bottom gasifier operating experience as published by SASOL.

c. Partial Oxidation Unit experience as published by Shell for their process.

d. Claus plant operating experience published by the Amoco Production Company.

Another source of data are the Lurgi Single-Train Centrifugal-type ammonia plant surveys. Data from the 1975-1976 survey for 30 plants, which covers the period from 1969 through 1976 reports on the operating experience of 30 plants.

Available Data

Air plants outages have been reported as follows:

Scheduled	4.3 days/year
Unscheduled	3.0 days/year
	7.3 days/year

It is assumed that scheduled outages for the air plant can be set so as to occur simultaneously with the scheduled outages for other units. Thus, the air plant availability becomes 0.9918. This is without oxygen storage.

Information from air products indicates that a liquid oxygen storage system has an availability of about 0.9996 which corresponds to an availability of about 3½ hours per year. Parallel gaseous storage with a capacity of 15 minutes should bring the total oxygen supply system to an availability of 1.0000 and has been assured in this analysis.

SASOL reports that a design operating factor of 35% for the Lurgi dry bottom gasifier is to be used for SASOL II. This availability derives from their more than twenty years of experience with SASOL I. The other gasifier with considerable commercial experience is the Kopper Totzek. However, published gasifier availability data has not been found for this process. Since the other gasifiers have not yet been demonstrated in large commercial plants, the SASOL experience has been used as a bench mark.

Amoco Production Company has reported availability data for five plants. These show an average of 6.0 days outage per year for an availability of 0.9836.

Shell has reported gasifier and overall plant availability for their partial oxidation plants. The reported data for oil fed gasification plants shows an average gasifier availability of 0.9388. This data was used to estimate the availability of each section of the plant. This data is shown in Table 1.

TABLE 1

		SHELL (1) PO PLANTS	
		Availability	Days
1.	Air Separation	0.9988	0.4
2.	Coal Prep	-	-
3.	Gasification	0.9478	19.1
4.	Shift Converter	0.9988	0.4
5.	Acid Gas Removal	0.9988	0.4
6.	Gas Compression	0.9988	0.4
7.	Sulfur Plant	0.9988	0.4
8.	Tail Gas Treating	-	-
9.	CO Removal	0.9988	0.4
10.	Ash Handling	-	-
11.	Utility System	0.9988	0.4
12.	Waste Treatment	0.9988	0.4
		0.9387	22.4

(1) For PO Plants known values are gas and overall availability other equally distributed

Large single-train centrifugal type ammonia plants have been surveyed for a number of years in relation to their operating experience. The survey for 1975-1976 for 30 plants shows outages as follows:

TABLE 2

	Shutdowns (1)	Down Days (2)	Percent of Down Days	Availability
Instrument Failures	1.5	1.5	3	0.9959
Electrical Failures	1	1.5	3	0.9959
Major Equip. Failures	6	22	44	0 9397
Turnarounds	0.5	20	.40	0.9452
Others	2	5	10	0.9863
	11	50	100	(0.8630)

(1) number of shutdowns per year per plant

(2) down time in day/year/plant

Major equipment failures average for the ammonia plants surveyed into the categories shown in Table 3.

TABLE 3

Total Days Outage	50		45.5		49		50	
Calendar Year	1969-1970		1971-1972		1973-1974		1975-1976	
	%	Days/ Year	%	Days/ Year	%	Days/ Year	%	Days/ Year
Syn. Gas Compr.	13	6.5	16	7.3	16	7.8	25	12.5
Tubes, Risers & Manifolds	19	9.5	17	7.7	19	9.3	13	6.5
Piping, Flanges & Relief Valves	-	-	-	-	5	2.5	11	5.5
Exchangers	10	5.0	9	4.1	8	3.9	11	5.5
Waste Heat Boilers	21	10.5	10	4.6	-	-	8	4.0
Transfer Header	6	3.0	-	-	6	2.9	7	3.5
Air Compressor	-	-	11	5.0	9	4.4	-	-
Ammonia Conv.	-	-	8	3.6	-	-	-	-
Htr. Convert. Coils	5	2.5	-	-	-	-	-	-
Total Days For Reasons Listed		37.0		32.3		30.8		37.5

The outages shown in Table 3 were used to calculate the average availabilities shown in Table 4.

TABLE 4

Item	Availability
Syn. Gas Compr.	0.9767
Tubes, Risers & Manifold	0.9774
Piping, Flys & Relief Valves	0.9891
Exchangers	0.9873
Waste Heat Boiler	0.9825
Transfer Header	0.9914
Air Compressor	0.9871
Ammonia Conv.	0.9901
Convection Section Coils	0.9932

Project Reliability Requirements

Three gasification schemes were selected to set the approximate availability requirements for each section. These are shown in Tables 5, 6, and 7. The selected processes were:

Lurgi Dry Bottom Gasifier
BGC/Lurgi Slagging Gasifier
B&W Slagging Gasifier

These tables were used to estimate the need for sparing of major equipment to achieve the design module availability of 90%. An availability requirement of 0.95 was set for the gasifiers so as to meet the overall availability requirement. The critical item in this analysis is the gasifier availability.

Figure 1 shows the results for the B&W Gasifier. In this case, a total of three would be installed per module with two operating units required. From Figure 1, it can be seen that a single gasifiers' availability of about 0.875 would be required to achieve a gasification unit availability of 0.95 at the 100% plant capacity. Since this requirement is somewhat above the single gasifier availability achieved at SASOL, a second case was plotted in Figure 2. This figure shows that with four units installed and two operating, the single gasifier availability drops to about 0.75. This case was chosen as the design case.

The Lurgi dry bottom is shown in Figure 3. This shows that for seven operating gasifiers, 10 units are required to achieve a single gasifier reliability of 0.850.

Similarly, Figure 4 shows that five BGC/Lurgi Slagging Gasifiers are required with three operating units to meet the system requirement. Figure 5 shows three operating Texaco Gasifiers with a total of five units installed.

The B&W and Texaco Gasifiers require the use of coal pulverizers. These units are believed to be somewhat less reliable than the gasifier. A single unit availability of 0.775 was selected for the design. It was assumed that one pulverizer per module would be required for a total of four pulverizers. The total installed requirement is seven to meet the design's availability. This calculation is summarized in Figure 6.

Figure 7 shows a plot of oxygen availability versus hours of oxygen storage. Curve 1 shows that a system availability of 1.00 can be achieved with a minimum of six hours liquid storage. Short term gaseous storage - say about 15 minutes - should be provided to handle unusual situations such as a stuck valve.

The above analysis has looked at process systems as complete blocks. Examination of specific equipment items has not been considered necessary for the present effort. The final design should include the type of redundancies and backup typically found in Process Plants. Examples of this are spare pumps, spare turbine driven pumps with automatic start systems, and handwheels in control valves.

Product Gas Storage

Product Gas Storage is an alternative method of achieving the plot design operating factor. Various forms of storage include:

Line Packing

Underground Storage

Pressured above ground storage

Atmospheric above ground storage

The volume of gas is quite large. One days' storage is about 284 MMSCF. Storage costs, excluding any compression costs will run in excess of $1.00/Ft.^{3}. Previous work by the contractor has demonstrated that large volume storage is too costly to merit serious consideration for this project.

TABLE 5

TVA STUDY

LURGI DRY BOTTOM

		Availability	Outage Days/Year
1.	Air Separation	1.0000 (1)	0.0
2.	Coal Prep	0.9998	0.1
3.	Gasification	0.9500	18.3
4.	Gas Cooling & Scrub	0.9945	2.0
5.	Acid Gas Removal	0.9945	2.0
6.	Gas Compression	0.9918	3.0
7.	Sulfur Plant	0.9974	0.9
8.	Tail Gas Treating	0.9974	0.9
9.	Gas Liquor Sep.	0.9945	2.0
10.	Ash Handling	0.9945	2.0
11.	Utility System	0.9945	2.0
12.	Waste Treatment	0.9945	2.0
		0.9066	34.1

(1) Based on liquid & gaseous backup storage

TABLE 6

TVA STUDY

BGC/LURGI SLAGGING

		Availability	Outage Day/Year
1.	Air Separation	1.0000 (1)	0.0
2.	Coal Prep.	0.9998	0.1
3.	Gasification	0.9500	18.3
4.	Gas Cooling & Scrub	0.9945	2.0
5.	Acid Gas Removal	0.9945	2.0
6.	Gas Compression	0.9918	3.0
7.	Sulfur Plant	0.9974	0.9
8.	Tail Gas Treating	0.9974	0.9
9.	Gas Liquor Sep.	0.9945	2.0
10.	Ash Handling	0.9945	2.0
11.	Utility System	0.9945	2.0
12.	Waste Treatment	0.9945	2.0
		0.9066	34.1

(1) Based on liquid & gaseous backup storage

TABLE 7

TVA STUDY

B & W

		Availability	Outage Days/Year
1.	Air Separation	1.0000 (1)	0.0
2.	Coal Prep.	0.9850	5.5
3.	Gasification	0.9500	18.3
4.	Gas Cooling & Scrub	0.9945	2.0
5.	Acid Gas Removal	0.9945	2.0
6.	Gas Compression	0.9918	3.0
7.	Sulfur Plant	0.9974	0.9
8.	Tail Gas Treating	0.9974	0.9
9.	Gas Liquor Sep.	-	-
10.	Ash Handling	0.9945	2.0
11.	Utility System	0.9945	2.0
12.	Waste Treatment	0.9945	2.0
		0.8981	37.2

