FE-2371-36(Vol.3)
Distribution Category UC-90e

EVALUATION OF A PRESSURIZED-FLUIDIZED BED COMBUSTION (PFBC) COMBINED CYCLE POWER PLANT CONCEPTUAL DESIGN

Final Report

TRADE-OFF STUDIES

SUBTASK 1.3

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Date Published -June, 1978

PREPARED FOR THE UNITED STATES
DEPARTMENT OF ENERGY

Under Contract No. EX-76-C-01-2371

Abstract

In June 1976, the U. S. Department of Energy (DOE) awarded a contract to an industry team consisting of Burns and Roe Industrial Services Corp. (BRISC), United Technologies Corp. (UTC), and the Babcock & Wilcox Company (B&W) for an "Evaluation of a Pressurized, Fluidized Bed Combustion (PFBC) Combined Cycle Power Plant Design".

The results of this program indicate that pressurized fluidized bed combustion systems, operating in a combined cycle power plant, offer great potential for producing electrical energy from high sulfur coal within environmental constraints, at a cost less than conventional power plants utilizing low sulfur coal or flue gas desulfurization (FGD) equipment, and at higher efficiency than conventional power plants.

As a result of various trade-off studies, a 600 MWe combined cycle arrangement incorporating a PFB combustor and supplementary firing of the gas turbine exhaust in an atmospheric fluidized bed (AFB) steam generator has been selected for detailed evaluation.

The overall program consists of the following Subtasks:

1.1 - Commercial Plant Requirements Definition

1.2 - Commercial Plant Design Definition

1.3 - System Analysis and Trade-Off Studies

1.4 - Reliability and Maintainability Evaluation with Advanced Technology Assessment

1.5 - Environmental Analysis

1.6 - Economic Analysis

1.7 - Evaluation of Alternate Plant Approaches

1.8 - PFB/Gas Turbine/Waste Heat Boiler Cycle Study

1.9 - PFB/Gas Turbine/Power Turbine Reheat Cycle Study

This Interim Report discusses the results of studies performed under Subtask 1.3.

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1.0 SUMMARY

Efforts under subtask 1.3 have included a study of the sensitivity of plant performance and cost to various design variables. Variations in pressure and temperature of the gas and of the steam, as well as various cycle configurations have been studied. Alternatives have also been evaluated in the areas of PFB combustor design (i.e., shop assembled versus field assembled combustors), high temperature/high pressure particulate removal, and sorbent regeneration. These efforts have served to optimize the commercial plant design (Refer to Subtask 1.2. Report) and to provide guidance for further development efforts.

1.1 ALTERNATIVE COMBINED CYCLE ARRANGEMENTS

Several power plant configurations have been studied using different schemes to improve either the gas turbine or steam system performance. All of the system configurations considered utilize gas turbines, a steam bottoming system, and pressurized fluidized bed combustors. The systems are described below and are shown schematically in Figures A-3 through A-6.

Case	Description
A	COmbined Gas And Steam (COGAS), Fig. A-3
B	COGAS + Exhaust-fired air-cooled AFB, Fig. A-4
C	COGAS + Exhaust-fired steam-cooled AFB, Fig. A-5
D	COGAS + Power Turbine Reheat, Fig. A-6

In the COGAS case (Case A), a PFB combustor replaces the conventional gas turbine combustor, and the exhaust from the gas turbine is used in an unfired waste heat recovery boiler (WHRB) for the generation of steam. Cases B through D utilize additional stages of combustion to increase the efficiency and output of each gas turbine and/or its associated portion of the steam plant.

On the basis of a performance and economic screening analysis Case C was selected for further study in the commercial plant conceptual design effort (i.e., Subtask 1.2).

The selected PFB power plant configuration (Case C) is an excellent compromise between complexity, efficiency, and cost. Other configurations that have been studied have shown either better overall efficiency or less complexity than the selected case and may be attractive alternatives.

The configuration using reheat before the power turbine (Case D) has the highest overall efficiency and appears equivalent in overall cost to the selected case. It was passed over as the preferred configuration primarily on the bases of complexity and higher technical risk. However, since the costing analysis was performed on an order of magnitude basis, a more detailed analysis of the reheat cycle is warranted and has been authorized by D.O.E. as an extension to the current contract. These additional studies will be performed under Subtask 1.9.

The COGAS PFB power plant (Case A) utilizes only one stage of combustion which results in less output per gas turbine module than the other configurations studied. This power plant may be attractive as a smaller interim system to demonstrate technology and bridge the gap between present power plants and more efficient PFB/AFB cycles. Furthermore, Case A may be more competitive in the smaller plant sizes (say, less than 100 MWe). Since the maximum generating capacity of each gas turbine and associated waste heat steam system is only 90-100MWe, based on the largest gas turbines available to date, multiple PFB/G.T./WHRB modules must be used for plants over 100 MWe. Therefore, reductions in specific cost ($/kW) associated with the "economy of scale" are expected to be relatively small for Case A plants over 100 MWe in capacity. However, in the small utility or industrial market, only one module would be required. Hence, Case A should be more competitive with other alternatives which utilize a larger proportion of steam turbine to gas turbine power since the steam power equipment must be reduced to the same scale as the gas turbine equipment. Thus, a more detailed analysis of Case A is also justified.

On the basis of the studies described above, the Department of Energy (DOE) has authorized an extension to the present contract to cover additional studies on Case A and Case D. These studies will be performed as Subtasks 1.8 and 1.9, respectively, under the extended contract.

1.2 SENSITIVITY ANALYSES ON DESIGN VARIABLES

Sensitivity studies have indicated that the power output and efficiency of the selected PFB/AFB combined cycle are relatively insensitive to changes in gas and steam pressure drops. A 100% increase in these pressure drops above the base design value would only reduce plant output and efficiency by 1-2.7%. Plant efficiency would decrease by 1-2% for a 100°F decrease in gas turbine inlet temperature (T.I.T.) or a 100% increase in overall pressure ratio (OPR). While efficiency shows little sensitivity to these parameters, the same changes in T.I.T. and OPR would cause decreases in plant output of 9.4% and 23%, respectively. However, one of the unique features of the PFB/AFB cycle is that, subject to limits imposed by the oxygen content of the turbine exhaust gases, only minor changes to the bottoming cycle need be made so that additional fuel could be burned in the AFB in order to restore the lost output. Since the PFB/AFB combined cycle plant performance is not greatly affected, a significant reduction in the cost of electricity (COE) and an increase in plant availability might be obtained if the equipment were to be designed for the less severe pressure drop and temperature criteria. Parametric studies on capital cost versus these criteria would be required to evaluate the potential savings.

1.3 FABRICATION METHODS FOR PFB COMBUSTORS

The Subtask 1.2 conceptual plant design utilizes two PFB combustor systems per gas turbine. Each PFB system consists of one coal feed system, one dolomite feed system, one PFB combustor, one PFB ash letdown system and one dust cleanup system, which includes two Aerodyne 22000 SV-FBC Dust Collectors in series. Since the size of the combustor vessel prohibits rail shipment of fully assembled vessels, three alternate methods of fabricating, assembling and shipping the combustor vessel have been investigated. One method involves fabricating and assembling the PFB combustor vessel at Babcock & Wilcox's Mt. Vernon shops and shipping the complete vessel by barge to the jobsite. All necessary non-destructive testing of the vessel is done in the Mt. Vernon shops.

The second method consists of fabricating segments of the combustor vessel at Babcock & Wilcox's Barberton shops, shipping the segments by rail to the jobsite and assembling the segments in the field. Due to the size of the vessel, complete head and shell assemblies cannot be shipped by rail. Consequently, fairly elaborate field fabricating facilities are required for final machining and assembling of the vessel segments. Additional field facilities must be supplied for welding, stress relieving and non-destructive testing of the combustor vessel.

The third alternative is to use a larger number of smaller PFB combustors. Being able to ship an entire PFB combustor vessel by rail eliminates the need for costly barge unloading facilities and elaborate field fabrication facilities. Consequently, a smaller size PFB combustor has been designed in order to make use of the advantages of shipping an entire combustor vessel by rail. Drawings showing the arrangement and details of the smaller PFB combustor have been prepared. Five rail shippable combustors are equivalent to one base PFB combustor. Each rail shippable combustor is provided with a corresponding particulate collection system.

The total fabrication, transportation and erection costs for all three alternatives are as follows:

Alternative 1	-	$33,820,000
Alternative 2	-	34,140,000
Alternative 3	-	60,755,000

These costs include the combustors and corresponding solids feed systems, ash letdown systems, and particulate removal systems. It is clear that the large number of shop fabricated, rail shippable modules make this alternative uneconomical. The difference between alternatives 1 and 2 is not statistically significant. Alternative 2 has been selected for the commercial plant conceptual design since it would not require the construction of expensive barge unloading facilities. In addition, a specific sight may not have access to a navigable waterway. In any case, a thorough and more detailed cost estimate must be performed in order to justify the choice of either Alternative 1 or 2 over the other on the basis of economics.

1.4 PFB PARTICULATE REMOVAL SYSTEMS

Various particulate removal systems (conventional cyclones, high efficiency cyclones, and granular bed filters) have been investigated as part of the Trade Off Studies performed under Subtask 1.3. The results of this investigation indicate that conventional cyclones are incapable of removing all particulates greater than 10 micron diameter and are relatively inefficient in the range of 2 to 10 microns. Consequently, these devices are unable to meet the performance requirements for particulate removal equipment.

The high efficiency cyclones that were investigated are Aerodyne Development Corporation's "SV-FBC" Series of Dust Collectors and Donaldson Company's "Tan-Jet" System. The particular "SV-FBC" Dust Collector that has been studied is the Model 22000SV as shown in Figure C-3. Its design is an extension of the Aerodyne equipment presently used in low temperature, low pressure applications. The major modification consists of placing the collector inside a refractory lined pressure vessel which becomes an initial stage cyclone dust collector. This results in a single unit having two stages of collection. Calculations indicate that two of these units operating in series would be required for each PFB combustor. The dust loading entering the gas turbine in the critical 2 to 10 micron size range is projected to be 1/3 of the allowable level.

The "Tan-Jet" system of Donaldson is essentially a high efficiency multiclone which employs a secondary stream of clean compressed air from a booster compressor to generate a strong vortex in each of the multiclone tubes. Literature supplied by Donaldson indicates that the "Tan-Jet" system is capable of removing all particulates greater than 10 microns in diameter. However, the predicted collection efficiency of the "Tan-Jet" system in the range of less than 4 microns is lower than that of the Aerodyne Model 22000SV Dust Collector. Furthermore, the price of the Donaldson "Tan-Jet" system per PFB is significantly higher than that of the Aerodyne system.

Granular bed filters have been considered as an alternate method of removing particulates from the PFB gas stream. A fixed granular bed system design has been developed by the Ducon Company under subcontract to Babcock & Wilcox (B&W purchase order 717363DU). The study report by Ducon is contained in Appendix 8.1.

A conventional cyclone dust collector is located upstream of the granular bed filter in order to reduce the dust loading to the filter. This reduces the granular bed cleaning requirements and the overall system pressure drop. Ducon predicts that no particles greater than 2 microns will enter the gas turbine. The predicted collection performance of the Ducon system is greatly superior to that of the Aerodyne system for particle diameters of less than 10 microns. However, the price of the Ducon system per PFB is $14,930,000, compared to $7,310,000 for the Aerodyne system. Furthermore, the anticipated operating availability of the Ducon system is less than that of the Aerodyne system. For example, the Aerodyne system has no moving parts and is continuously self cleaning, whereas the Ducon system makes use of auxiliary equipment such as compressors and valves to periodically backwash (clean) individual filter elements. Consequently, Aerodyne's Model 22000SV Dust Collector has been selected for use in the conceptual plant design.

It should be noted that, before a PFB plant can become a commercial reality, a considerable amount of development effort must be expended in the following areas:

1. Determination of the quantity, characteristics and composition of the particulates leaving the PFB combustors.

2. Determination of the maximum quantity of PFB particulates that can be tolerated by the gas turbine as a function of size distribution, composition, and other characteristics.

3. Verification of the predicted performance, reliability, and costs for the equipment being proposed to remove PFB particulates from the gas stream.

1.5 EVALUATION OF REGENERATION OF SULFATED SORBENTS FROM FLUID BEDS

An engineering study of the regeneration of sulfated additives from a 600 MWe coal-fired fluidized bed power plant has been performed.

The work has involved a review of the literature, selection of a viable process to be used, preparation of conceptual flow diagrams, identification of required equipment and order-of-magnitude cost estimates for various sulfated sorbent regeneration and handling systems.

Several alternative arrangements of the one-step regeneration process have been studied and compared to a once-through sorbent system. Two involve use of a Claus sulfur recovery plant, and the third uses Foster Wheeler's RESOX system.

The results of this study indicate the following:

1. From a technical viewpoint, sorbent regeneration appears feasible. However, more experimental data is required if a commercial plant is to be designed with significant confidence levels.

2. Sorbent regeneration utilizing sulfur recovery processes with commercial operating experience, such as the Claus system, cannot be economically justified unless sorbent costs approach $30 per ton.

3. Additional development efforts are required in order to achieve an economical additive regeneration system. These efforts must be focused on the development of an economical sulfur recovery system, such as RESOX, as well as on the regenerator itself. Development of one without the other will be of no use economically.

4. If the currently projected costs of a RESOX system prove realistic, sorbent regeneration utilizing this system for sulfur recovery may be more economical than a once-through sorbent system based on a sorbent cost of over $7 per ton. However, to our knowledge, there is no RESOX system in commercial operation today.

5. From an environmental viewpoint, the amount of solid wastes leaving a plant with regeneration is only 35-40% of the amount produced in a "once-through" sorbent system. Therefore, the environmental impact of the waste disposal is greatly reduced.

6. It is recommended that further development work be performed on the regeneration of spent additive using the one-step regeneration process at 2000°F (1093°C) and atmospheric pressure. In conjunction with this work, development effort should be expended on economical sulfur recovery system such as Foster Wheeler's RESOX system or an equivalent.

7. For the Claus based system, land requirements are approximately 40,000 ft^2 for the entire regeneration plant if hydrogen sulfide is purchased "over-the-fence" and 60,000 ft^2 if it is manufactured in plant. For the RESOX based plant, approximately 50,000 ft^2 are required.

8. The number of personnel required for the various spent sorbent processing systems are as follows:

"Once-through" system	- 11
Regeneration with Claus (Purchased H_2S)	- 33
Regeneration with Claus (In-plant Manufacture of H_2S)	- 50
Regeneration with RESOX	- 31

With reasonable training, all personnel listed above are interchangeable with power plant personnel.

9. Without taking credit for sulfur sales, the operating costs for the regeneration systems range from 2.9-11 mills/kWh depending upon the sulfur recovery system selected.

1.6 ADDITIONAL GUIDES FOR FURTHER DEVELOPMENT

During the course of the Commercial Plant design (Subtask 1.2) and trade-off studies (Subtask 1.3), various cost improvement ideas have been conceived that could not be developed to any extent within the cost and schedule constraints of this program. At the risk that some obvious fallacies have been overlooked due to insufficient analysis, these ideas are mentioned in this section in the hope that they can be evaluated in some future effort and can perhaps contribute to an improvement in the technology.

One such idea involves a modification to the base commercial plant concept (per Subtask 1.2) to incorporate a baghouse as a replacement for the hot electrostatic precipitator. In addition to potential cost savings, this modification would also eliminate uncertainties involved with the use of an electrostatic precipitator on FBC exhaust gases.

The second idea discussed concerns the potential for improving plant efficiency and costs by taking advantage of the lower acid dew point temperatures expected for FBC flue gases. A reduction in stack temperature will result in some improvement in efficiency (i.e., fuel cost) and reduction of costs for I.D. fans, stacks, etc. These cost improvements must be compared to increases in costs for the turbine and the plant heat rejection system. All of these factors must be evaluated in order to determine whether a reduction in stack temperature is cost effective regardless of any relaxation in dew point criteria.

A third idea involves the use of a double-ended generator drive in lieu of the single-ended drive used in the base concept. In the case of the selected cycle, doubling the size would allow economical use of a hydrogen cooled generator in lieu of air-cooled, with a reduction in specific cost and a slight increase in overall efficiency. Other advantages also result. These must be weighed against potential increased costs associated with plant layout.

One final idea involves the use of very high gas turbine pressure ratios, say, 40 to 100. The higher pressure would be obtained by cooling the current compressor discharge air and supplying the cooled air to a high pressure compressor which discharges to the high pressure PFBC. In passing through the turbine, the gas would be reheated one or more times. The system would be more complex than the selected cycles but offers the possibility of reduced PFB and cleanup equipment sizing, improved efficiency and specific work, and lower cost.

2.0 SELECTION OF PLANT DESIGN PARAMETERS

2.1 PLANT SIZE

A utility system, normally, decides the size of a base loaded generating unit depending on its own present generating capacity, projected growth of electricity demand, the power grid system to which it is connected and overall economics. Base loaded generating units come in all sizes from under 100 MWe to over 1300 MWe. This can be seen from the following table (Ref. 1):

NUMBER AND AGGREGATE CAPACITY OF NEW
THERMAL POWER PLANT UNITS GROUPED
ACCORDING TO SIZE RANGE AS OF APRIL 1, 1976

In Terms of Manufacturers' Ratings of the Units

Size Range, kW	Number of Units	Aggregate Capacity, kW	Percent of Total Units	Percent of Total kW
1,300,000 and Larger	28	37,313,356	5.9%	12.8
1,200,000 to 1,299,999	33	40,748,473	6.9%	14.0
1,100,000 to 1,199,999	40	46,601,829	8.4%	16.0
1,000,000 to 1,099,999	8	8,469,119	1.7%	2.9
900,000 to 999,999	22	20,820,265	4.6%	7.1
800,000 to 899,999	40	33,370,518	8.4%	11.4
700,000 to 799,999	21	15,546,612	4.4%	5.3
600,000 to 699,999	37	23,863,366	7.8%	8.2
500,000 to 599,999	57	30,150,204	12.1%	10.3
400,000 to 499,999	33	14,174,473	6.9%	4.9
300,000 to 399,999	20	7,082,892	4.2%	2.4
200,000 to 299,999	22	5,461,312	4.6%	1.9
100,000 to 199,999	23	3,392,507	4.8%	1.2
Under 100,000	91	4,812,435	19.3%	1.6
TOTAL THERMAL UNITS	475	291,807,361	100.0%	100.0

From operating experiences of utility plants it has been found that overall efficiency and cost of electric generation improve as unit size increases. It is obvious that only a very large utility system would opt for a unit capacity of over 1000 MWe, whereas, units under 100 MWe would be preferred for peaking service or for use in a small municipality. In the mid-ranges there is a relatively large demand for units sized between 500 to 599 MWe, as is apparent from the above Table.

Operating availability and on-line experience are also critical factors in deciding the size of a generating unit. The following data were collected by Edison Electric Institute for the operating availability of fossil fired units (Ref. 2).

Unit Size (MWe)	Operating Availability %
300-399	81.2
400-599	78.3
600-799	74.3
800 and above	74.4

This Table shows that a unit of 400 to 599 MWe capacity has a good operating availability.

The utilities also have considerable on-line experience with units of 400-599 MWe capacity as is evident from the following Table (Ref. 2) which shows units that are currently in operation:

Unit Size(MWe)	Number of Units	Number of Unit Years
300-399	117	703
400-599	125	604
600-799	69	266
800 and above	33	111

For this study, the selection of the size of a generating system utilizing pressurized fluidized combustion was also influenced by available sizes of gas turbines, and the anticipated need to scale the capacity of the plant over the full range up to large commercial size (i.e., to 1300 MWe and larger).

A nominal size of 600 MWe has been selected for this conceptual study because; a) there is an ample demand for this size; b) utilities have good on-line experience with these sized units, and; c) operating availability has been good for this size.

2.2 CYCLE CONFIRGURATION AND PARAMETRIC STUDIES

2.2.1 Summary

The objective of this effort is to select a Pressurized Fluidized Bed (PFB) power plant configuration that utilizes coal and produces approximately 600 MWe of electric power in an efficient and economic manner. To accomplish this task, analyses of the gas turbine cycle and integration of several PFB gas turbine steam system configurations have been performed. Also, preliminary economic evaluations of the various alternative systems have been used to identify the configuration that is likely to be the most cost effective. The results of these studies are presented below:

1. As a result of the economic and performance sensitivity studies, the selected power plant configuration incorporates an air-cooled PFB combustor for the gas turbine with supplementary firing of the turbine exhaust in a steam cooled AFB to provide heat for the 2400 psig/1000 F/1000F steam bottoming system. The estimated coal-pile-to-busbar efficiency is 40.9 percent (Gross,HHV).

2. The relatively low operating temperature (1650F) of the air-cooled PFB coal combustor results in low open-cycle gas turbine cycle efficiency. Also, the gas turbine exhaust temperature is low so that an exhaust-fired AFB is necessary to provide enough high-level energy to generate steam for a high-efficiency steam cycle.

3. An overall pressure ratio (OPR) of 10 provides a good combination of gas turbine efficiency and power output. Also, when integrated with the remaining subsystems of the selected configuration, this pressure ratio provided the best overall system efficiency.

4. The performance studies indicate that a configuration using gas reheat before the power turbine exhibits the highest system efficiency (43.1 percent) of all the configurations investigated whereas the simple, unfired heat recovery combined cycle configuration shows the lowest system efficiency (38.4 percent) versus the selected cycle at 40.9 percent.

5. Since the preliminary economic evaluations of the different system configurations were of a rough order of magnitude, systems other than that configuration selected are worthy of additional study. In particular, the cycle with reheat before the power turbine exhibits the best efficiency and has an estimated cost close enough to the selected system to warrant further attention.

   The system with the unfired waste heat recovery boiler might be attractive to the utilities as an initial PFB system. It would have a higher efficiency than a conventional coal-fired steam power plant with stack gas cleanup. Therefore, further evaluation of this cycle is also warranted.

2.2.2 Introduction

Pressurized Fluid Bed (PFB) combustors are constrained to operate at temperatures of $1600^{\circ}F$ to $1700^{\circ}F$. This temperature range is low enough to limit the vaporization of the alkali metals from the fluidized bed, the deposition of particulates in downstream equipment, and corrosion of turbine blades. Also, this temperature will assure an operating margin below the coal ash softening temperature to prevent agglomeration within the bed. Unfortunately, the operation at these bed temperatures will allow turbine inlet temperature of only $1550^{\circ}F$ to $1650^{\circ}F$, values obtained over a decade ago in commercial units.

When an air-cooled PFB is used as a combustor for a gas turbine, the resulting open cycle gas turbine efficiency is relatively low as a result of the low turbine inlet temperature. Also, the exhaust temperature is not high enough to allow a high-efficiency steam bottoming cycle to be used. An unfired heat recovery combined cycle using an air-cooled PFB would have an overall efficiency only a few percentage points better than conventional coal-fired power plants with flue gas desulfurization. To make PFB power plant performance more attractive to the utilities, cycle modifications can be employed to increase efficiency.

The objective, then, of this task is to identify the best configuration of gas turbines, pressurized fluid bed combustion, heat recovery equipment, and steam bottoming cycle for a 600 MWe power plant.

It should be recognized that all performance estimates utilized during this trade-off study were made during the early stages of the program. At the time, design efforts on Subtasks 1.2, 1.8, and 1.9 had not yet been initiated. Therefore, certain assumptions had to be made regarding cycle pressure losses, steam cycle configuration and throttle conditions, fuel HHV/LHV ratios, etc. For this reason, the plant performance estimates presented in this section differ from the final values reported under the other subtasks for nominally similar cycles which are based on a more detailed analysis of more highly optimized configurations.

2.2.3 Approach

Several alternate power plant configurations have been evaluated that provide relatively good efficiency. Some of the candidate configurations increase gas turbine exhaust temperature by either supplementary firing of the turbine exhaust or by reheating the turbine flow between stages. Reheat leads to increases in turbine output as well as exhaust temperature allowing the use of a high-efficiency steam bottoming cycle to increase the overall system performance. For comparison purposes, a conventional combined cycle without reheat or supplementary exhaust firing has also been evaluated.

In addition to the screening of different configurations, variations in the gas turbine operating parameters have been analyzed to establish the conditions giving the best combination of efficiency and power output for the overall system. The range of parameters evaluated are within the present commercial gas turbine technology level. By focusing on this technology level it is possible to identify a low-risk design for the gas turbine subsystem.

In addition to defining the performance of alternate systems, a relative economic evaluation of the same power plants has been made. The estimates of relative capital costs are of a rough order of magnitude since certain pieces of equipment have not been included in the total cost estimates. The efficiency differences between the various systems have been evaluated on a capitalized cost basis and a relative cost determined. The system with the lowest net relative cost has been selected as the PFB power plant configuration to be studied under Subtask 1.2 (Commercial Plant Design Definition).

Sensitivity studies have been carried out on the selected system to further identify the operating parameters resulting in the most attractive performance.

2.2.4 <u>Performance Evaluation</u>

All cycle performance calculations have been made with United Technologies Corporation's (UTC) <u>S</u>tate-<u>o</u>f-the-<u>A</u>rt <u>P</u>erformance <u>P</u>rogram (SOAPP).

2.2.4.1 Gas Turbine Performance

The initial studies performed for the PFB power plant defined the performance of a gas turbine that operates at a turbine inlet temperature of approximately $1500^{\circ}F$ - $1700^{\circ}F$.

In order to utilize a large gas turbine and yet maintain realism, a revised version of UTC's FT50 gas turbine has been used. The FT50 industrial gas turbine is an axial flow machine with a twin-spool gas generator and free turbine on a third spool. The initial design engine airflow of 815 lb/sec has been retained, and the turbine has been rematched to handle this airflow at various turbine inlet temperatures and pressure ratios. Figure A-1 presents the performance of this machine at the different operating conditions. The turbine inlet temperature range of $1500^{\circ}F$ to $1700^{\circ}F$ is representative of that resulting from the split stream air-cooled PFB combustor. The pressure range varies from that of current industrial turbines (8:1) to that for the FT50A-4 (18:1).

Gas turbine exhaust temperatures in some cycles are not high enough to produce steam for a high performance bottoming cycle. One method for increasing the exhaust temperature of a gas turbine is to use reheat before the power turbine. A comparison of the exhaust temperatures for the cases with and without reheat is shown in Fig. A-2. The heat available to the steam cycle can also be increased by supplementary firing of the gas turbine exhaust in a coal fired Atmospheric Fluid Bed (AFB).

2.2.4.2 Performance of Alternative Combined Cycle Power Plants

2.2.4.2.1 Alternative Combined Cycle Plant Descriptions

Several power plant configurations have been studied using different schemes to improve either the gas turbine or steam system performance. All of the system configurations considered utilize gas turbines, a steam bottoming system, and pressurized fluidized bed combustors. The systems are described below and are shown schematically in Figures A-3 through A-6:

SIMPLE–CYCLE GAS TURBINE WITH PFB COMBUSTOR

Wa (GT) = 815 LBS/SEC

T_{AMB}=59° F

3.5" H_2O INLET PRESSURE LOSS

5.0" H_2O EXHAUST PRESSURE LOSS

10% COMBUSTOR PRESSURE LOSS

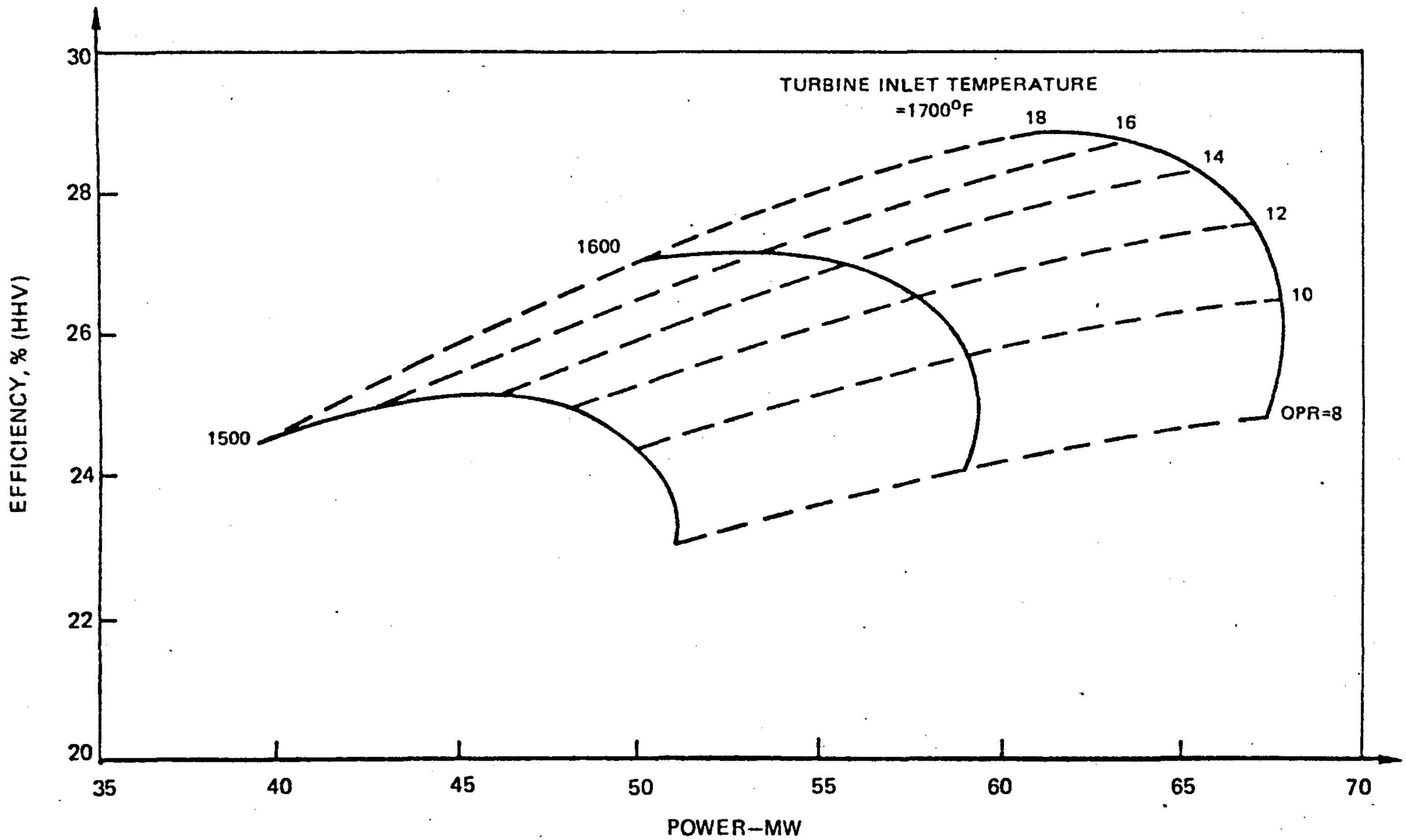


FIGURE: A-1

GAS TURBINE EXHAUST TEMPERATURE

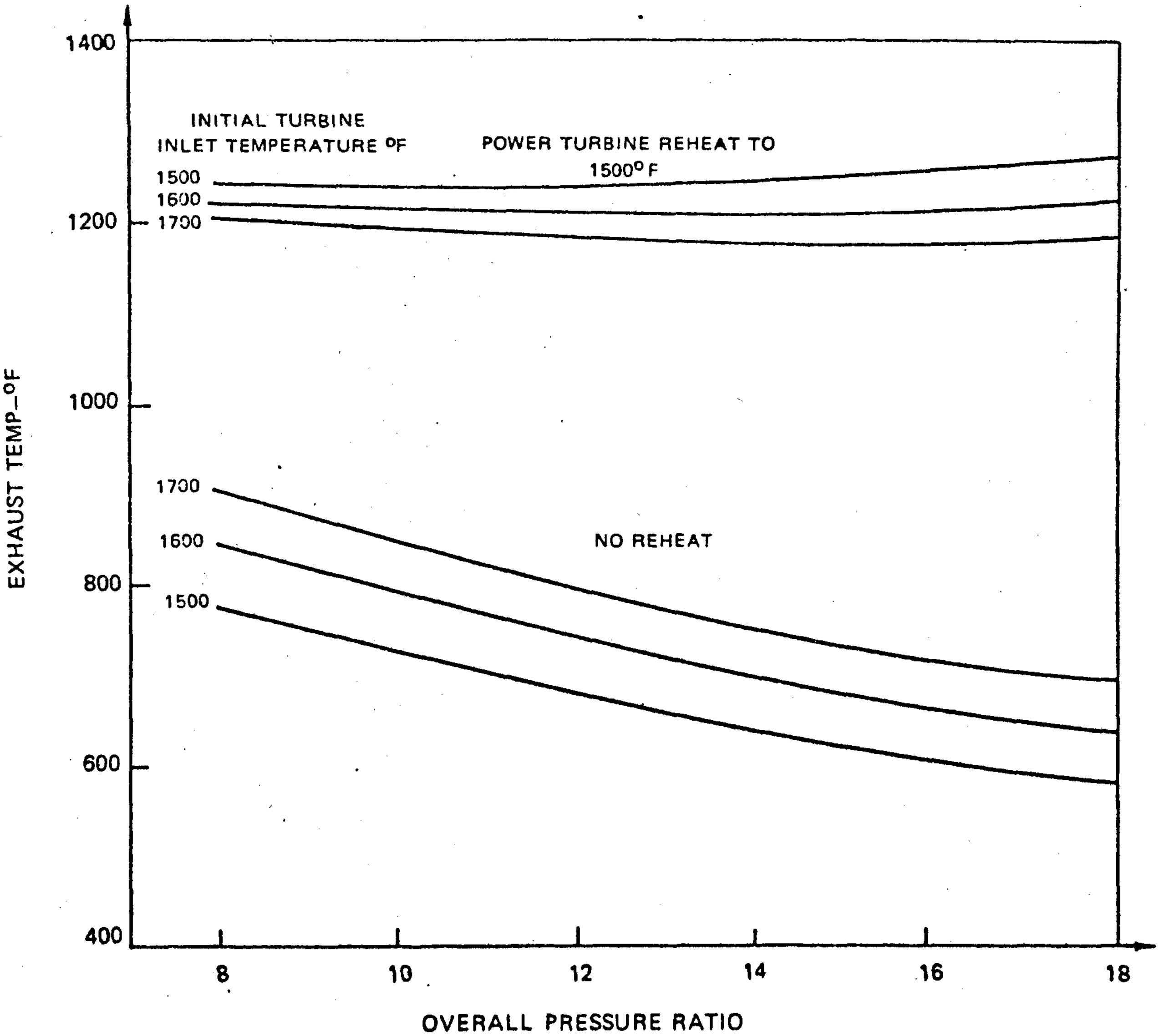


FIGURE: A-2

Case	Description
A	COmbined Gas And Steam (COGAS), Fig. A-3
B	COGAS + Exhaust-fired air-cooled AFB, Fig. A-4
C	COGAS + Exhaust-fired steam cooled AFB, Fig. A-5
D	COGAS + Power Turbine Reheat, Fig. A-6

In the COGAS case (Case A), a PFB combustor replaces the conventional gas turbine combustor, and the exhaust from the gas turbine is used in an unfired waste heat recovery boiler for the generation of steam. Cases B through D utilize additional stages of combustion to increase the efficiency and output of each gas turbine and/or its associated portion of the steam cycle.

Cases B and C utilize the gas turbine exhaust to burn additional coal thereby increasing the heat available to the steam system. Since the gas turbine combustion process utilizes less than one-quarter of the available oxygen in the air, the turbine exhaust gas can support considerable firing of additional coal. An atmospheric fluid bed (AFB) combustor has been considered for exhaust-firing because it provides a means for capturing the sulfur released in the coal combustion process. Case B utilizes an air-cooled AFB where the exhaust stream from the gas turbine is split and part of the air is used for combustion in the bed. The remaining gas cools the bed by passing through tubes immersed in the bed to absorb heat. After mixing the two streams back together, the gas is used to generate steam in a waste heat recovery system.

Case C is similar to Case B except that the AFB is steam-cooled. The gas turbine exhaust flow is also split with a portion of the gas being used for combustion. This stream is also used for reheating steam and is then recombined with the remaining exhaust gas and passed over a high-pressure economizer. The AFB is cooled by having high-pressure evaporator and superheater tubes immersed in the bed to absorb heat and generate steam.

Reheating before the power turbine, Case D, increases the complexity of the gas turbine and requires two pressurized coal combustion stages and associated particulate cleanup, solids feed, and spent bed material removal systems. The gases exiting from the compressor drive turbines are split, similar to the scheme in the main PFB combustor, and are used for combustion and cooling the bed. The exhaust gases from the exit of the power turbine are then used to generate steam in a high-performance steam bottoming cycle.

SIMPLE COGAS CONFIGURATION

(CASE A)

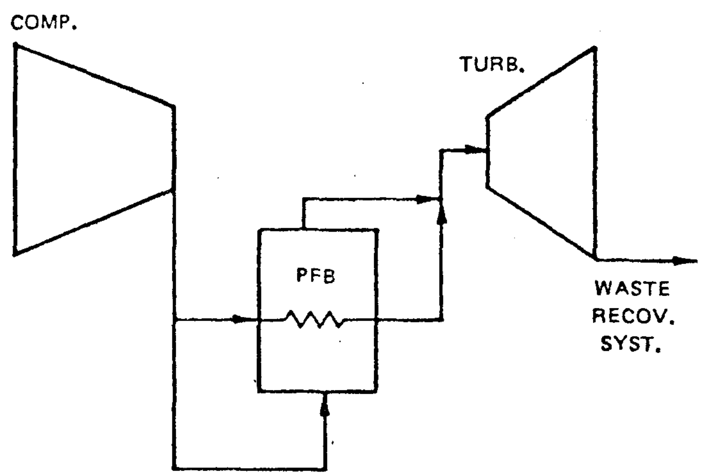


FIGURE: A-3

COGAS WITH EXHAUST–FIRED AIR–COOLED AFB

(CASE B)

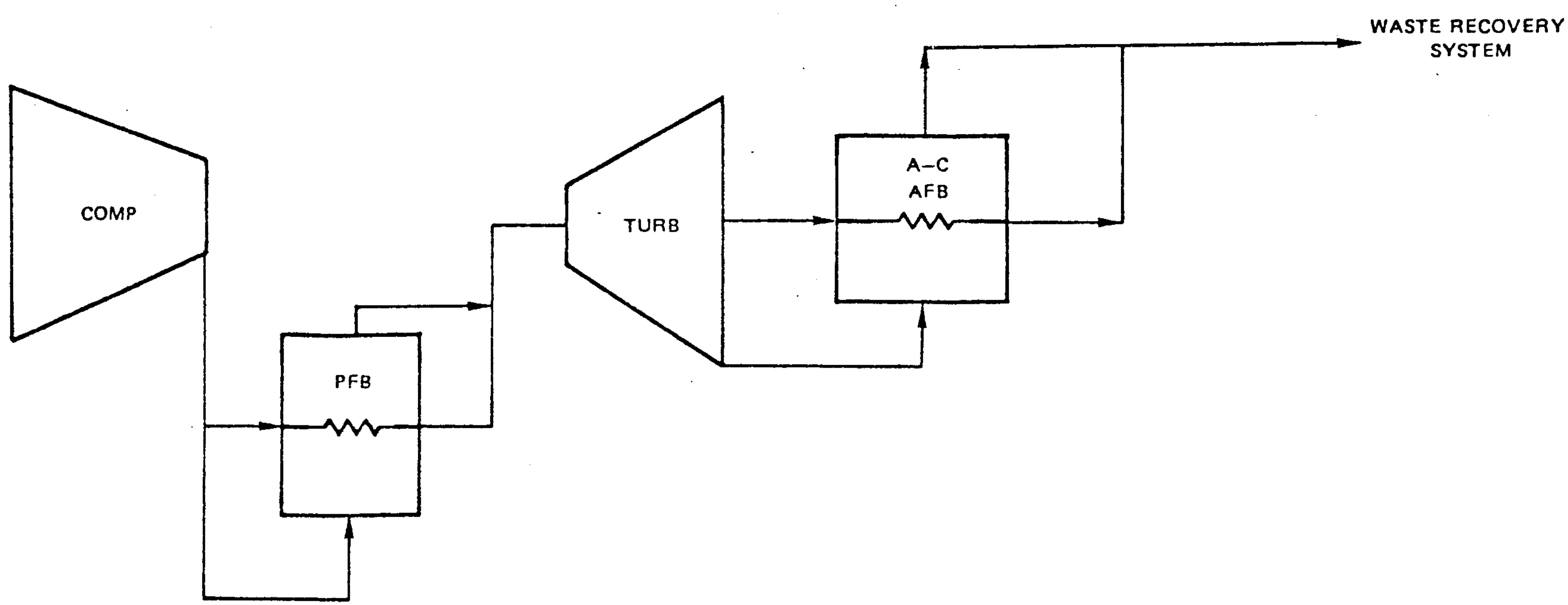


FIGURE: A-4

COGAS WITH EXHAUST-FIRED STEAM COOLED AFB

(CASE C)

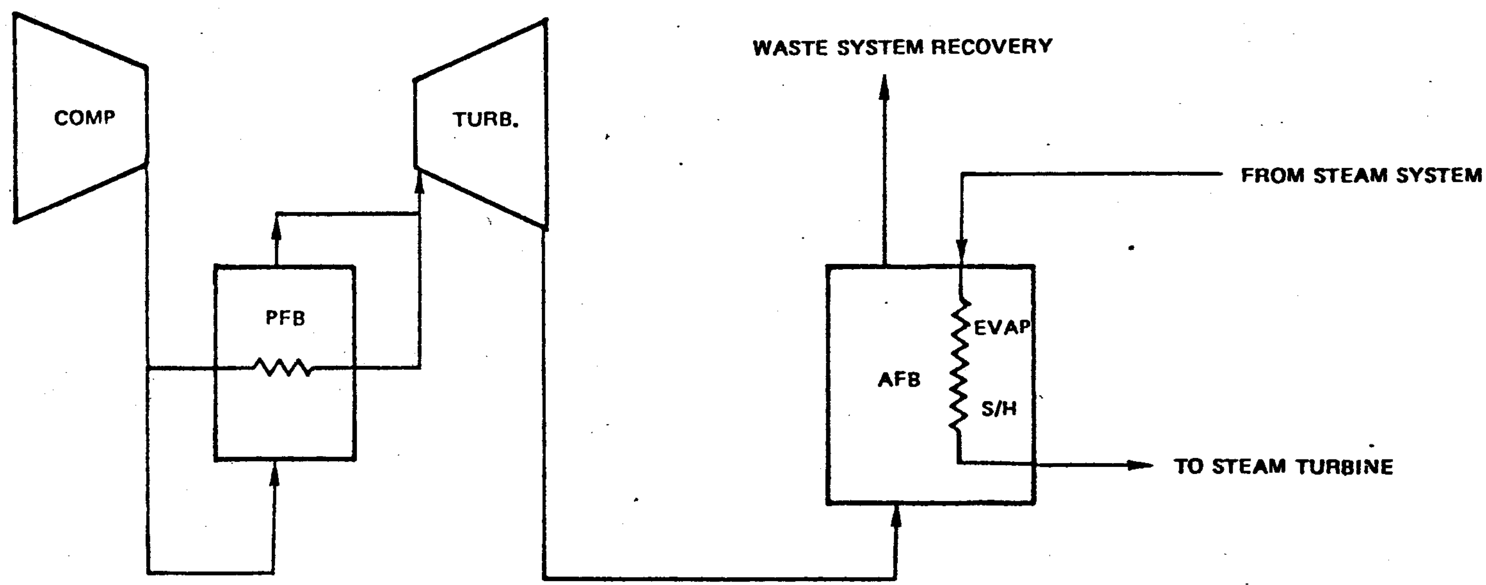


FIGURE: A-5

REHEAT BEFORE POWER TURBINE
(CASE D)

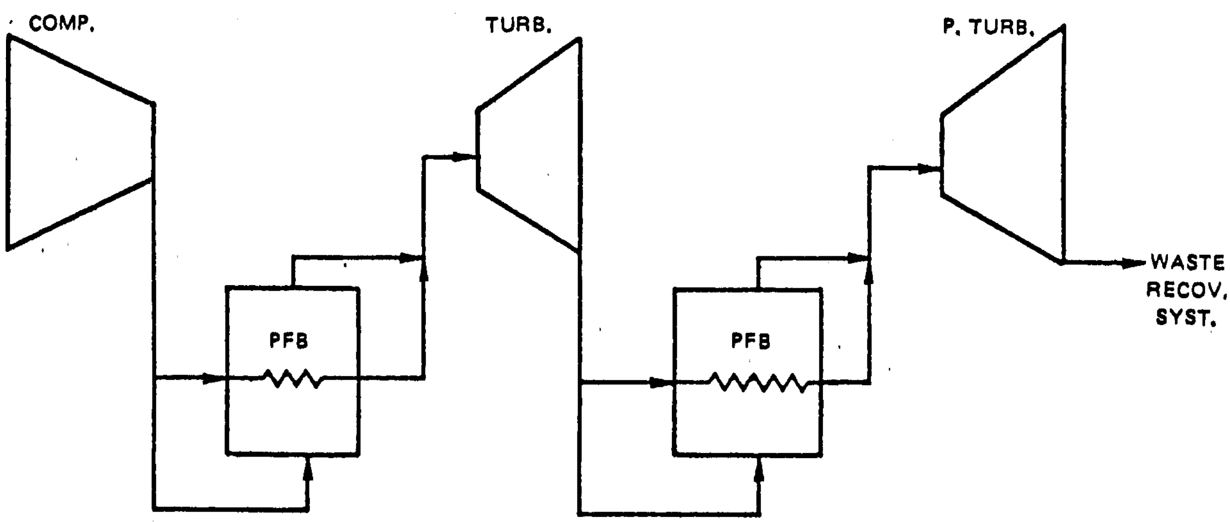


FIGURE: A-6

2.2.4.2.2 System Performance Evaluation

Table A-1 shows the assumptions used in simulating the various systems and Table A-2 presents the performance results along with a description of the steam bottoming cycles. While the optimum overall pressure ratio (OPR) is 10 for several of the cycles studied, cases utilizing an OPR of 16 have also been examined since higher pressures imply smaller combustors, piping, cyclones, etc., and, therefore, potentially lower overall costs. Efficiencies and specific works obtainable for different OPR's and different cycles are shown in Figure A-7.