(1) Based on liquid & gaseous backup storage

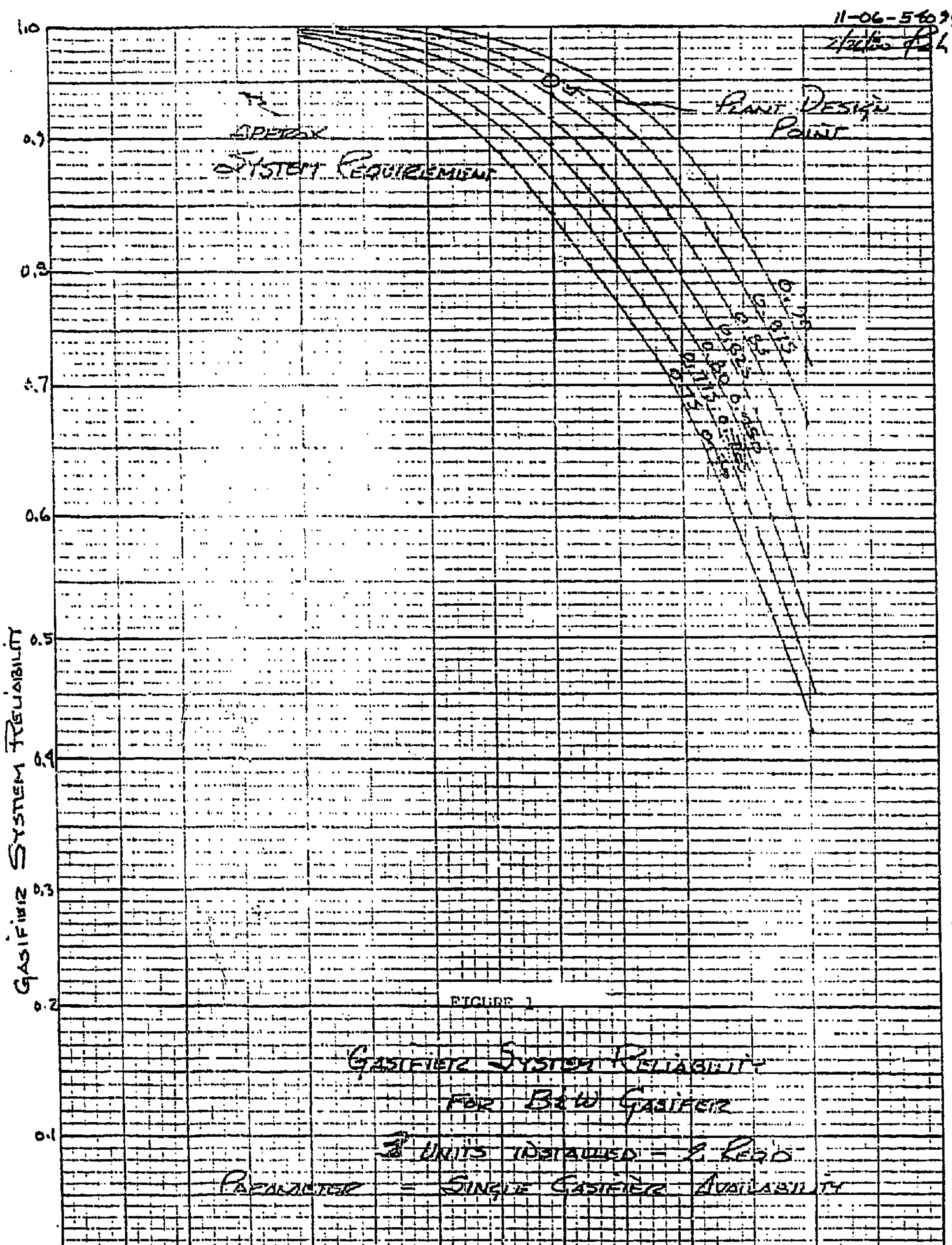


FIGURE 1

GASIFIER SYSTEM RELIABILITY FOR B&W GASIFIER

3 UNITS INSTALLED = 2 REQD

PARAMETER = SINGLE GASIFIER AVAILABILITY

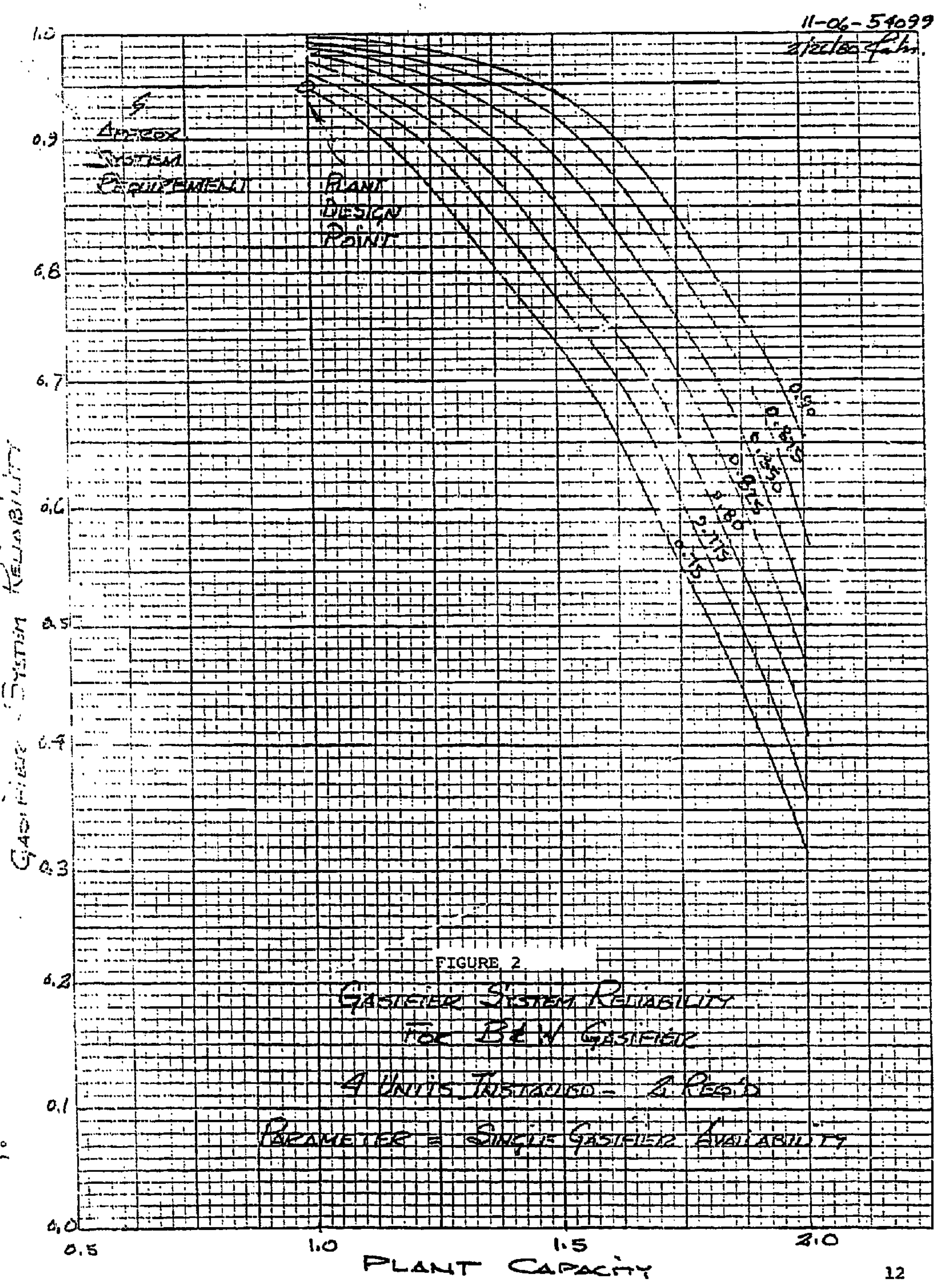


FIGURE 2

Gasifier System Reliability For B&W Gasifier

4 Units Installed - 2 Req'd