The COGAS (Case A) configuration results in a relatively low gas turbine exhaust temperature at the inlet of the waste heat boiler (WHB). The steam plant efficiency is a function of the WHB effectiveness and steam cycle efficiency. For a given steam pressure, an increase in gas turbine exhaust temperature results in increased steam production, lower stack temperatures, and increased WHB effectiveness. In addition, the higher gas temperature permits higher steam temperatures. As a result of these two effects, steam plant efficiency increases with the increase in gas turbine exhaust temperature that accompanies a decrease in OPR. The gas turbine cycle efficiency also increases as OPR decreases down to a certain value below which the gas turbine cycle efficiency decreases. Therefore, the combined cycle plant efficiency also reaches a maximum at a certain optimum OPR (generally around 10:1 for a gas turbine inlet temperature of 1600°F). However, since the gas temperature entering the boiler surfaces in the COGAS case is low compared to the other configurations studied, the steam plant efficiency and, thus, the combined cycle efficiency also tends to be relatively low.

Case A only utilizes one stage of combustion per gas turbine; therefore, each gas turbine/WHB/Steam Turbine module is less complex than the corresponding modules of the other alternatives studied. While the overall plant layout might prove more complex because of the greater number of modules required, it was decided to use Case A as the reference cycle for the economic analysis of all the systems. Since each module is less complex, Case A may be of interest as an initial PFB demonstration plant or small commercial plant (i.e., approximately 100 MWe or less).

Cases B and C show that supplementary exhaust firing increases system efficiency to 40-41 percent. The steam-cooled AFB configuration, Case C, shows a higher specific work output than the air-cooled AFB configuration (Case B) since more fuel may be burned in a steam cooled bed without exceeding the desired 1550°F bed temperature, and thus more steam is made. In both cases, the 10 OPR case results in a better efficiency (i.e., is closer to the optimum) than the 16 OPR case since the available heat in the gas turbine exhaust is higher at 10 OPR than at 16 OPR, therefore less coal is required in the AFB to achieve the same steam conditions and flow.

Case D utilizes gas reheat before the power turbine. This method has two advantages; first, the turbine exhaust temperature is

TABLE A-1

System Assumptions for Performance Analysis

Combustion Efficiency, %		
PFB, main and reheat	99.0	
AFB (with carbon burnup bed)	98.5	
Pressure Loss, % of local gas pressure		
	Bed	Cooling Tubes
PFB, main and reheat	10.0	10.0 (air)
AFB	9.2	- (steam)
Temperature, °F		
	Bed	Cooling Tubes
PFB, main	1650	1575 (air)
PFB, reheat	1550	1475 (air)
AFB	1550	- (steam)
Component Efficiency, %		
Electric generator (steam turbine)	98.4	
Electric generator (gas turbine)	98.7	
Electric motors	95.0	
Boiler feed pump	82.0	
Boiler feed pump drive turbine	75.0	
Condensate pump	82.0	
ID fan	70.0	

TABLE A-2

Alternative Powerplant Configuration Performance Study

Parameters	Case A Basic Combined Cycle		Case B Exhaust-Fired Air-Cooled AFB		Case C Exhaust-Fired Steam-Cooled AFB		Case D Gas Reheat at Power Turbine Inlet
Overall Press. Ratio	10	16	10	16	10	16	16
No. Gas Turbines Used	6	8	4	4	3	4	4
Comp. Inlet Air Flow (lb/sec)	4896	6528	3264	3264	2448	3264	3264
Total Gross Plant Output (MW)	592	648	584	568	592	706	628
Gross Efficiency (HHV),%	38.4	37.2	41.0	40.3	40.9	40.4	43.1
<u>Steam Cycle</u>							
Throttle Pressure, psig	250	250	2400	2400	2400	2400	2400
Superheat Temp., ^{o}F	734	605	1000	1000	1000	1000	950
Reheat Temp., ^{o}F	-	-	1000	1000	1000	1000	950

ALTERNATIVE PFB POWERPLANT PERFORMANCE COMPARISON

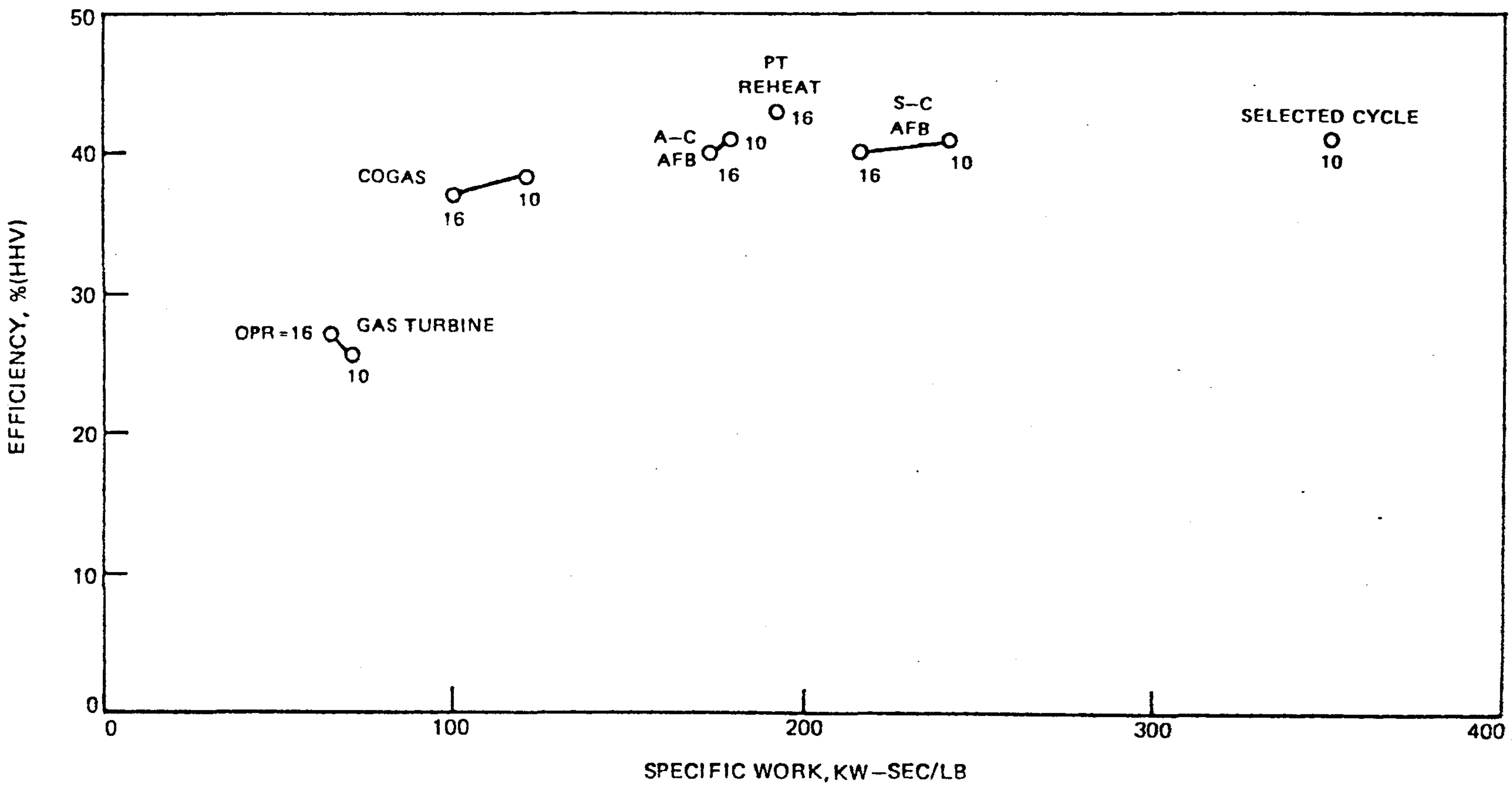


FIGURE: A-higher and therefore, a better steam system (than Case A) is possible, and second, the power turbine is able to produce more power per pound of airflow. These factors lead to a highly efficient power plant (43 percent); but, the complexity of the plant is increased because, to reheat the gas, two additional PFB coal combustors are required along with their associated particulate removal and pressurized solids injection and removal systems. In addition, the gas turbine requires considerable redesign in order to adapt it to the Case D configuration.

An alternate reheat location has been briefly studied. This configuration places the gas turbine reheat before the low-pressure turbine (LPT), but the overall efficiency is not as high as for the power turbine reheat case. Also, placing the reheat combustor before the LPT causes many mechanical design problems for the gas turbine. For these reasons work on this configuration has not been carried further.

2.2.5 Economic Screening Analysis

The selection of the commercial plant configuration cannot be made on the basis of performance alone. The most important selection criterion is overall cost of electricity. Therefore, an order of magnitude estimate has been made of the relative capital and operating costs of the alternative configurations. The operating cost differences due to variations in cycle efficiencies between alternatives are expressed in terms of the present worth of fuel savings per kW capacity. The economic parameters assumed are shown in Table A-3.

The results of the economic screening analysis are given in Table A-4. All costs are given as incremental costs relative to Case A (the basic combined cycle unfired waste heat recovery system). Case C (exhaust-fired, steam-cooled AFB) and Case D (power turbine reheat) have the lowest evaluated net relative costs. The cost differential between these two systems is not statistically significant. The power turbine reheat cycle requires a more complex gas turbine design with additional hot particulate removal and pressurized lock hopper solids injection equipment. In addition, little data are available for design of a PFB combustor at the 2.5 atmosphere pressure existing at the reheat point. Therefore, the PFB cycle with an exhaust-fired, steam-cooled AFB would offer less technical risk.

Capital costs have not been estimated for all major pieces of equipment or systems required in a plant. Table A-5 lists those items which have been considered. Some major systems (such as the coal and sorbent feed systems to the AFB and low-pressure reheat PFB combustors) have been omitted which would tend to decrease the advantage of the reheat and exhaust-fired cycles. However, the differences in the costs of these systems should not be large enough to offset the differneces shown in Table A-4. Therefore, there is a strong probability that the trends shown in this study will be confirmed by a more detailed design and cost estimate of the alternates. On the basis of these results, the Department of Energy has authorized additional studies on Cases A and D under an

TABLE A-3

ASSUMED ECONOMIC PARAMETERS

Plant Life	30 years
Capacity Factor	65%
Output Factor	100%
Dollar Base Used	Mid-1977
Escalation	0
Discount Rate	8%
Interest During Construction	8%
Period of Construction	5 years
S-Curve Expenditure Schedule	
Fixed Capital Charge	18%
Replacement Energy Cost	25 mills/kWh
Cost of Coal (3.01% or more sulfur)	0.87/M Btu, $20/ton
Cost of Limestone and Dolomite	$7/ton
Cost of Disposal of Ash and Spent Sorbent	$3/ton
Effective Cost of Fuel to Account for Sorbents and Disposal of Ash and Spent Sorbent at Full Load Condition.	$0.979/M Btu $22.46/ton of coal input

TABLE A-4

PFB Combined Cycle Power Plant Cost Summary

	Screening Analysis						
	Case A		Case B		Case C		Case D
Cycle Type	Basic COGAS		Exhaust-Fired Air-Cooled AFB		Exhaust-Fired Steam-Cooled AFB		Power Turbine Reheat
Gas Turbine Pressure Ratio	10	16	10	16	10	16	16
No. Feedwater Heaters	0	0	0	0	0	0	0
Output per Gas Turbine, MW							
Gas Turbine	66.7	62.0	60.0	55.0	63.4	58.4	77.0
Steam Turbine	32.0	19.0	86.0	87.0	134.0	118.0	88.0
Total	98.7	81.0	146.0	142.0	197.5	176.4	165.0
Specific Work, kW-sec/lb-air	121	100	179	174	241	216	202
Gross Efficiency, % (HHV)	38.4	37.2	41.0	40.3	40.9	40.4	43.1
Relative Equipment Cost, $/kW							
Combustion System	Base	+33	-8	-26	-16	-12	-42
Prime Movers and Electrical	Base	+55	-38	-17	-49	-32	-24
Miscellaneous	Base	-5	+76	+79	-10	-2	+20
Subtotal	Base	+83	+30	+36	-75	-46	-46
Present Worth of Fuel Savings, $/kW	Base	+18	-35	-26	-34	-28	-61
Net Relative Cost, $/kW	Base	+101	-5	+10	-109	-74	-107

TABLE A-5

Major Equipment Included in Cost Summary

Main PFB Coal/Sorbent Feed System

Gas Turbines/Generators

PFB Main Combustors

PFB Reheat Combustors

AFB Combustors (Excluding: Flues, Duct, Cyclones, Fans, Coal/Limestone Feed System)

Electrical Equipment

Steam Turbine/Generator

Waste Heat Boilers

Electrostatic Precipitators

extension to the present contract. These studies are being performed under Subtasks 1.8 and 1.9.

It should be recognized that the cost estimates do not consider some of the material, equipment, and other costs normally associated with the items indicated on Table A-5. In addition, little more than conceptual outline drawings were available for many items that were considered. This preliminary analysis is not intended to provide a definitive evaluation of the economic merits of each alternative cycle. Rather, the intent of the effort was to provide a systematic approach for summarizing the relative pros and cons of each cycle on the basis of preliminary design calculations. While each pro and con was, in effect, weighted on a cost basis, it would be misleading to consider the numbers shown as anything more than a rough order of magnitude.