Parameter = Single Gasifier Availability

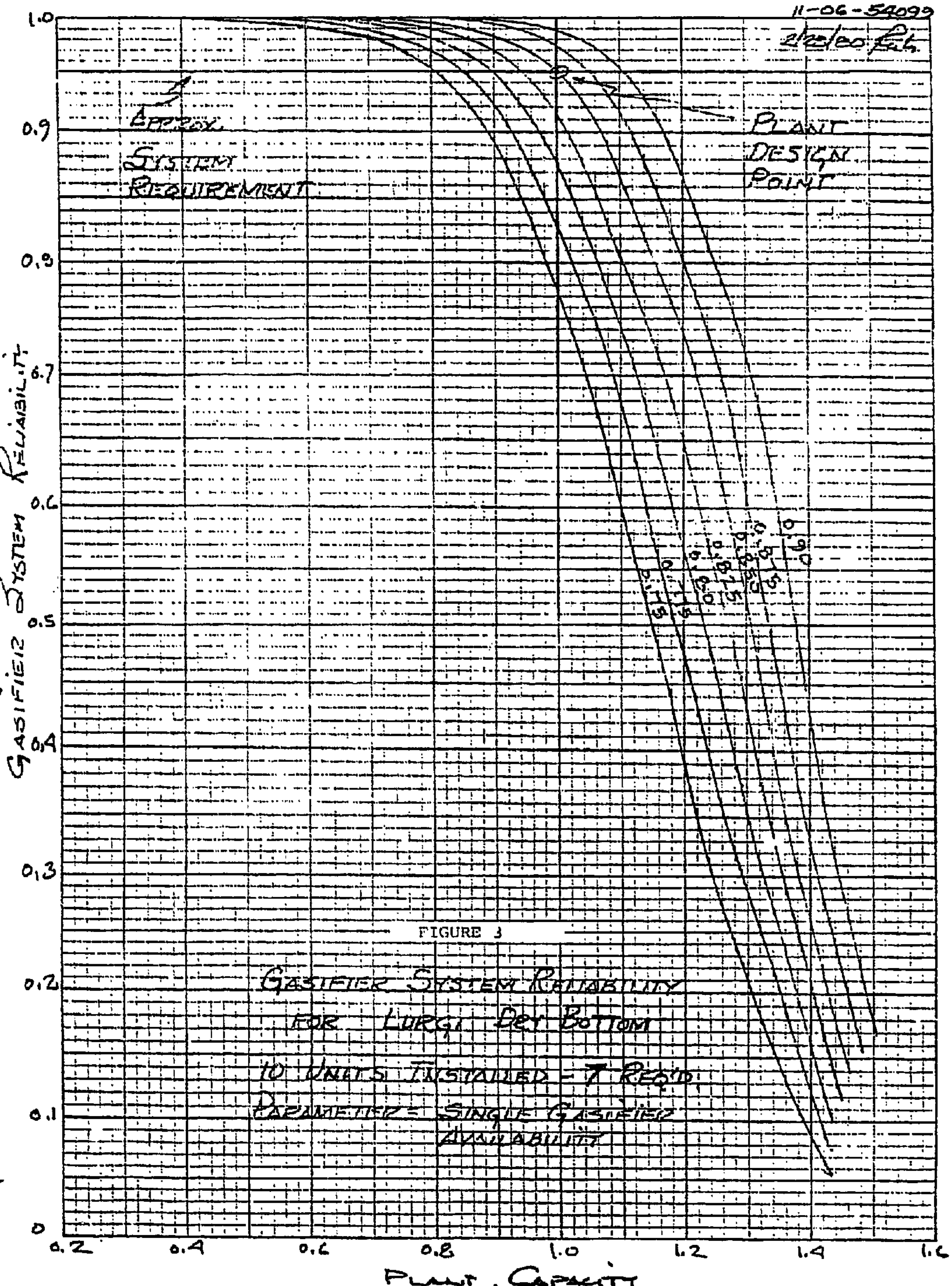


FIGURE 3

Gasifier System Reliability
for Lurgi Dry Bottom
10 Units Installed - 7 Req'd.
Parameter = Single Gasifier
Availability

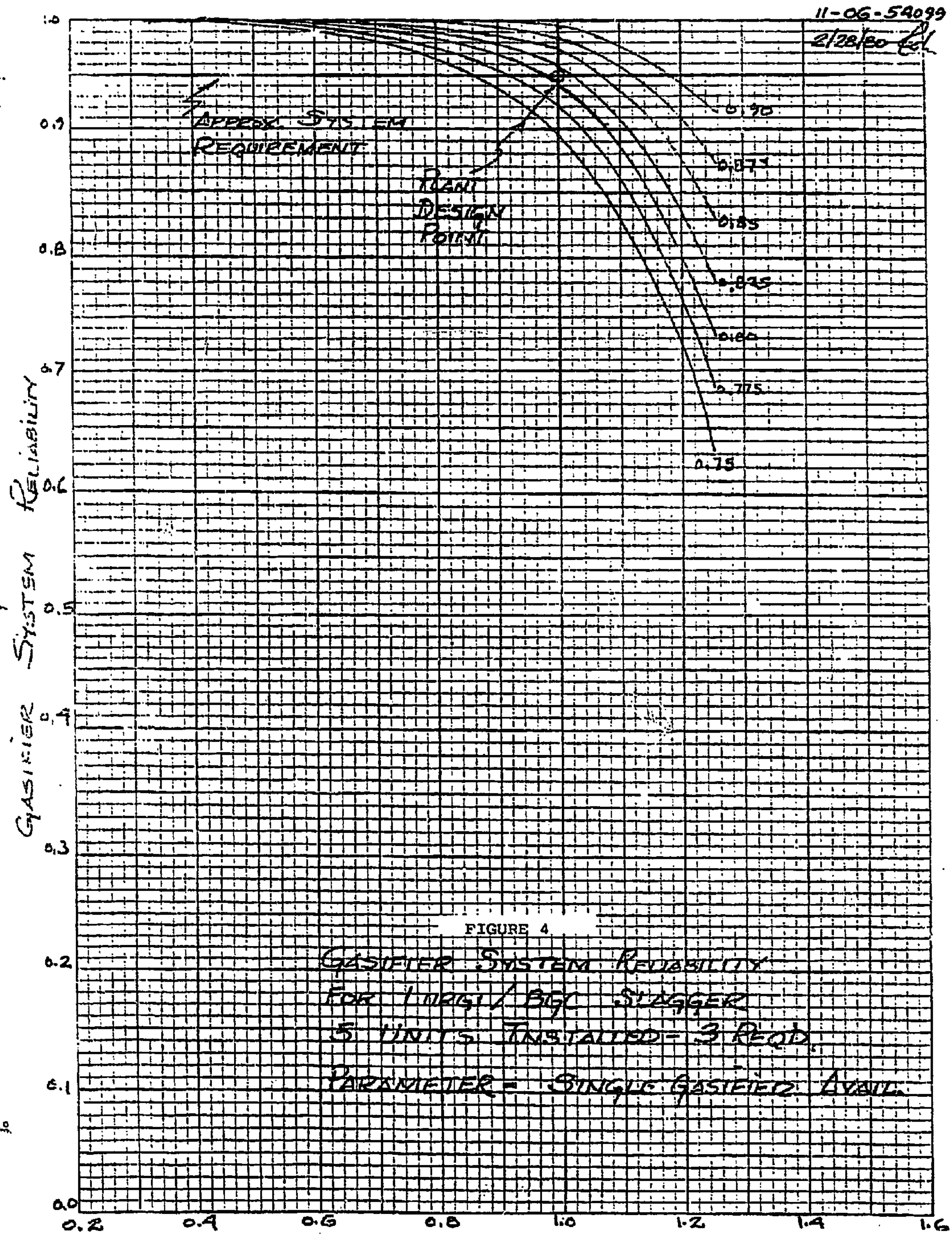


FIGURE 4

GASIFIER SYSTEM RELIABILITY FOR LURGI/BGC SLAGGER 5 UNITS INSTALLED - 3 REQD. PARAMETER = SINGLE GASIFIER AVAIL.

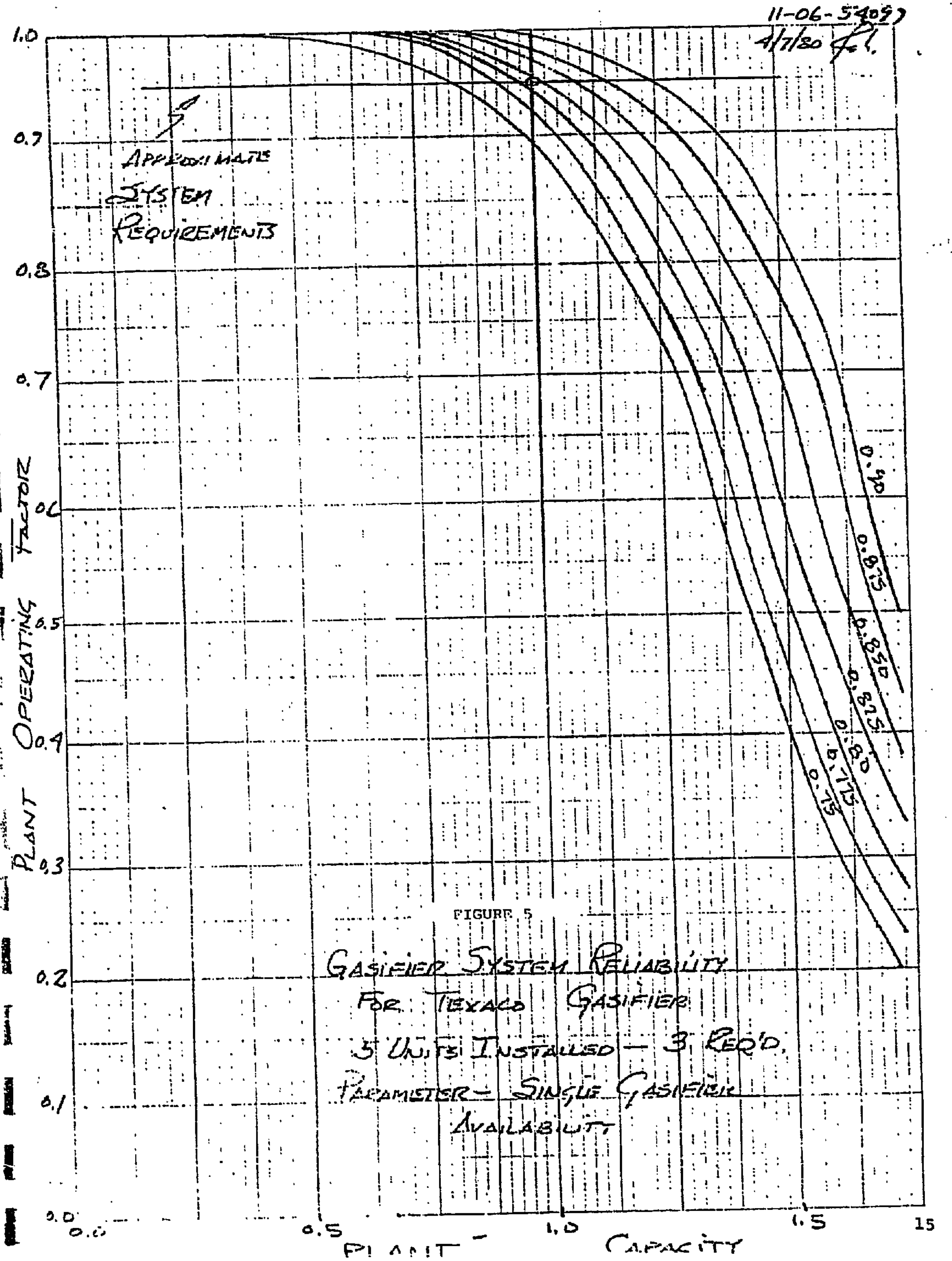


FIGURE 5

GASIFIER SYSTEM RELIABILITY
FOR TEXACO GASIFIER
3 UNITS INSTALLED — 3 REQ'D.
PARAMETER — SINGLE GASIFIER AVAILABILITY

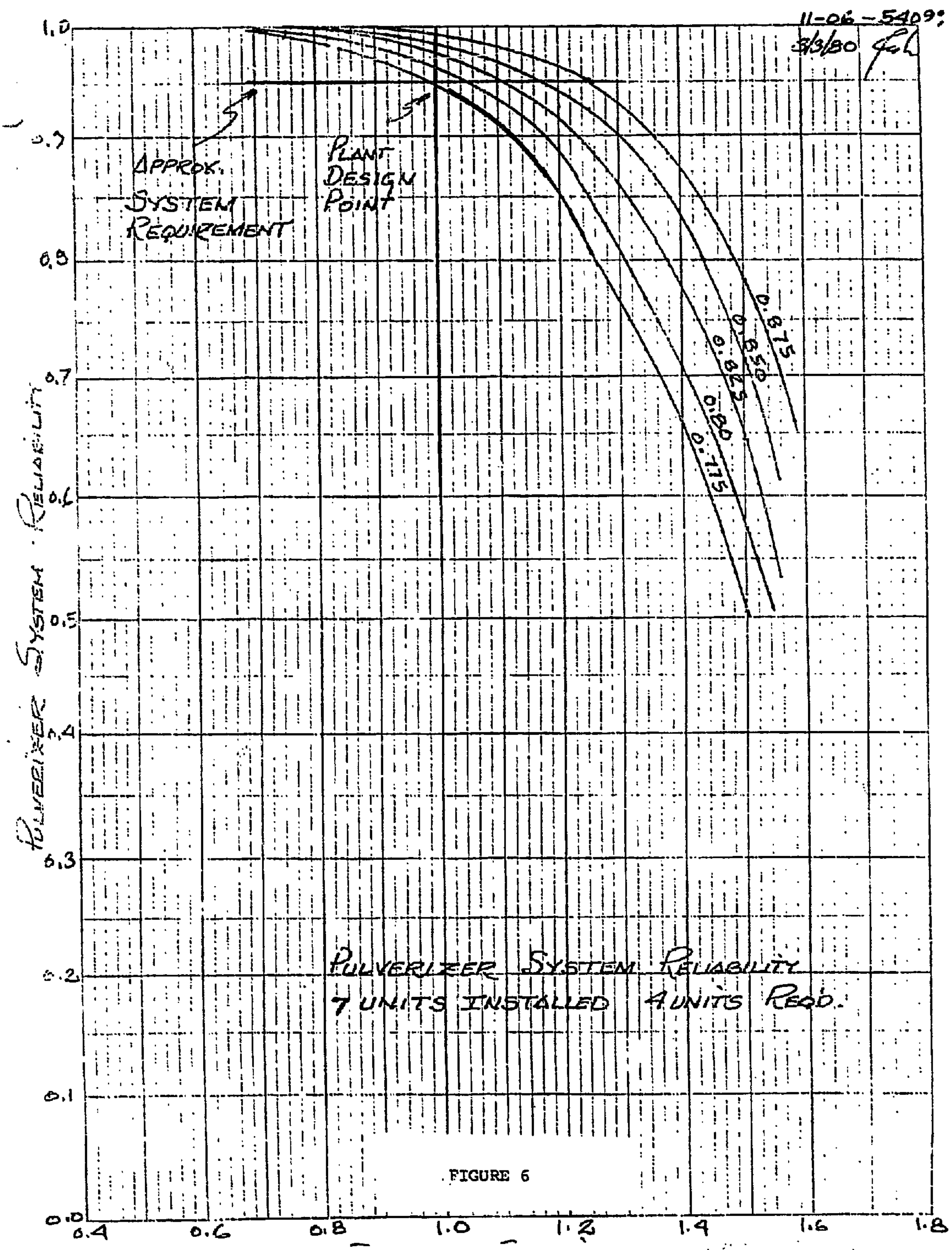
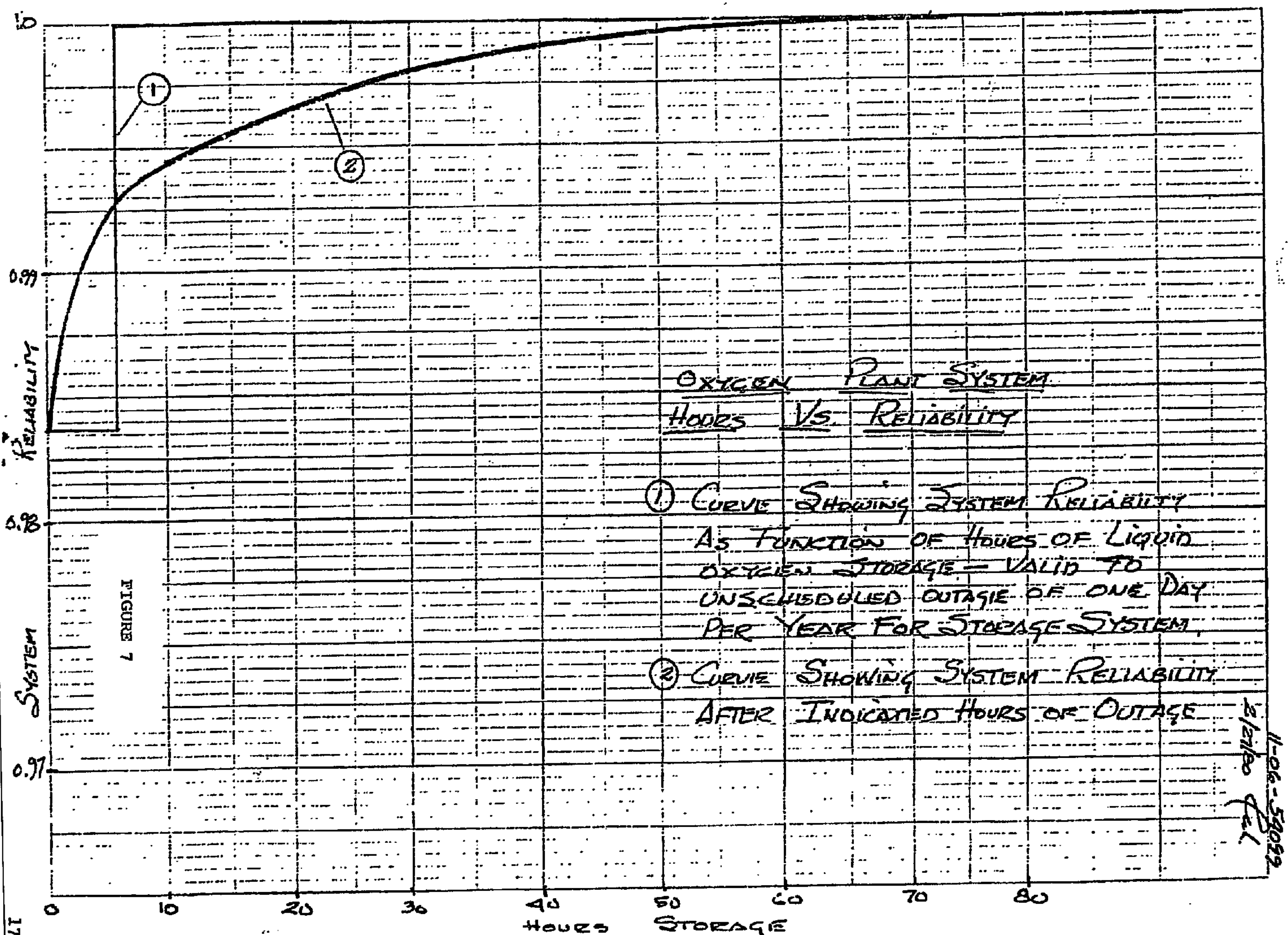

11-06-5409:
3/3/80
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.0
0.4
0.6
0.8
1.0
1.2
1.4
1.6
1.8
PULVERIZER SYSTEM RELIABILITY
APPROX. SYSTEM REQUIREMENT
PLANT DESIGN POINT
0.875
0.850
0.825
0.80
0.775
PULVERIZER SYSTEM RELIABILITY
7 UNITS INSTALLED 4 UNITS REQD.