2.2.6 Selected Commercial Plant Configuration

2.2.6.1 Plant Description

On the basis of the screening analysis, the PFB combined cycle power plant with the steam-cooled AFB has been selected for the commercial plant conceptual design study. Further optimization of the selected configuration has led to incorporation of three stages of regenerative feedwater heating and an adjustment in the relative power split between the gas and steam turbine.

The amount of supplementary firing of the gas turbine exhaust flow has been chosen to provide the highest system efficiency as shown in Fig. A-8. Firing in the gas turbine system alone (100 percent Q_{GT}/Q_{Total}) results in a low-temperature steam system, providing approximately 38 percent system efficiency. As additional fuel is fired in the AFB, a higher efficiency steam system may be used resulting in higher overall system efficiencies. At the other extreme of no gas turbine participation, the efficiency ($\sim$36 percent) falls off since the combustion heat energy is recovered only at steam cycle efficiency and not at combined cycle efficiency.

In the resulting system, shown in Figure A-9, two gas turbines are used operating at 10:1 pressure ratio with 1600^{o}F inlet temperature. Each gas turbine requires two PFB combustors. The gas turbines exhaust into a single exhaust-fired AFB steam generator which generates steam at 2400 psig/1000^{o}F/1000^{o}F to drive a single steam turbine.

2.2.6.2 Sensitivity Studies

Key design parameters have been varied to determine their effect on the performance of the selected power plant. Table A-6 lists the parameters that have been varied. In addition to those shown, the effects of various steam system pressure drops have also been calculated for other cycle configurations. These calculations have indicated that the addition of a 10% pressure loss ($\Delta P/P$) to the steam system would decrease overall plant efficiency by 0-0.2 percentage points, depending on location (with the reheater being the most sensitive area).

OVERALL SYSTEM EFFICIENCY VARIATION WITH GAS TURBINE EXHAUST SUPPLEMENTARY FIRING

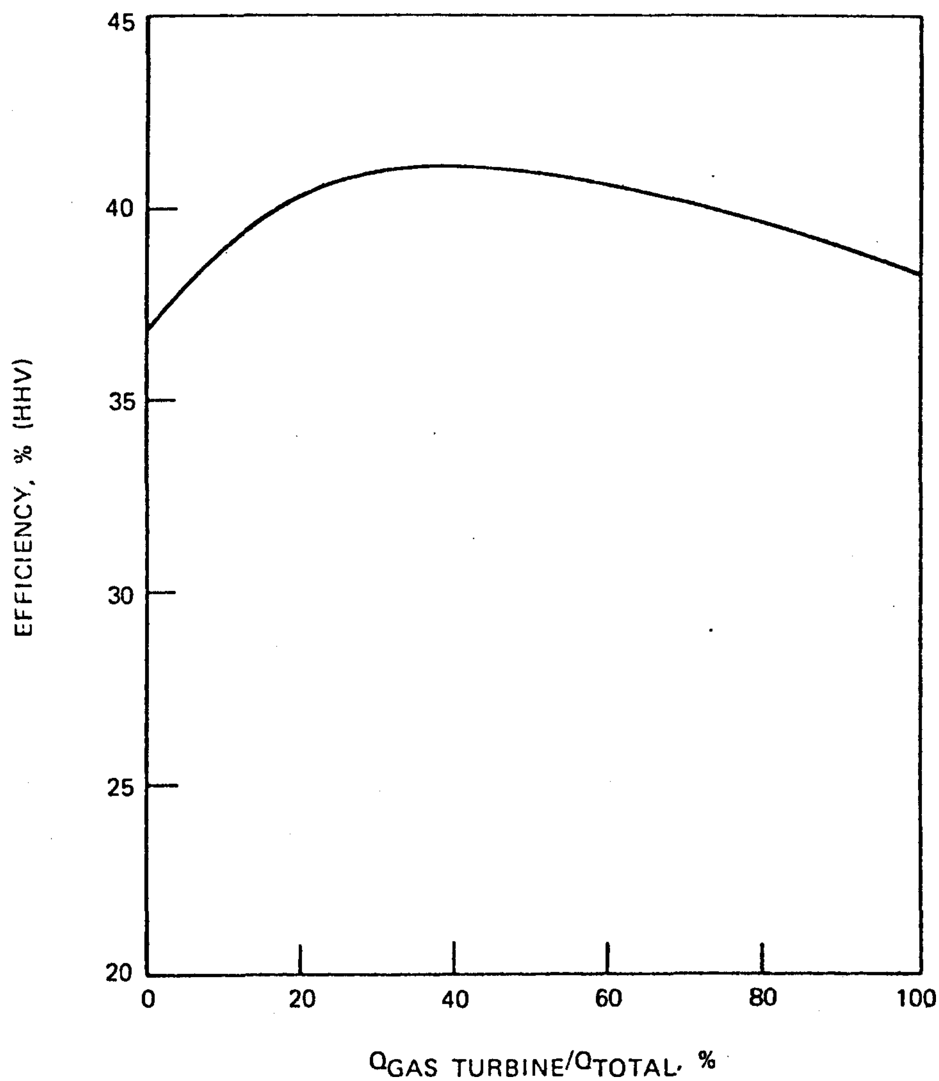


FIGURE: A-8

SCHEMATIC OF SELECTED PFB/AFB COMBINED CYCLE POWER PLANT CONFIGURATION

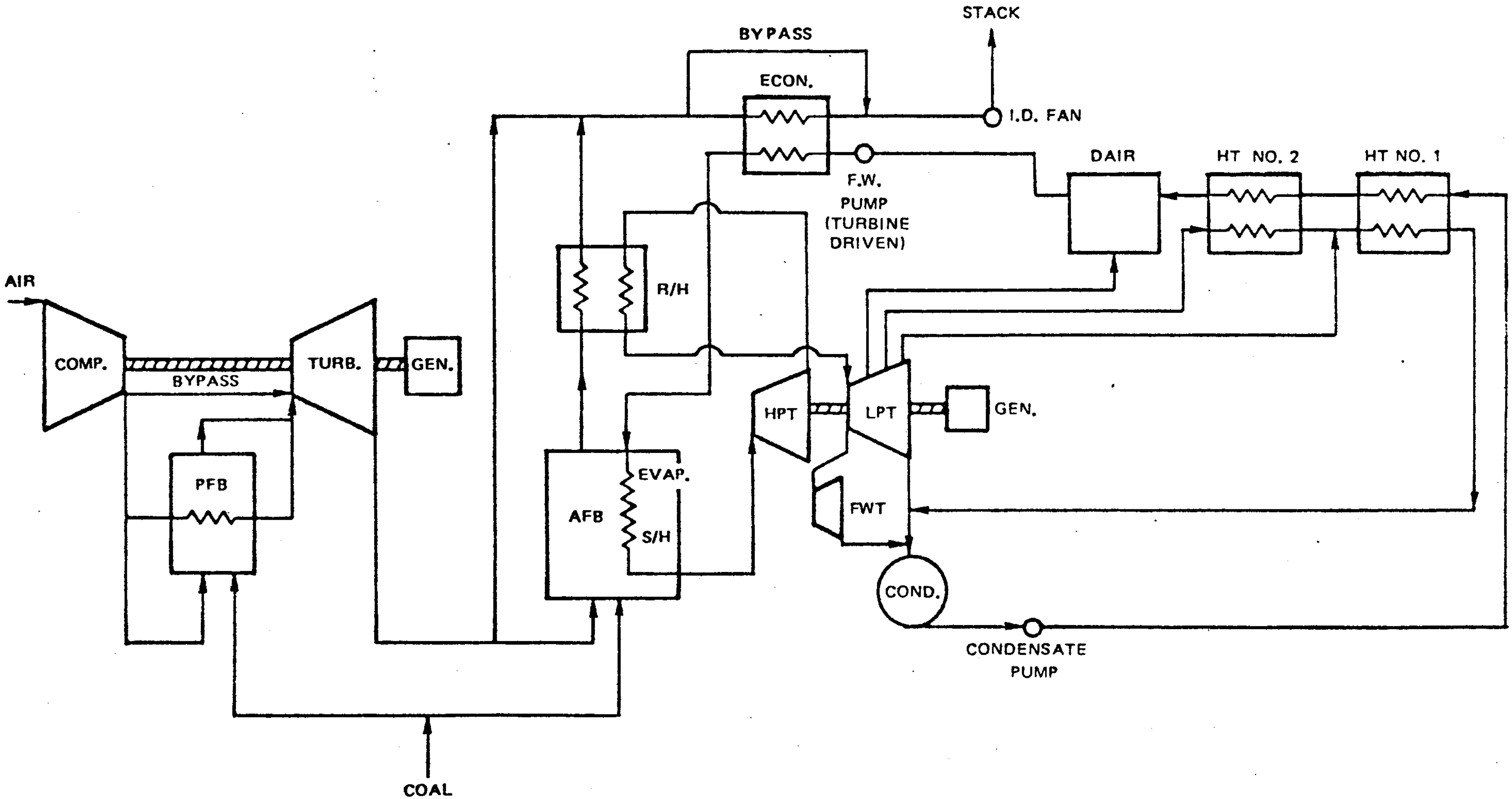


FIGURE: A-9

TABLE A-6

Parameters for Sensitivity Study

Parameters Changed	Base Value	Changed Value	
Turbine Inlet Temperature, $^{\circ}F$	1600	1500	1700
Turbine Cooling Airflow, % WAE	3.5	0	7.0
PFB Combustor $\Delta P/P$, %	10	0	20
AFB Combustor $\Delta P/P$, %	9.2	0	20
Condenser Pressure, " Hg	2.0	2.0	3.0

Table A-7 shows the effects on performance of changing the variables listed in Table A-6. Condenser pressure variation and gas turbine inlet temperature have the largest effect on overall system performance. As the steam condenser pressure increases, the amount of power generated by the steam cycle decreases due to the reduced expansion ratio (proportional to output power) of the steam turbine. Since the steam system provides approximately 80 percent of the overall power, this variable shows a stronger effect on the overall system than a proportional change in pressure loss of the gas turbine system.

Turbine inlet temperature variation has a strong effect on system power. A decrease of 100^{o}F in temperature causes approximately a 9.4 percent reduction in power. Not only does the gas turbine power output decrease, but the heat input to the AFB system from the gas turbine exhaust decreases. It would be possible to restore the system power output by increasing the firing rate in the AFB.

Figures A-10 through A-19 show the gas turbine system, steam system, and total system performance as functions of the variable cycle parameters. As turbine inlet temperature is increased, system power and efficiency also increase as shown in Figure A-10 (power output) and Figure A-11 (efficiencies). Unfortunately, the bed temperature is limited to about 1650^{o}F.

Increasing the amount of turbine cooling airflow (TCA) causes a decrease in system power and efficiency as shown in Figures A-12 and A-13. The energy level throughout the turbine decreases since the additional TCA dilutes the mainstream gas flow and, also, less gas flow is available to do work in the initial stages of the turbine. In addition, the gas turbine exhaust contains less heat which affects the steam system fuel requirements. Therefore, it is advantageous to limit the amount of cooling airflow and to minimize any leakages in the turbines to maintain high power output and efficiency.

Increased pressure losses in the gas turbine system cause a power and efficiency penalty. Figures A-14 through A-17 show the effects of varying the PFB and AFB combustor pressure losses on power output and system efficiency. The pressure level entering the turbine decreases as the PFB combustor pressure loss increases. This effect reduces the amount of expansion (power output) in the turbine since the turbine exit pressure is fixed at approximately ambient conditions. However, the gas turbine exit flow contains more energy thus benefiting the steam cycle. Steam system power increases with increasing pressure loss because of the higher gas turbine exhaust temperatures. This increase in steam power is not enough, though, to offset the loss in gas turbine power so that the net effect is a decrease in overall system power and efficiency.

Pressure loss in the AFB combustor also affects the expansion ratio of the turbine. To maintain the AFB furnace pressure at approximately ambient conditions, the turbine expansion must be adjusted to make up for any pressure losses between the turbine and the top of the beds. Figures A-16 and A-17 show the power and efficiency variations with AFB combustor pressure losses.

TABLE A-7

PFB/AFB Powerplant Performance Sensitivity Analysis

Variable Changed	GT Pwr,[1] MW	GT η, %	Steam Pwr, MW	Total Net Pwr., MW	% Δ Pwr System	% η System	$\Delta\eta$ Points
No Change (Base Case)	127.0	25.3	461	588	-	41.1	-
$-100^{\circ}F$ Turbine Inlet Temp.	108.0	24.2	429	533	-9.4	40.6	-.5
Turbine Cooling Air Flow = 7% (Twice Base)	116.2	24.0	452	564	-4.1	40.9	-.2
PFB Combustor $\frac{\Delta P}{P}$ = 20% (Twice Base)	111.6	22.2	465	573	-2.6	40.6	-.5
AFB Combustor $\frac{\Delta P}{P}$ = 20% (Twice Base)	110.4	22.0	465	572	-2.7	40.6	-.5
Condenser Pressure = 3" Hg (Base - 2½" Hg)	127.0	25.3	450	573	-2.6	40.5	-.6

(1) Based on 2 gas turbines per plant.

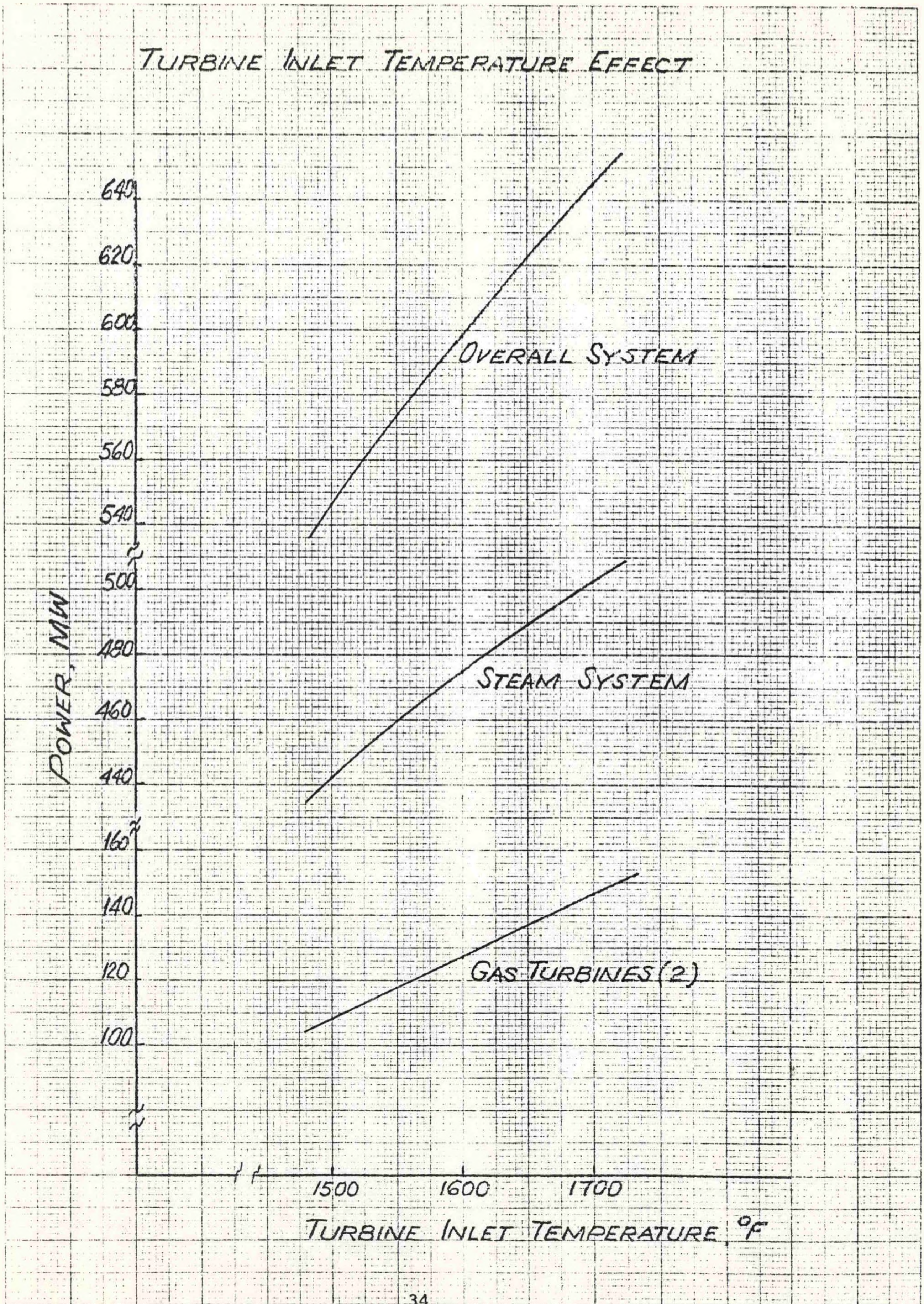

TURBINE INLET TEMPERATURE EFFECT
OVERALL SYSTEM
STEAM SYSTEM
GAS TURBINES (2)
POWER, MW
640
620
600
580
560
540
500
480
460
440
160
140
120
100
1500
1600
1700
TURBINE INLET TEMPERATURE, °F

FIGURE: A-10

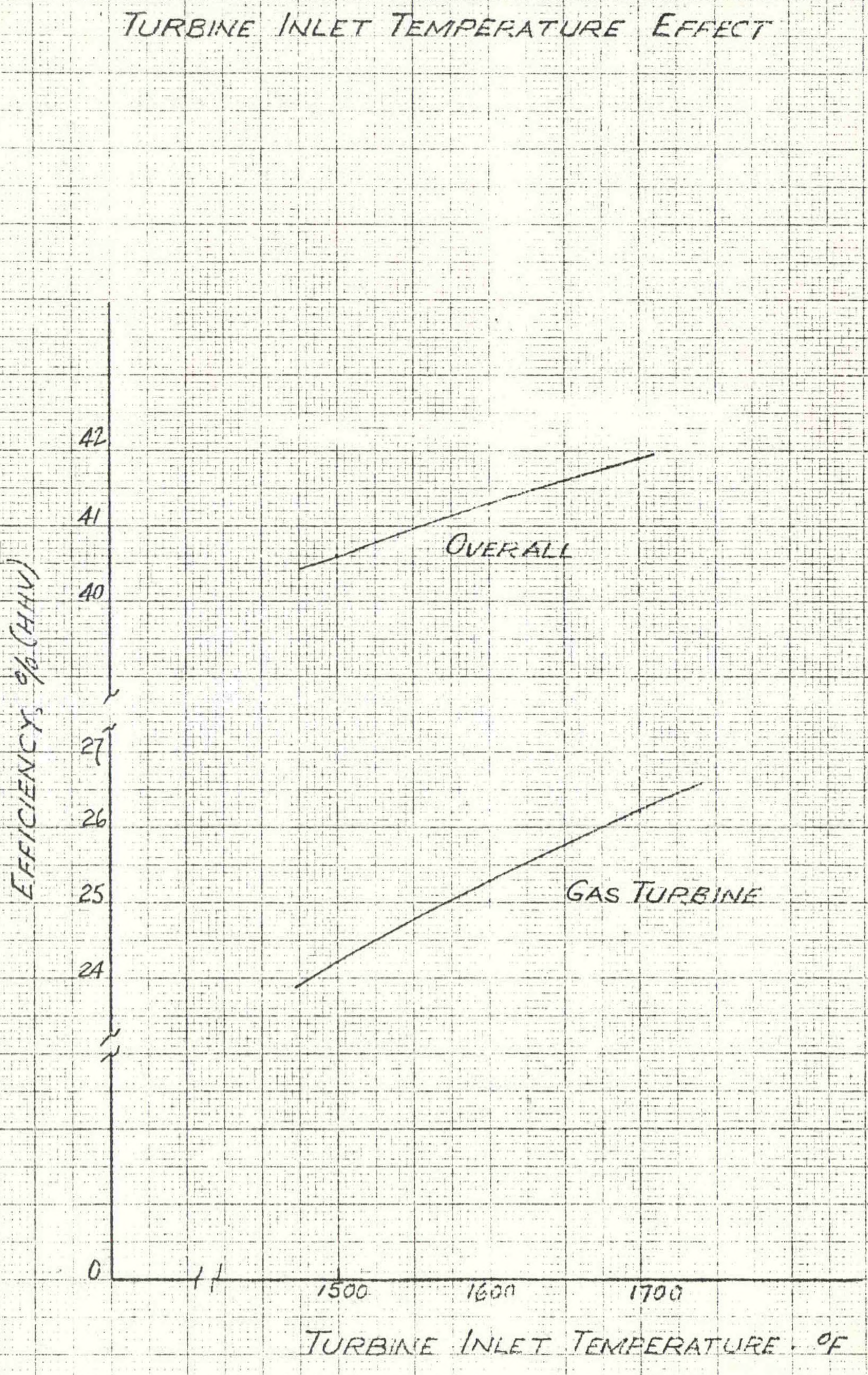
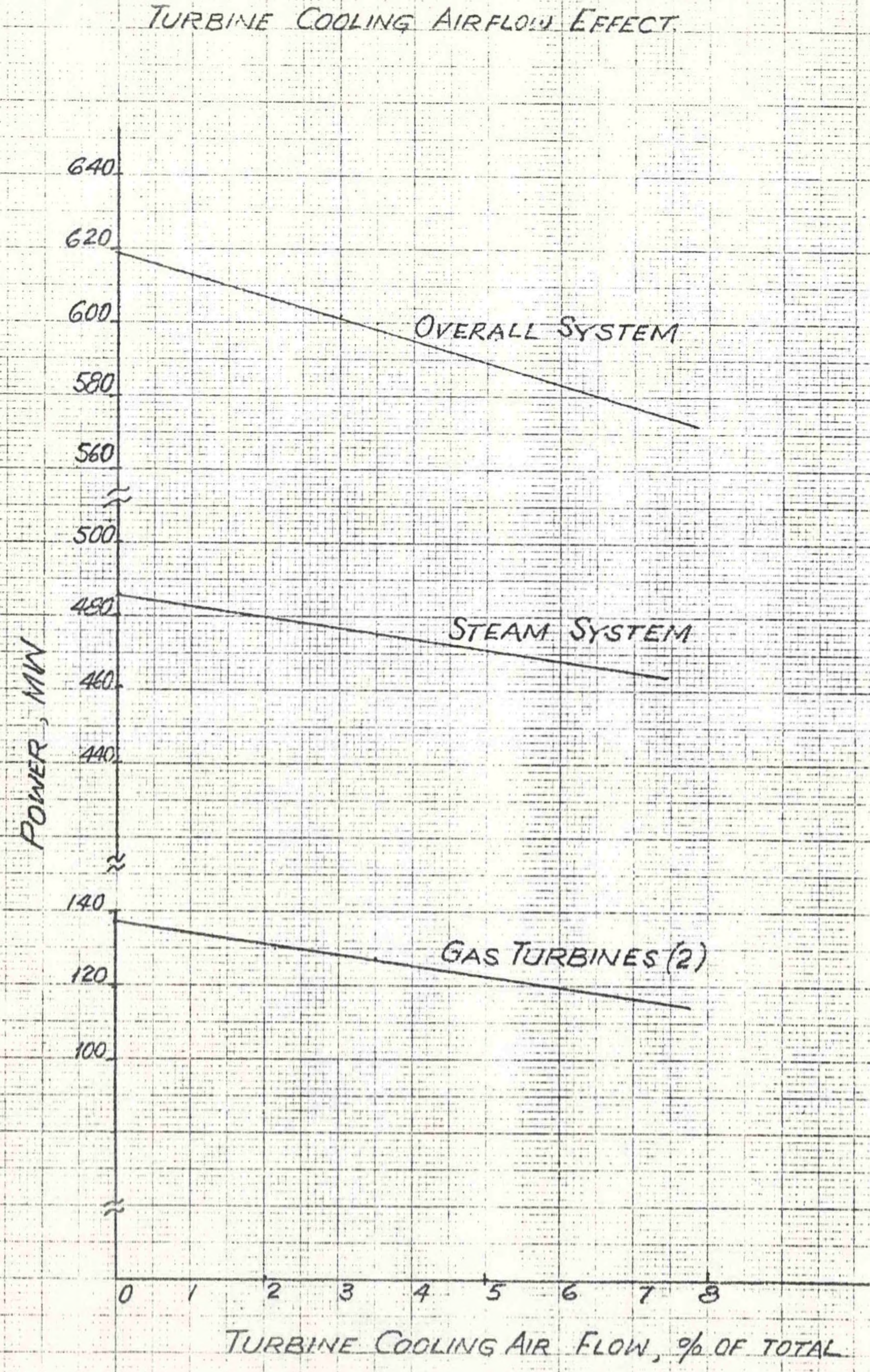

TURBINE INLET TEMPERATURE EFFECT
42
41
40
27
26
25
24
0
OVERALL
GAS TURBINE
EFFICIENCY, % (HHV)
1500
1600
1700
TURBINE INLET TEMPERATURE, °F

FIGURE: A-11



TURBINE COOLING AIRFLOW EFFECT
640
620
600
580
560
500
480
460
440
140
120
100
POWER, MW
OVERALL SYSTEM
STEAM SYSTEM
GAS TURBINES (2)
0
1
2
3
4
5
6
7
8
TURBINE COOLING AIR FLOW, % OF TOTAL

FIGURE : A-12

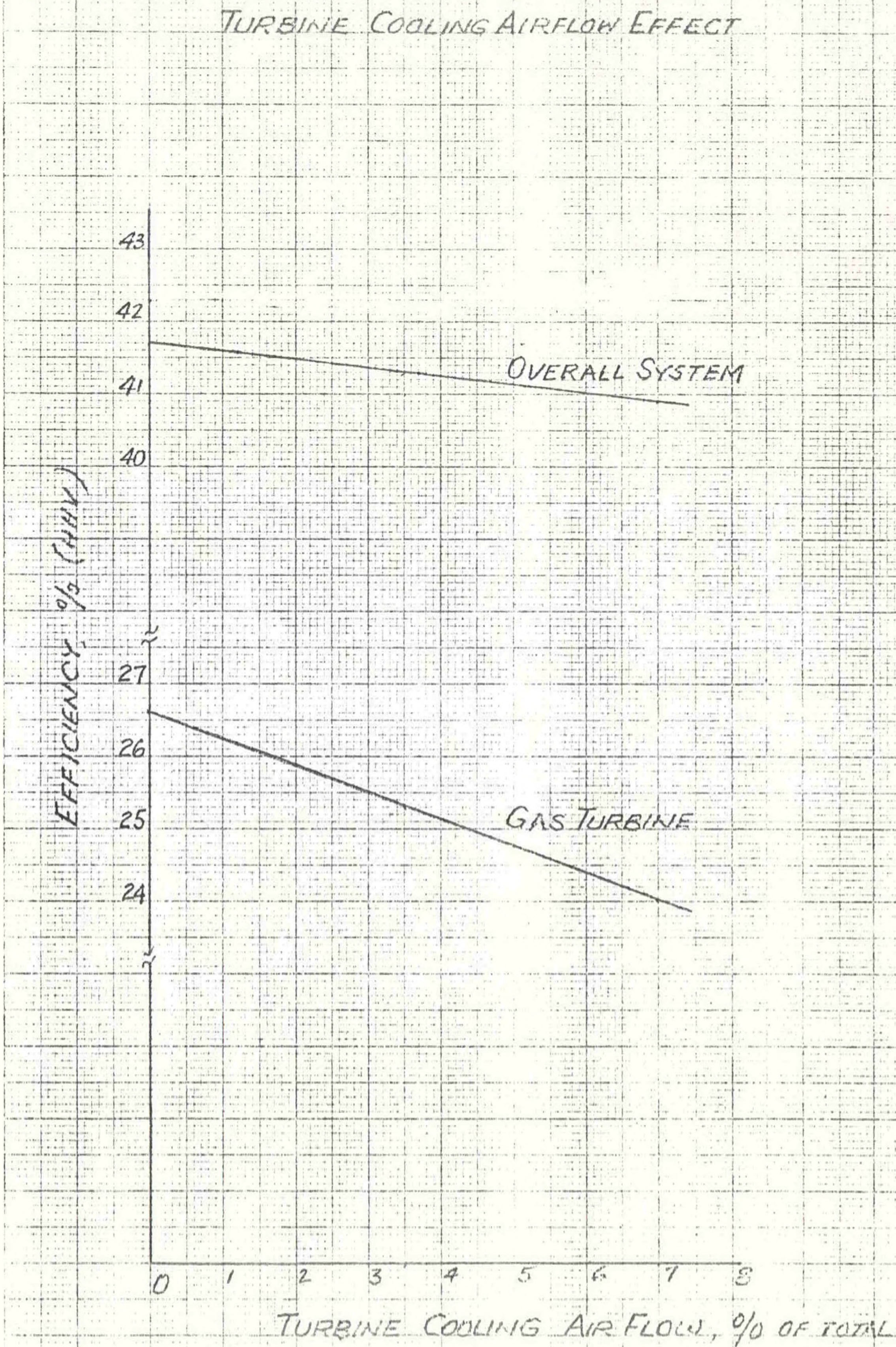
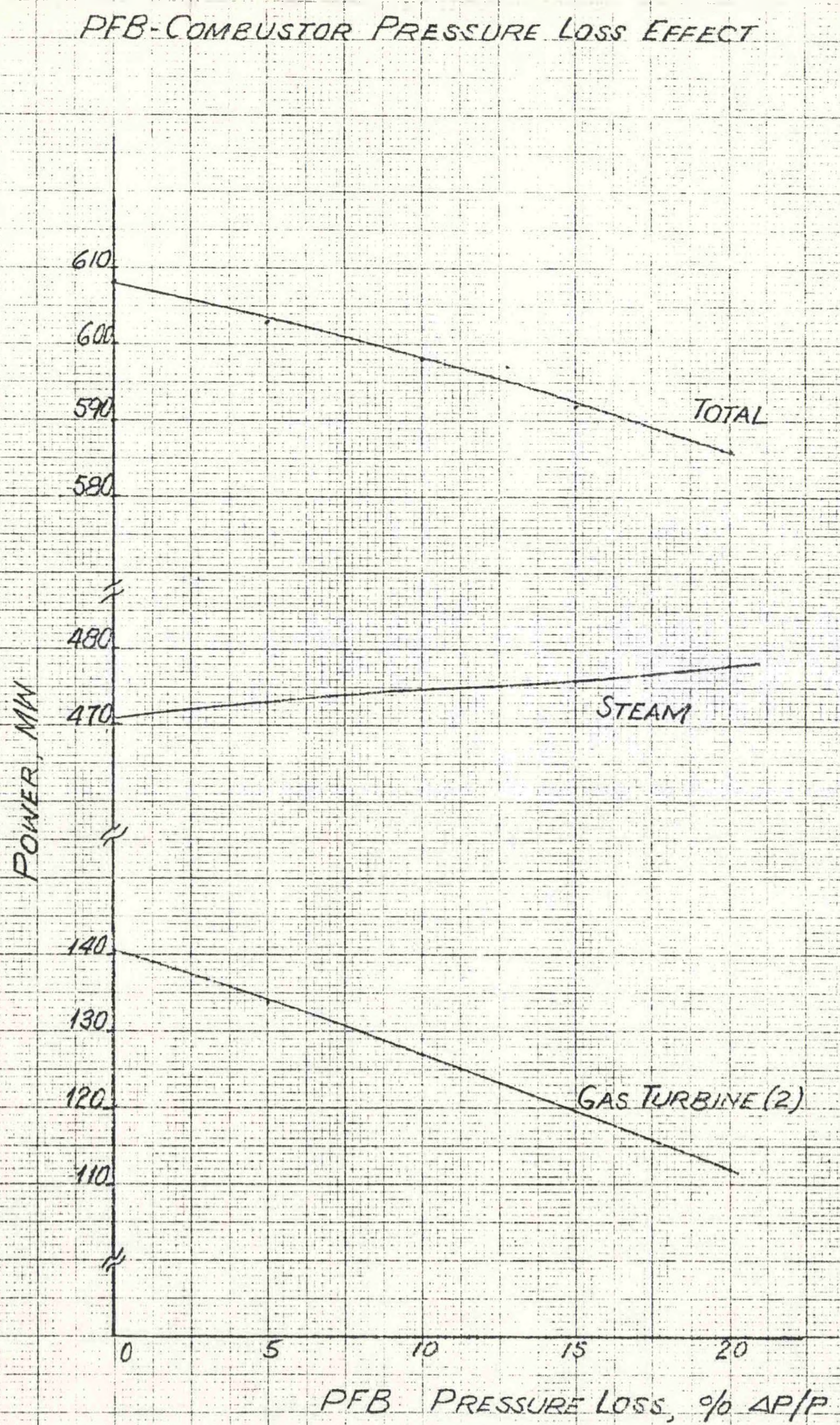