FIGURE 6



1.0
0.99
0.98
0.97
System Reliability
0 10 20 30 40 50 60 70 80
Hours Storage
1
2
Oxygen Plant System
Hours Vs. Reliability
1 Curve Showing System Reliability As Function of Hours of Liquid Oxygen Storage — Valid To Unscheduled Outage of One Day Per Year For Storage System.
2 Curve Showing System Reliability After Indicated Hours of Outage
11-06-59099
2/27/80

FIGURE 7

ASSESSMENTS AND PROCESS SELECTIONS

GAS DELIVERY PRESSURE

Introduction

It is recognized that variations in gasifier pressure can have significant impact on plant operation. The optimum operating pressure is that pressure where minimum capital and operating cost occur. Reliability of operation is another factor that is considered when an operating pressure level is selected.

Benefits of Operation at Increased Pressure

In general, an increase of pressure is beneficial for the following reasons:

1. It may be possible to eliminate raw gas compression, thus decreasing both capital and operating costs, and also increasing plant reliability by avoiding raw gas compressor downtime and possible plant shutdown. The compression cost of oxygen increases, but since the oxygen volume is approximately one-third that of the product gas, a considerable compression savings is realized by increased gasifier pressure operation.
2. There will be lower capital and operating costs in the equipment downstream of the gasifiers due to the reduced volume of gas at higher pressures.
3. Higher pressure operation provides greater flexibility for rapidly controlling gasifier response to load variations.
4. Acid gas removal systems operate more efficiently at higher pressures; this is particularly true for absorbers in which a physical solvent (for example, SELEXOL) is employed. Also, the investment cost for the gas purification system is reduced.

5. Higher pressures enhance heat transfer and also provide reduced heat losses.
6. Increased pressure results in higher gasification rates/ unit reactor volume. Both capital investment and heat loss are reduced.

Disadvantages of Increased Pressure Operation

The disadvantages of increased gasifier pressure operation, especially applied to the production of a medium BTU gas, are:

1. As the pressure increases the methane yield also increases. This would be beneficial for high BTU (SNG) production, but for MBG product necessitates the incorporation of a reforming step to convert the unwanted methane into hydrogen and carbon monoxide. The effects of higher pressures on methane yield can be counterbalanced by operating the gasifier at higher temperatures, however.
2. Higher pressure operation usually increases the difficulties associated with feeding a gasifier through a lock hopper system. Also, it can be anticipated that the design of gasifiers not already capable of operating at elevated pressures would have to be modified. Possible benefits relating to the gasifier itself might be: a) more efficient gasification by improved gas mixing, b) reduction in the number of gasifiers.

Specific Gasifiers

Despite the obvious advantages of increased pressure on gasification plant design, the limiting factor is the highest operating gasifier operating pressure stipulated by the gasifier developer. For the gasifiers considered in our study, the recommended pressure conditions recommended by the gasifier licensors are as follows:

Gasifier	Pressure (psig)
Lurgi Dry Ash	450
BGC/Lurgi Slagger	450
Babcock & Wilcox	225
Koppers-Totzek	15
Texaco	630

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ASSESSMENTS AND PROCESS SELECTIONS

GAS DELIVERY PRESSURE

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Lurgi Dry Ash	450
BGC/Lurgi Slagger	450
Babcock & Wilcox	225
Koppers-Totzek	15
Texaco	680

We have confirmed these levels of gasifier operating pressures with each licensor and cite below this rationale. For the Dry Ash, Slagger, B & W and K-T gasifiers these levels pretty much represent the pressures at which actual units have operated, and are thus guaranteed. The Texaco gasifier has operated over the range of 350 to 1200 psig; the selection of 680 psig was made to maintain the gasifier at the level necessary to generate product gas at the required 600 psig level with no attempt made to optimize the operating pressure with respect to product gas cost.

Rationale Provided by Licensors for Selection of Gasifier Operating Pressures

Babcock and Wilcox Gasifier

"The gas side pressure in the B&W gasifier, designed for TVA was established at 240 psig for metallurgical reasons. At all times the steam side pressure should be greater than the gas side pressure so that any leakage will be of steam into the gas stream instead of dirty gas into the steam. Because of H_2S corrosion, the maximum O.D. temperature has been limited to about 650^{o}F, so as to give a 20 year predicted life time for the tubes. A temperature of 650^{o}F corresponds to a steam pressure of 250 psig".

Lurgi Dry Ash Gasifier

"At present, Lurgi offers only one gasifier design which is capable of going to 450 psig. This pressure level is therefore the limit for Lurgi's commercial offering. The 450 psig design was originally selected by Lurgi for coal gasification plants to give overall economy; i.e. high gasifier throughput and downstream gas processing. Presently, pilot scale work at 170 TDP is being conducted to 1500 psig. The Lurgi gasifier can be operated at pressures less than 450 psig. Low pressure operation results in lower throughput per gasifier as well as less methane in the raw gas. Atmospheric units are in operation in Pakistan, but Lurgi would not recommend this approach".

Koppers-Totzek Gasifier

"The Atmospheric Entrained Bed Coal Gasification (EBCG) Process has provided its commercial application in the past 30 years using a large number of different solid feedstocks, including U.S. coals. The main advantage of the EBCG Process are: high reliability, easy operation and maintenance, suitability for a wide range of solid feedstocks, no environmental restrictions, and flexibility with regard to load changes. The main expected advantages for a future entrained bed high-pressure coal gasification process are:

1. Higher thermal efficiency because of energy savings due to oxygen compression instead of raw gas compression and a higher yield of useful gas per KG of coal especially in case of feeding a low reactive coal.
2. Higher throughput per gasification stream.

None of the processes under development at present has proven its industrial maturity and appears to be commercially available for large plants before the second half of this decade".

Texaco Gasifier

"The Texaco Coal Gasification Process has been operated successfully on a 15-ton-per-day pilot-unit scale with pressures up to 1200 psig and on a 160 ton-per-day demonstration scale at pressures up to 590 psig. The closely related Texaco Synthesis Gas Generation Process, using liquid hydrocarbons as feedstock, has been operated successfully on a pilot-unit scale at pressures up to 2500 psig and on a commercial scale at pressures up to 1200 psig.

Since the Solicitation of Proposals issued by TVA for this project specified a delivered gas pressure of 600 psig and since the Texaco Coal Gasification Process can operate over a wide range of pressures as indicated in the above paragraph, it was decided that the optimum gasification pressure would be that which would be just high enough to eliminate any intermediate compression of product gas. Thus, a gasifier pressure of 750 psig was selected and used to prepare the Conceptual Design dated February 11, 1980, which was sent to you as a basis for your study. We believe that a pressure close to 750 psig takes maximum advantage of the pressure flexibility of the Texaco Coal Gasification Process since, after allowance for pressure drops downstream on the gasifier, it will deliver a product gas close to 600 psig.

A gasifier pressure in the 750-psig range does not present any problems in obtaining an adequate supply of high-pressure oxygen since, as indicated above, oil-based gasification plants are now operating commercially at pressures up to 1200 psig with significantly higher oxygen delivery pressures".

Effects of Pressure on Product Gas Cost

An internal study at Foster Wheeler has explored the effects of pressure on product gas cost for a selected gasifier. This study demonstrated that there is a definite advantage to raise the gasification pressure to a level just high enough to eliminate the raw gas compressor. A reduction in product gas cost of 10.8¢/MM BTU was attained by operating the gasifier at a pressure 75 psig above distribution requirements. By raising the pressure 400 psig above the distribution pressure, the gas cost can be reduced by an additional 3.5¢/MM BTU.

For the two pressure levels cited, the compression investment costs represented the largest effects. At the lower pressure, a 75% decrease in net (oxygen, raw gas) compression costs was achieved over the normal gasifier operating (75 psig) pressure; at the higher level only a 40% decrease was achieved. This was due to less savings in the oxygen compressor at the higher level. However, the use of a higher pressure was more than compensated for by a decrease in the operating requirements. The principal advantage of the higher pressure level was due to the large net change in auxiliary steam generation, which was credited at fuel gas equivalent value.

SECTION 3.8

EFFECTS OF SCALE

ASSESSMENTS AND PROCESS SELECTIONS
EFFECTS OF SCALE

An important consideration in the planning of a coal gasification project is the capacity of the plant to be installed. There are obviously many aspects to this consideration, including the size and nature of the market available for the product, availability of feedstock, type and amount of financing that can be arranged for, and economies of scale of operation affecting the production cost of the product. The last aspect - economies of scale of operation - is the subject of this assessment task for TVA's proposed coal gasification demonstration plant project. For the purposes of this assessment, it is assumed that market considerations, feedstock availability, etc. do not impact on plant sizes which can be considered.

Components of Product Cost

Before discussing the specific aspects of economies of scale, it is necessary to recognize the major components of coal gasification product cost and to indicate the effects of scale of operation on each component. A variety of methods can be used to carry out "cost of product" calculations depending upon specific business requirements. The main components of product cost included in these calculations, however, are generally those listed in Table 1. This table was prepared for a coal gasification plant operation and reflects TVA's approach to product costs. In the case of a coal gasification plant, cost components include raw materials, utilities, direct operations, and indirect charges.

The raw material for the plant is coal. Utilities include electric power, raw water, and catalysts and chemicals. Included in the utilities category is the annual cost of disposal of ash. Direct operations includes shift operators and supervision as well as plant maintenance - labor and materials. Indirect charges include plant overhead, payments in lieu of taxes, depreciation of plant, interest on debt, and TVA corporate overhead.