TURBINE COOLING AIRFLOW EFFECT
43
42
41
40
27
26
25
24
EFFICIENCY, % (HHV)
OVERALL SYSTEM
GAS TURBINE
0
1
2
3
4
5
6
7
8
TURBINE COOLING AIR FLOW, % OF TOTAL

FIGURE: A-13



PFB-COMBUSTOR PRESSURE LOSS EFFECT
610
600
590
580
480
470
140
130
120
110
POWER, MW
TOTAL
STEAM
GAS TURBINE (2)
0
5
10
15
20
PFB PRESSURE LOSS, % ΔP/P

FIGURE: A-14

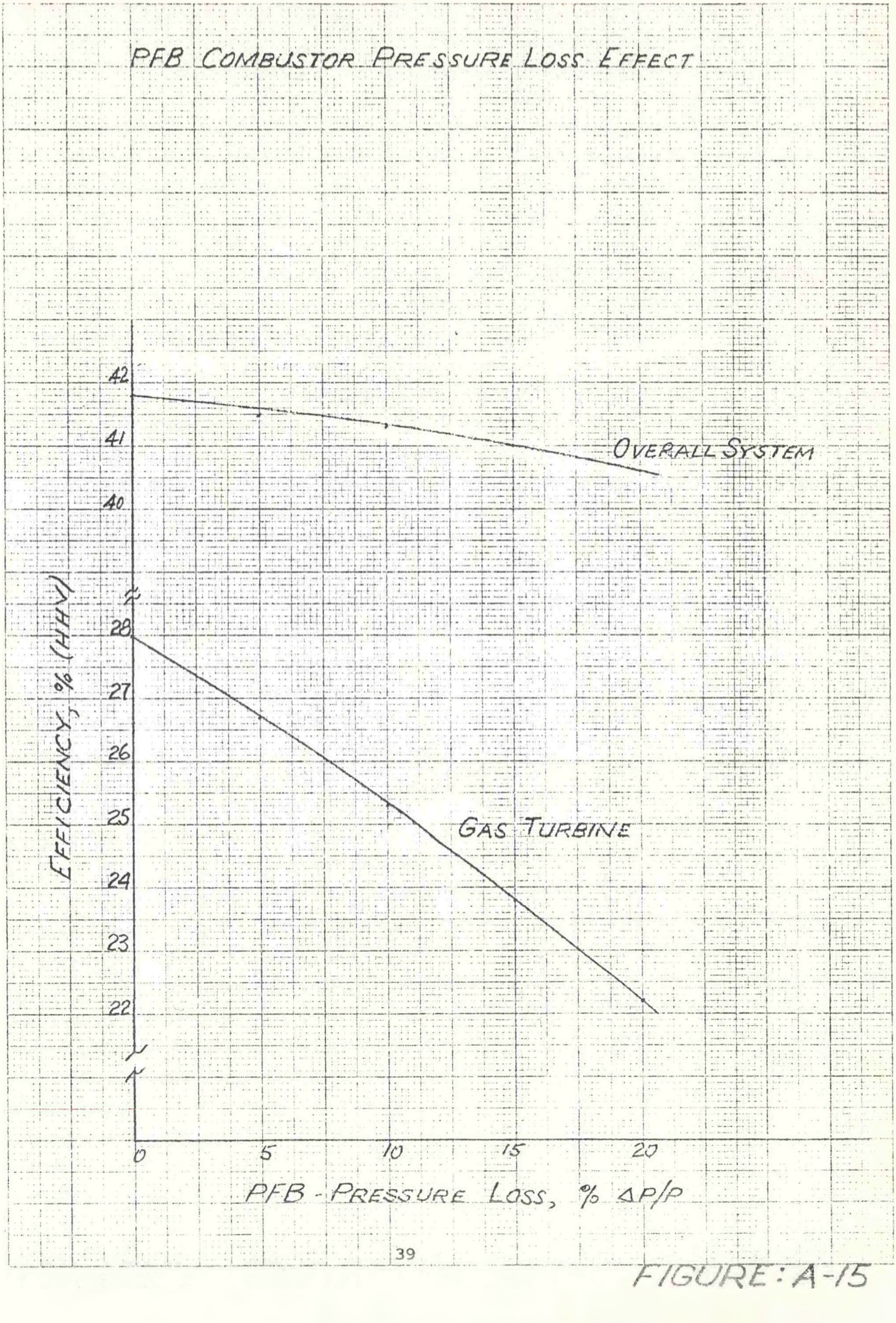

PFB COMBUSTOR PRESSURE LOSS EFFECT
EFFICIENCY, % (HHV)
42
41
40
28
27
26
25
24
23
22
OVERALL SYSTEM
GAS TURBINE
0
5
10
15
20
PFB - PRESSURE LOSS, % ΔP/P

FIGURE: A-15

AFB PRESSURE LOSS EFFECT

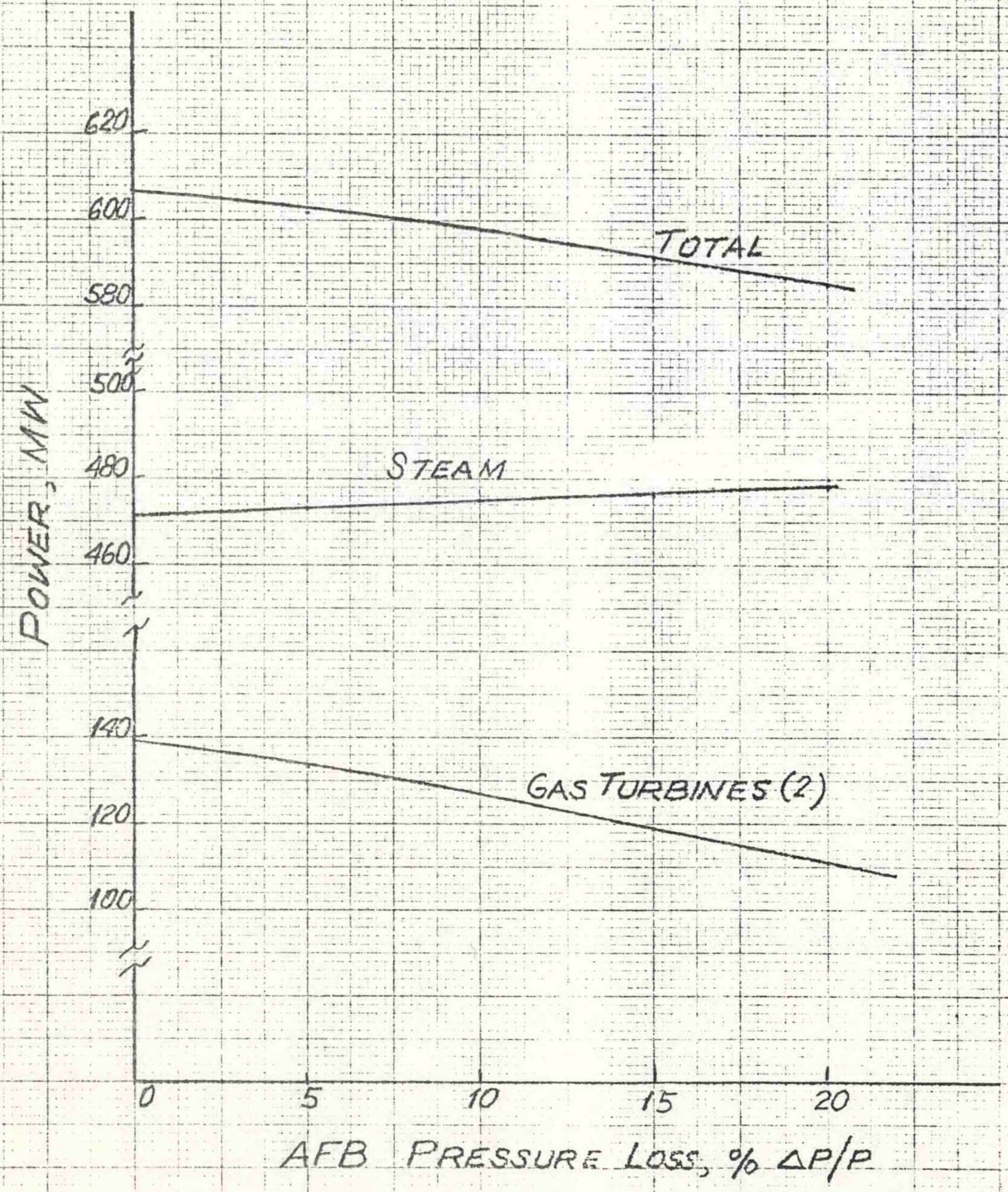
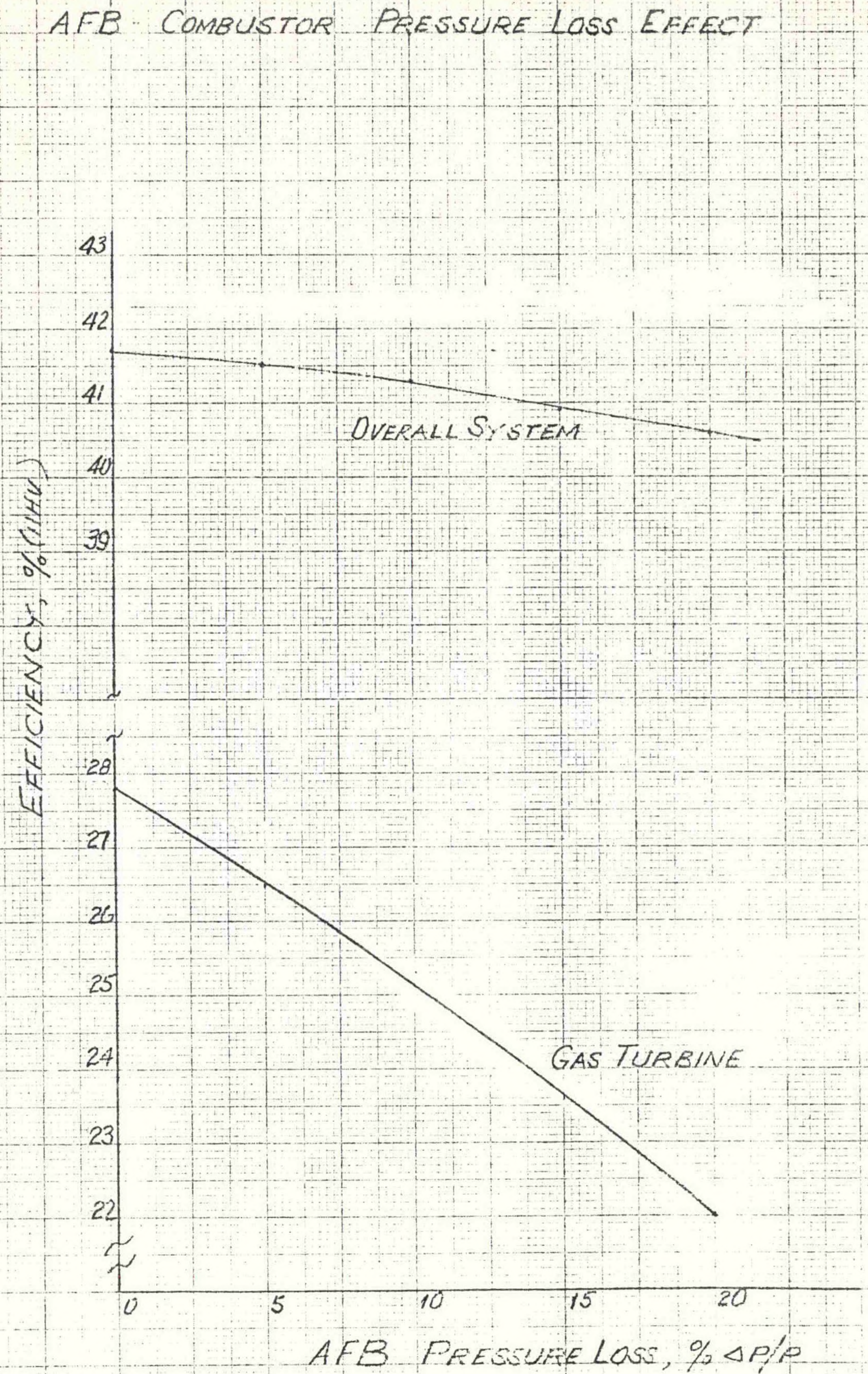


FIGURE: A-16



AFB COMBUSTOR PRESSURE LOSS EFFECT
43
42
41
40
39
28
27
26
25
24
23
22
EFFICIENCY, % (HHV)
OVERALL SYSTEM
GAS TURBINE
0
5
10
15
20
AFB PRESSURE LOSS, % ΔP/P

FIGURE: A-17

Figures A-18 and A-19 present the effects of steam condenser pressure on the power plant output and efficiency (the gas turbine system is not affected at all by this variable). An increase in back pressure for the steam turbine decreases the overall expansion ratio and, thus, lowers steam power output. Since the amount of energy input to generate steam is the same, the steam cycle efficiency is lower with reduced power output.