Effects of Scale

Cost items listed in Table 1 tend to vary with plant capacity but the extent of variation depends on the specific cost item considered. The raw material - coal - and utilities - raw water, electric power, and catalyst and chemicals, as well as ash disposal are usually considered to be directly proportionally to plant size, although this is not necessarily true for small plants, for example, less than 1000 TPD coal consumption. The contract price of delivered coal could be appreciably higher for small plants than for large plants since the latter would represent a base load customer for a large share of the output of a given mine or group of mines. Efficiency of motors and drivers in a small plant would tend to be lower than that of similar equipment in a large plant. Large electric power customers currently benefit from lower unit cost rate although this situation may not exist in the future.

Plant operating labor required per unit of plant capacity tends to decrease with increasing plant size since large capacity equipment can be operated by about the same number of operators assuming adequate instrumentation and control systems. Multiple modules, however, usually require national staff. Maintenance labor and materials are usually expressed as a percentage of plant investment so that the cost of these items for unit of product drecreases with increasing plant capacity in a manner similar to the variation of plant cost with capacity. Indirect charges are also expressed as percentages of plant investment and consequently, vary with plant capacity in the same manner.

The major factor in the consideration of economy of scale is the relationship between installed cost of plant and plant capacity. Experience indicates that this relationship is given by the equation:

$$I = Io\left(\frac{C}{Co}\right)^{x}$$

Where I is the installed cost of plant having capacity C, Io is the installed cost of a reference plant having capacity Co and x is a dimensionless exponent. The value of x varies between about 0.6 and about 0.9 depending on the type and capacity of plant. The exponent has a value of about 0.6 for processing plant consisting primarily of

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equipment such as process heaters, boilers, filters, and solids handling equipment for which an increase in capacity is achieved by a proportional increase in surface area of tubes, filtering surface, etc. A value of 0.7 to 0.75 for x usually applies where the processing plant contains a mixture of different types of equipment.

The exponent x is approximately constant, however, only over a limited range of capacity, the limits of the range being defined as capacities where multiple trains or modules of equipment are required because of technical or practical limitations on equipment sizes. At these points, there is usually a discontinuity in the cost capacity relationship. A two train plant, however, generally costs somewhat less than twice a single train plant of the same total capacity because of savings in support facilities.

Product Cost Estimates as a Function of Plant Capacity

Approximate calculations were made of the cost of medium BTU fuel gas produced by a coal gasification plant over a range of plant capacities. Purpose of these calculations was to illustrate the effects of scale of operation on product cost. These calculations were made in accordance with the parameters given in Table 2 which specify the variation of cost components discussed previously. It should be noted that the parameters of Table 2 do not represent any of the specific designs developed by Foster Wheeler in this study for TVA but are typical of such designs.

The base or reference plant capacity was taken as 2500 TPSD of coal. At this capacity, it was assumed that coal cost is $35/ton and that this cost varies as capacity increases with exponent of -0.1. Power cost was taken as $30/MWH and raw water cost was taken as $0.40/M gal. These costs were also assumed to vary with a capacity exponent of -0.1. Operating labor for the 2500 TPD plant was taken as 32 operators/shift. This parameter varies with capacity with an exponent of 0.25.

Long term debt was assumed to be 100% of total fixed investment with an interest rate of 12%. Payment in lieu of taxes and plant depreciation was taken as 2% and 5% of total fixed investment respectively. Investment for processing units and offsites for the 2500 TPD plant was assumed to be 300 million dollars. This investment varies with capacity as shown in Figure 1. Variation in investment cost was estimated using an exponent of 0.7 up to the limit of a module - 5000 TPD.

Details of the economic calculations are given in Tables 3, 4, 5 and 6. Calculated cost of product fuel gas is plotted as a function of plant capacity in Figure 2. Gas cost was about $8.0/MM BTU for the 2500 TPD plant. Gas cost decreased rapidly with increasing plant size up to a capacity of about 10,000 TPD and then decreased slowly as capacity increased to 20,000 TPSD. The results indicate that most of the economic benefits of large scale operation are available in plants having capacity of 15,000 TPD or higher. TVA's selected nominal capacity of 20,000 TPSD is in this category.

Variation in product cost with plant capacity has been reported by Farnsworth[1] and in the Department of Energy report on large scale methanol from coal plants. Variations described in these publications are similar to the variation shown in Figure 2. Cost variation reported by Kermode [2] was also similar.

(1) Farnsworth, J.F., Koppers Company, IGT meeting August 1973.

(2) Kermode, R.I., Nicholson A.F. and Jones, Jr., J.E., Chemical Engineering, February 25, 1980.

TABLE 1

EFFECTS OF SCALE

COMPONENTS OF PRODUCT PRODUCTION COST

COAL GASIFICATION PLANT FOR FUEL GAS PRODUCTION

Raw Materials
- Coal - gasification
  - boilers

Operation and Maintenance
- Utilities
  - Electric power
  - Raw water
  - Catalyst and chemicals
  - Ash disposal
- Direct Operations
  - Shift Operators
  - Operating Supervision
  - Maintenance labor and materials
- Indirects
  - Plant overhead
  - Annual payments in lieu of taxes
  - Depreciation of plant
  - Interest on long term debt
  - Corporate overhead

TABLE 2

EFFECTS OF SCALE

GAS COST PARAMETERS

Basis:

Range of Plant Capacity, TPSD Coal	2500 - 25,000
Maximum module size, TPSD Coal	5000

Parameter	Base Value	Capacity Exponent
Plant Capacity, TPSD as received	2500	0.0
Coal Cost, $/Ton	35	-0.1
Power Cost, $/MWH	30	-0.1
Raw Water Cost, $/M gal.	0.4	-0.1
Operating Labor/shift	32	0.25
Debt, %	100	-
Interest on debt, %	12	0.0
Payment in lieu of taxes, % TFI	2	0.0
Depreciation, % TFI	5	0.0
Plant Investment, BL & Offsites, MM	300	per Fig. 1

TABLE 3

PLANT: TVA Coal Gasification Plant - Fuel Gas Product

CAPACITY: 2500 TPD Coal	TYPE INVESTMENT ESTIMATE: Conceptual
LOCATION	INVESTMENT - B.L. 204
OPERATING DAYS/YR 328.5	INVESTMENT -OFFSITES 96
	INTEREST DURING CONSTR. 36
FUEL GAS PRODUCT	WORKING CAPITAL 5
45 MMM BTU/SD	STARTUP AND TESTING 1
	TOTAL FIXED INVESTMENT 356 MM$
	LONG TERM DEBT 100%

ITEMS OF COST OPERATION OF PLANT	UNIT	UNIT PRICE	UNIT CONSUMPTION	COST PER UNIT PROD.
RAW MATERIALS				
COAL	Tons	$35.0	2500/D	1.94
TOTAL RAW MATERIALS				1.94
OPERATION AND MAINTENANCE				
UTILITIES				
ELECTRIC POWER	MWH	$30.0	240/D	0.16
RAW WATER	M Gal.	$ 0.40	4000/D	0.04
CATALYST AND CHEMICALS		$600,000/Year		0.04
ASH DISPOSAL		$500,000/Year		0.04
TOTAL UTILITIES				0.28
DIRECT OPERATIONS				
SHIFT OPERATORS 32 OPERATORS/SHIFT @ $ 20,000 /YR.				0.19
OPERATING SUPERVISION @ 20% SHIFT OPER.				0.04
SUBTOTAL				0.23
MAINTENANCE				
SECTIONS 100 - 1000 @ 4% B.L.				0.55
SECTIONS 1200 - 2200 @ 2% OFFSITES				0.13
SUBTOTAL				0.68
TOTAL DIRECT OPERATIONS				0.91
INDIRECTS				
PLANT OVERHEAD @ 60% OF TOTAL DIRECT OPER.				0.55
ANNUAL PAYMENT IN LIEU OF TAXES 2% TFI				0.48
DEPRECIATION OF PLANT 5% TFI				1.20
INTEREST ON LONG TERM DEBT 12% TFI				2.89
CORPORATE OVERHEAD @ 1% OF OEM				0.05
TOTAL INDIRECTS				5.17
TOTAL OPERATION AND MAINTENANCE				6.08
TOTAL PRODUCTION EX CAPITAL RECOVERY				8.02

TABLE 4

PLANT: TVA Coal Gasification Plant - Fuel Gas Product

CAPACITY: 5000 TPD Coal		TYPE INVESTMENT ESTIMATE:	Conceptual
LOCATION		INVESTMENT - B.L.	343
OPERATING DAYS/YR: 328.5		INVESTMENT -OFFSITES	147
		INTEREST DURING CONSTR.	59
FUEL GAS PRODUCT		WORKING CAPITAL	10
90 MMM BTU/SD		STARTUP AND TESTING	24
		TOTAL FIXED INVESTMENT	583 MM$
		LONG TERM DEBT	100%

ITEMS OF COST OPERATION OF PLANT	UNIT	UNIT PRICE	UNIT CONSUMPTION	COST PER UNIT PROD.
RAW MATERIALS				
COAL	Tons	$32.6	5000/D	1.81
TOTAL RAW MATERIALS				1.81
OPERATION AND MAINTENANCE				
UTILITIES				
ELECTRIC POWER	MWH	$28.0	480/D	0.15
RAW WATER	M Gal.	0.37	8000/D	0.03
CATALYST AND CHEMICALS	$1,200,000/Year			0.04
ASH DISPOSAL	$1,000,000/Year			0.04
TOTAL UTILITIES				0.26
DIRECT OPERATIONS				
SHIFT OPERATORS 38 OPERATORS/SHIFT @ $ 20,000 /YR.				0.11
OPERATING SUPERVISION @ 20% SHIFT OPER.				0.02
SUBTOTAL				0.13
MAINTENANCE				
SECTIONS 100 - 1000 @ 4 % B.L.I.				0.46
SECTIONS 1200 - 2200 @ 2 % OFFSITES				0.10
SUBTOTAL				0.56
TOTAL DIRECT OPERATIONS				0.69
INDIRECTS				
PLANT OVERHEAD @ 60% OF TOTAL DIRECT OPER.				0.41
ANNUAL PAYMENT IN LIEU OF TAXES 2% TFI				0.39
DEPRECIATION OF PLANT 5% TFI				0.98
INTEREST ON LONG TERM DEBT 12% TFI				2.37
CORPORATE OVERHEAD @ 1% OF OEM				0.05
TOTAL INDIRECTS				4.20
TOTAL OPERATION AND MAINTENANCE				5.15
TOTAL PRODUCTION EX CAPITAL RECOVERY				6.96

TABLE 5

PLANT: TVA Coal Gasification Plant - Fuel Gas Product

CAPACITY: 10,000 TPD Coal		TYPE INVESTMENT ESTIMATE:	Conceptual
LOCATION		INVESTMENT - B.L.	655
OPERATING DAYS/YR		INVESTMENT -OFFSITES	255
		INTEREST DURING CONSTR.	109
FUEL GAS PRODUCT		WORKING CAPITAL	18
180 MMM BTU/SD		STARTUP AND TESTING	45
		TOTAL FIXED INVESTMENT	1082 MM$
		LONG TERM DEBT	100%

ITEMS OF COST OPERATION OF PLANT	UNIT	UNIT PRICE	UNIT CONSUMPTION	COST PER UNIT PROD.
RAW MATERIALS				
COAL	Tons	$30.5	10,000/D	1.69
TOTAL RAW MATERIALS				1.69
OPERATION AND MAINTENANCE				
UTILITIES				
ELECTRIC POWER	MWH	$26.1	920/D	0.13
RAW WATER	M Gal.	$0.35	16,000/D	0.03
CATALYST AND CHEMICALS	$2,400,000/Year			0.04
ASH DISPOSAL	$2,000,000/Year			0.04
TOTAL UTILITIES				0.24
DIRECT OPERATIONS				
SHIFT OPERATORS 45 OPERATORS/SHIFT @ $ 20,000 /YR.				0.06
OPERATING SUPERVISION @ 20% SHIFT OPER.				0.02
SUBTOTAL				0.08
MAINTENANCE				
SECTIONS 100 - 1000 @ 4% B.L.I.				0.44
SECTIONS 1200 - 2200 @ 2% OFFSITES				0.09
SUBTOTAL				0.53
TOTAL DIRECT OPERATIONS				0.61
INDIRECTS				
PLANT OVERHEAD @ 60% OF TOTAL DIRECT OPER.				0.37
ANNUAL PAYMENT IN LIEU OF TAXES 2% TFI				0.37
DEPRECIATION OF PLANT 5% TFI				0.91
INTEREST ON LONG TERM DEBT 12% TFI				2.20
CORPORATE OVERHEAD @ 1% OF OEM				0.05
TOTAL INDIRECTS				3.90
TOTAL OPERATION AND MAINTENANCE				4.75
TOTAL PRODUCTION EX CAPITAL RECOVERY				6.44

TABLE 6

PLANT: TVA Coal Gasification Plant - Fuel Gas Product

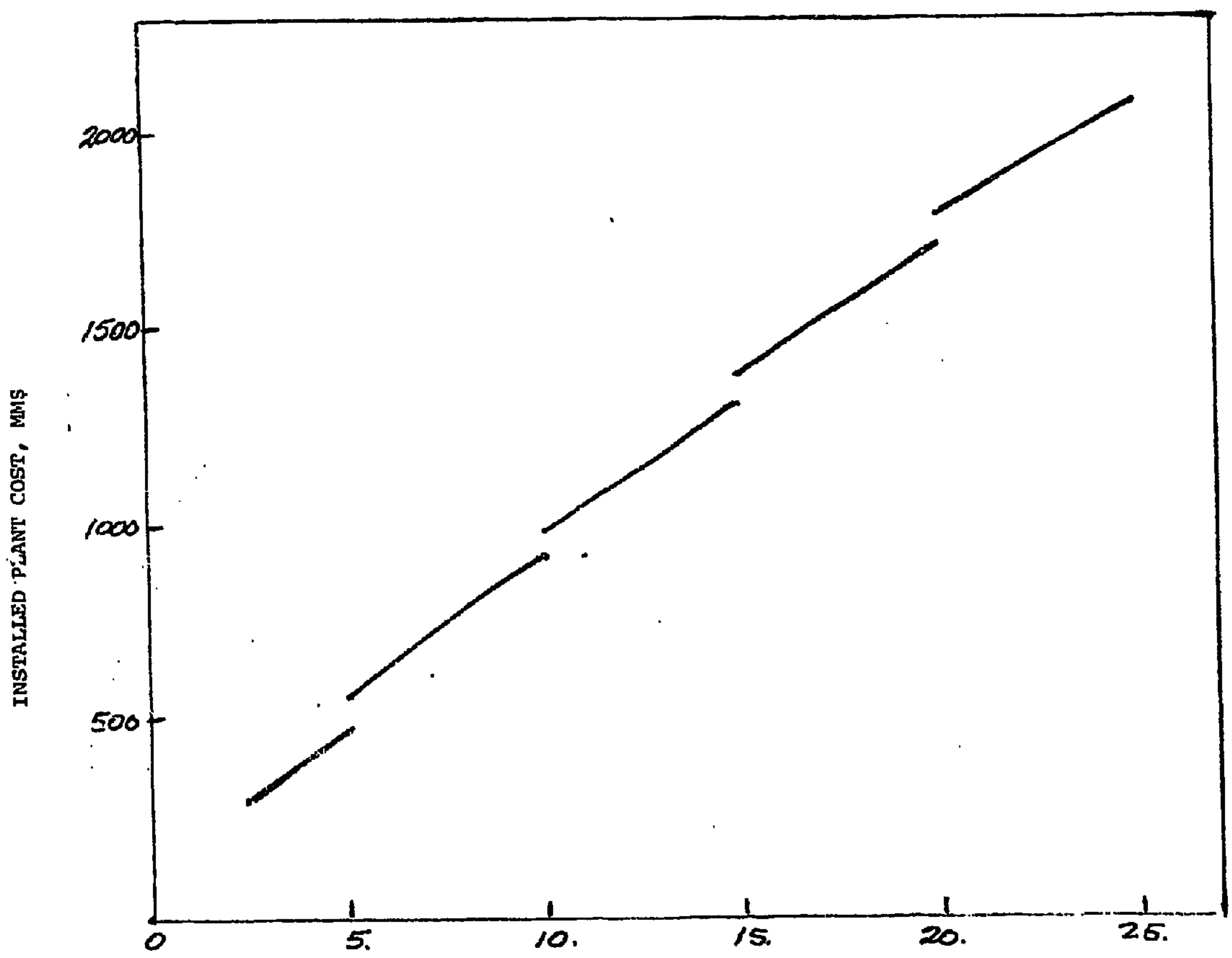
CAPACITY: 20,000 TPD Coal		TYPE INVESTMENT ESTIMATE:	Conceptual
LOCATION		INVESTMENT - B.L.	1323
OPERATING DAYS/YR	328.5	INVESTMENT -OFFSITES	46[illegible]
		INTEREST DURING CONSTR.	215
FUEL GAS PRODUCT		WORKING CAPITAL	34
360 MMM BTU/SD		STARTUP AND TESTING	90
		TOTAL FIXED INVESTMENT	2129 MM$
		LONG TERM DEBT	100%

ITEMS OF COST OPERATION OF PLANT	UNIT	UNIT PRICE	UNIT CONSUMPTION	COST PER UNIT PROD.
RAW MATERIALS				
COAL	Tons	$28.4	20,000/D	1.58
TOTAL RAW MATERIALS				1.58
OPERATION AND MAINTENANCE				
UTILITIES				
ELECTRIC POWER	MWH	$24.4	1840/D	0.12
RAW WATER	M Gal.	$0.32	32,000/D	0.03
CATALYST AND CHEMICALS	$4,800,000/Year			0.04
ASH DISPOSAL	$4,000,000/Year			0.03
TOTAL UTILITIES				0.22
DIRECT OPERATIONS				
SHIFT OPERATORS 54 OPERATORS/SHIFT @ $ 20,000 /YR.				0.04
OPERATING SUPERVISION @ 20% SHIFT OPER.				0.01
SUBTOTAL				0.05
MAINTENANCE				
SECTIONS 100 - 1000 @ 4 % B.L.I.				0.45
SECTIONS 1200 - 2200 @ 2 % OFFSITES				0.08
SUBTOTAL				0.53
TOTAL DIRECT OPERATIONS				0.58
INDIRECTS				
PLANT OVERHEAD @ 60% OF TOTAL DIRECT OPER.				0.35
ANNUAL PAYMENT IN LIEU OF TAXES 2% TFI				0.36
DEPRECIATION OF PLANT 5% TFI				0.90
INTEREST ON LONG TERM DEBT 12% TFI				2.16
CORPORATE OVERHEAD @ 1% OF OEM				0.05
TOTAL INDIRECTS				3.82
TOTAL OPERATION AND MAINTENANCE				4.62
TOTAL PRODUCTION EX CAPITAL RECOVERY				6.20

FIGURE 1

ESTIMATED VARIATION IN PLANT INVESTMENT

DEPENDENCE ON PLANT CAPACITY

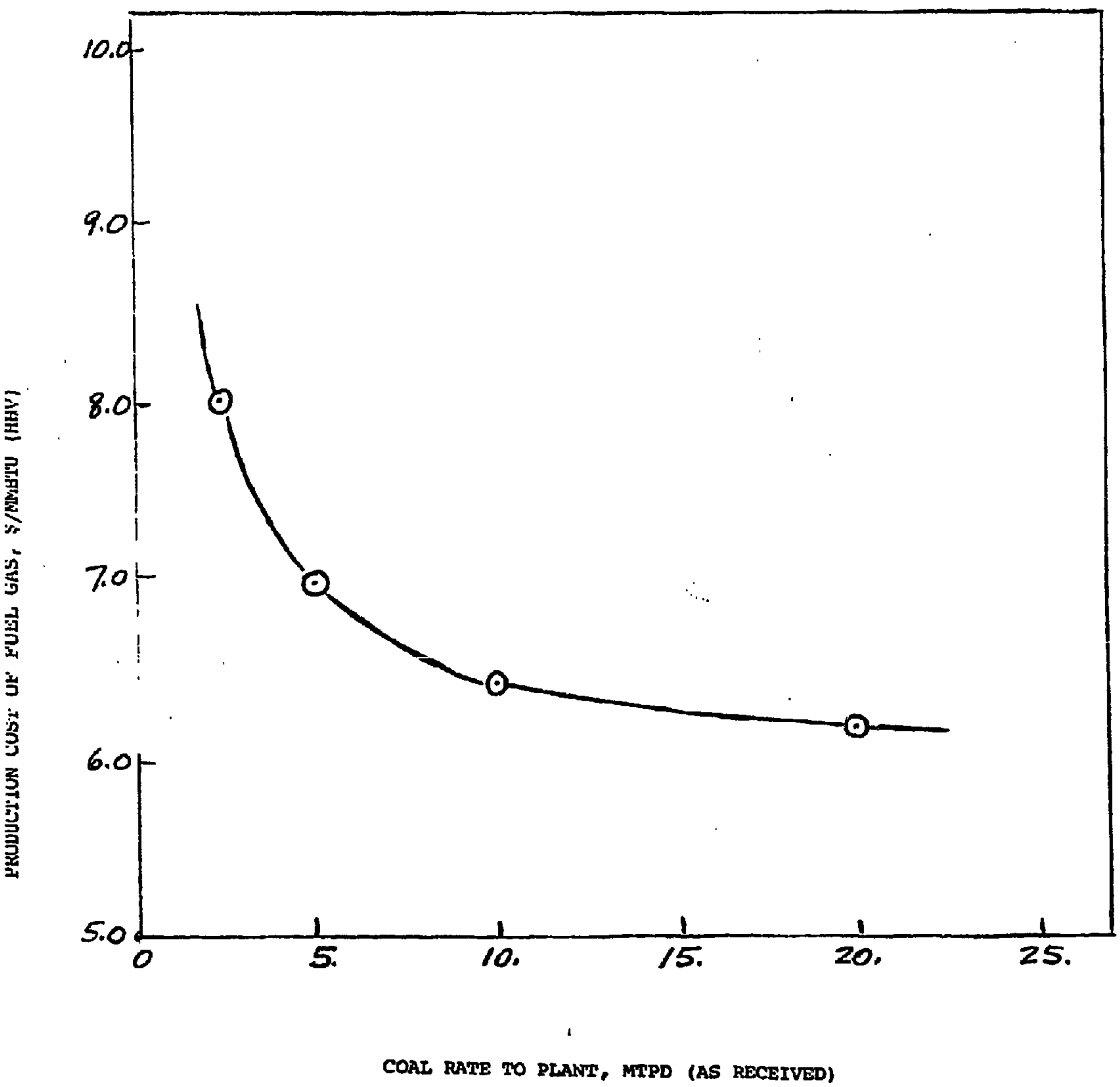


COAL RATE TO PLANT, MTPD (AS RECEIVED)

FIGURE 2

DEPENDENCE OF FUEL GAS PRODUCTION

COST ON PLANT CAPACITY



SECTION 3.9

LOAD CHANGES

ASSESSMENTS AND PROCESS SELECTIONS

LOAD CHANGES

The coal gasification plant being considered by TVA produces a medium BTU product gas which can be used as an industrial fuel gas. TVA is presently discussing use of the gas with potential customers but definite commitments to purchase gas have not yet been made. As a result, the demand pattern on a daily and seasonal basis is not known at this time.

Although gas demand will undoubtedly vary on both short and long term basis, TVA decided to base the conceptual designs of coal gasification plant on continuous operation at design capacity. The capabilities and limitation of the plants with respect to load changes would then be considered. Those considerations are the subject of this assessment.

General

Foster Wheeler carried out, as part of its work for TVA on this study, designs for five coal gasification plant producing medium BTU fuel gas at a rate of about 350 billion BTU per day. The plants use one of the following coal gasification processes:

- Lurgi Dry Ash
- Koppers Totzek
- B & W
- Texaco
- BGC/Lurgi Slagger

The plants consist of four operating modules, each module being essentially independent of the others except for certain spare systems which are shared among the four modules.

Production of fuel gas in these plants involves, in general, a sequence of processing steps which are directly involved with gas production:

- Air Separation
- Coal Gasification
- Acid Gas Removal
- Treated Gas Compression

There are other processing steps - sulfur recovery, sour water stripping, ash/slag handling in all plants, gas liquor separation, phenol recovery, ammonia recovery, in Lurgi Dry Ash and BGC/Lurgi slagger plants. These steps are, however, auxiliary to the main gas production sequence and are less severely affected by changes in plant output.

The main processing sequence involves gas production or gas treatment and there is no practical method of providing surge storage capacity of any significance for intermediate product from each processing step. As a result, changes in plant output are directly reflected in each processing step in the sequence. Each processing system in the sequence must respond to changes in load in concert with other systems in the sequence. If this cannot be accomplished, gas must be flared until conditions are satisfactory in all systems.

This situation in gas processing plants indicates, in general, that operation at a plant output which is essentially invariant is the preferable method of operation. Under these ideal circumstances, operation of each system can be lined out at optimum levels and product yield and quality can be maximized. This type of operation is also the most economical if the constant plant output corresponds to design capacity.

In practice, however, the circumstances of gas use in industrial plants result in variable gas demand on a daily, weekly and seasonal basis. The fuel gas producing plant will be required to follow these changes although the magnitude of the changes may be modified by the inherent storage capabilities of gas delivery pipeline and equipment. The transient response of the gas delivery pipeline to changes in gas input and output rates will be important with respect to short term variations.

Capabilities of Plants with Respect to Load Changes

Capabilities of the coal gasification plants with respect to load changes can be examined in terms of the individual processing steps in the gas production sequence. Table 1 lists for each of the five plants, the number of operating processing trains per module in air separation, gasification, raw gas compression, acid gas removal and clean gas compression sections. All of the plants contain multiple gasification trains - three in the case of Texaco gasifiers, up to eight in the case of Koppers-Totzek gasifiers. Assuming operability down to about 50% of design capacity for each train, raw gas output could be varied as follows:

Lurgi	7% to 100% in steps of 14%
K-T	6% to 100% in steps of 12.5%
B & W	12.5% to 100% in steps of 25%
Texaco	16% to 100% in steps of 33%
BGC/Lurgi	12.5% to 100% in steps of 25%

Other sections of the plant are less flexible, however, since the number of operating trains is less than that of the gasification section. In the case of the plant based on Lurgi dry ash gasifiers, the air separation, acid gas removal and clean gas compression sections are all single train. The plant based on BGC/Lurgi slagging gasifiers is similar in this respect.

Plants based on K-T, B & W, and Texaco gasifiers all have two trains of air separation, and are thus more flexible than plants based on Lurgi dry ash and BGC/Lurgi slagging gasifiers. Raw gas compression (for the K-T plant), acid gas removal, and clean gas compression (for K-T and B & W plants), however, are all single train. The limitation in these plants may be the compressors which generally experience surge problems at flows below about 70% of capacity. This limitation can be avoided by recycling gas to the compressor section.

Effects of Load Changes

Effects of load changes include loss of heat recovery efficiency and increased requirement for equipment. The loss in heat recovery efficiency occurs since heat losses tend to be independent of throughput. Additional equipment may be required to provide flexibility to melt a given pattern of load changes. Examples are coolers to recycle product gas to compressor suction, multiple trains, etc. Complete detailing of these effects requires definition of the load variation pattern.

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TABLE 1

TVA Coal Gasification Demonstration Plant Study
Task Assessment: Effect of Load Changes
Number of Operating Processing Trains per Modules

Process System	Lurgi Dry Ash	Koppers Totzek	B & W	Texaco	BGC/Lurgi Slagger
Air Separation	1	2	2	2	1
Gasification	7	8	4	3	4
Raw Gas Compression	-	1	-	-	-
Acid Gas Removal	1	1	1	1	1
Clean Gas Compression	1	1	1	-	1