Variations of the gas turbine overall pressure ratio (OPR) have also been examined, and the results are shown on Figures A-20 and A-21. As OPR increases, gas turbine power output decreases as indicated on Figure A-20 due to the additional amount of work that the turbine is required to perform to drive the compressors. Since the temperature (energy level) into the turbine is constant, less energy is available to drive the generator to produce power. The exhaust temperature decreases with increasing OPR so that less steam can be generated and, thus, the steam system produces less power. The design OPR is selected so as to produce optimum cycle efficiency. Figure A-21 shows the variation in gas turbine and overall system efficiency with OPR. The gas turbine efficiency peaks at pressure ratio between 14:1 and 18:1, but when integrated with the remaining power system, the optimum efficiency is shown to be at an OPR of 10:1.

2.2.7 Conclusions

The selected PFB power plant configuration (Case C) is an excellent compromise between complexity, efficiency, and cost. Other configurations that were briefly studied have shown either better overall efficiency or less complexity than the selected case and may be attractive alternatives.

The configuration using reheat before the power turbine (Case D) has the highest overall efficiency and appears equivalent in overall cost to the selected case. It was passed over as the preferred configuration primarily on the basis of complexity and higher technical risk. However, since the costing analysis was performed on an order of magnitude basis, a more detailed analysis of the reheat cycle is warranted.

The simple COGAS PFB power plant (Case A) is a less complex system than the other configurations studied. This power plant may be attractive as an interim system to bridge the gap between present power plants and more efficient PFB/AFB cycles. An overall efficiency of 37-38 percent is more efficient than today's steam plants (with scrubbers), and the utilities may find this power plant appealing. Also, certain equipment not included in the total cost estimates of the other alternative systems would not apply to the simple COGAS system, thus decreasing the difference in costs by increasing the relative costs of Cases B through D. Furthermore, Case A may be more competive in the smaller plant sizes (say less than 100MWe). Since the generating capacity of each gas turbine and associated waste heat steam system is only 90-100 MWe, based on the largest gas turbines available

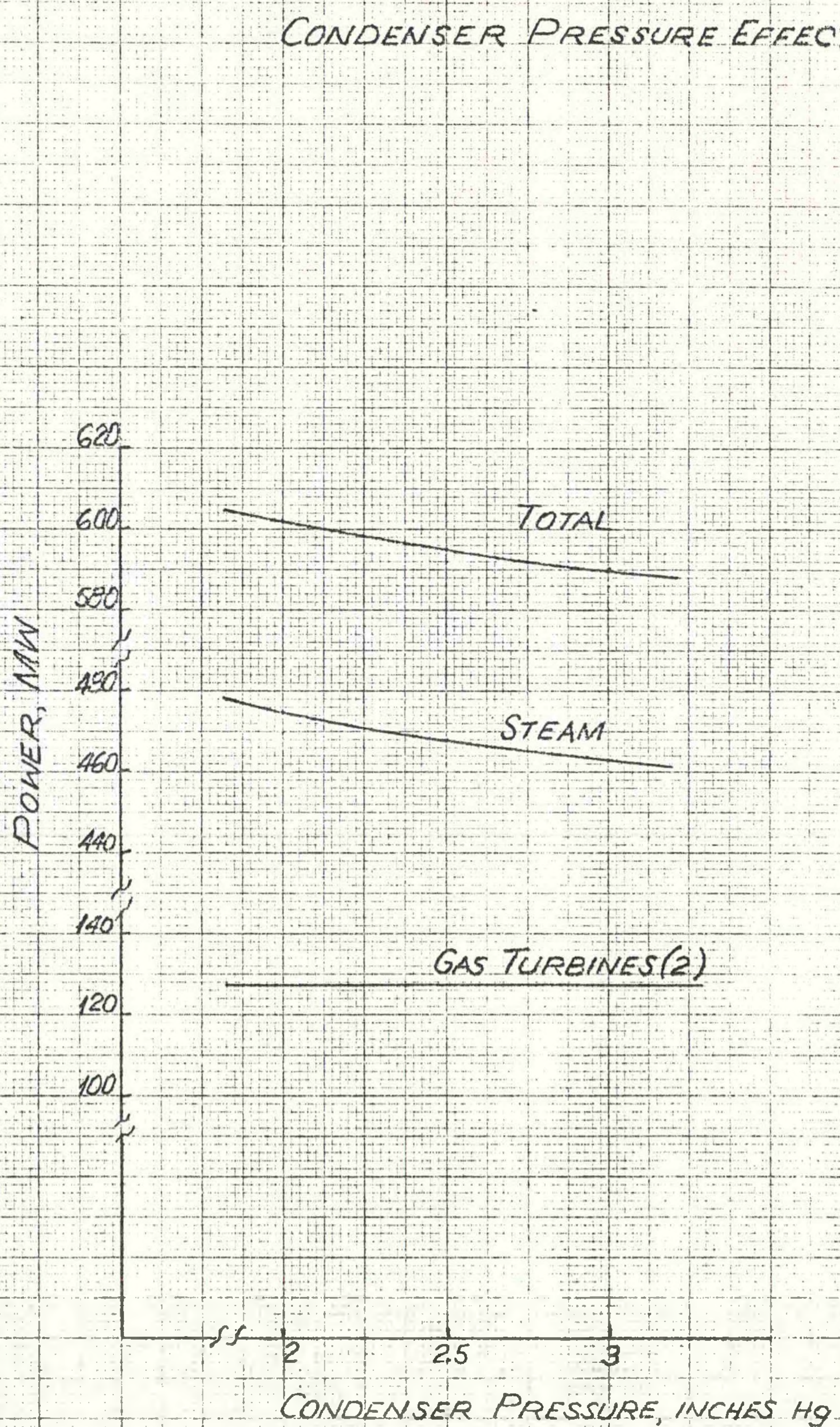

CONDENSER PRESSURE EFFECT
POWER, MW
620
600
580
490
460
440
140
120
100
TOTAL
STEAM
GAS TURBINES (2)
2
2.5
3
CONDENSER PRESSURE, INCHES Hg

FIGURE: A-18

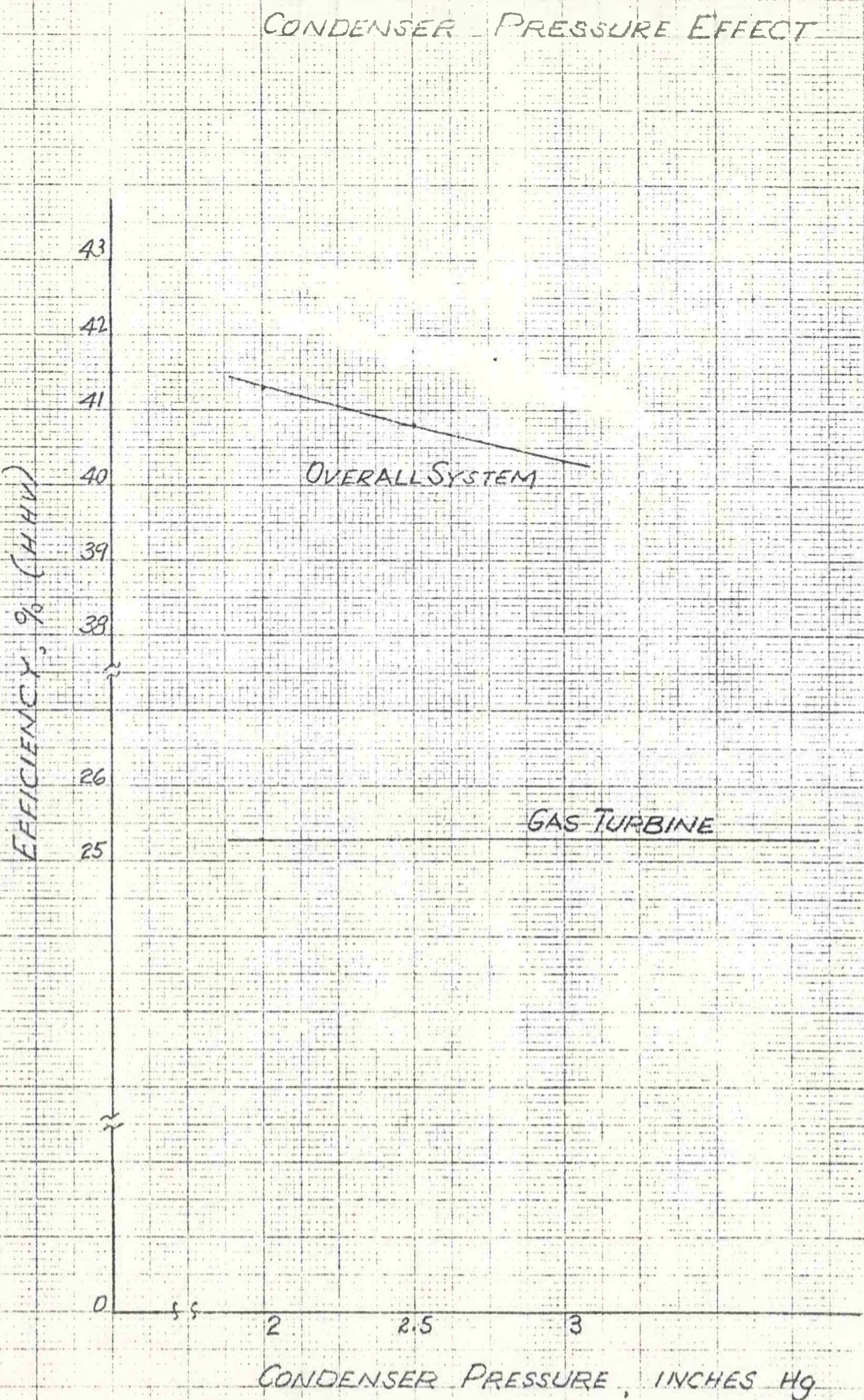

CONDENSER PRESSURE EFFECT
EFFICIENCY, % (HHV)
43
42
41
40
39
38
26
25
0
OVERALL SYSTEM
GAS TURBINE
2
2.5
3
CONDENSER PRESSURE, INCHES Hg

FIGURE: A-19

SELECTED PFB COMMERCIAL PLANT PERFORMANCE SENSITIVITY STUDIES

(GAS TURBINE OVERALL PRESSURE RATIO EFFECT)

ISO CONDITIONS

2400PSIG/1000F/1000F STEAM CYCLE

TURBINE INLET TEMPERATURE = 1600 F

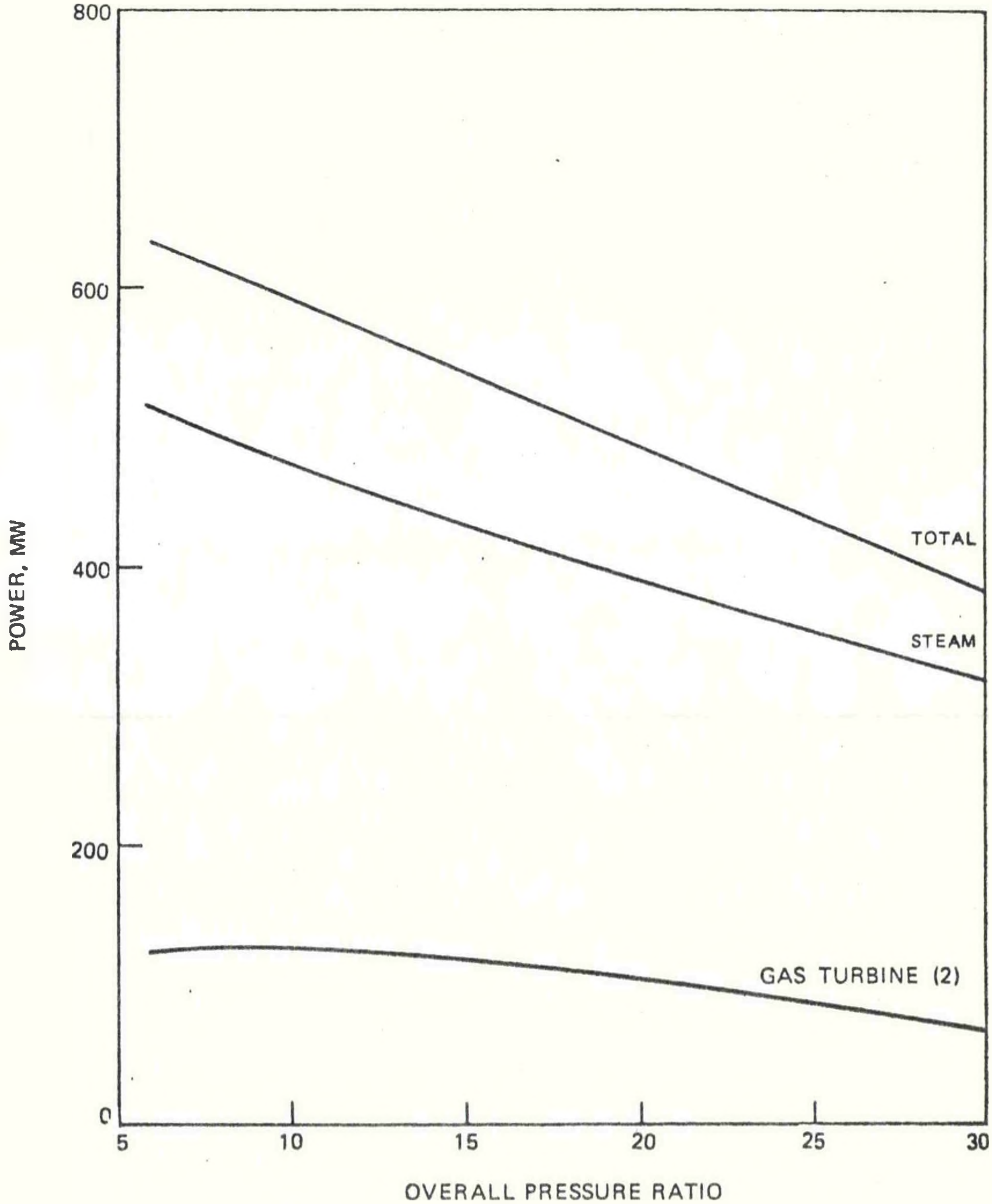


FIGURE: A-20

SELECTED PFB COMMERCIAL PLANT PERFORMANCE SENSITIVITY STUDIES
GAS TURBINE OVERALL PRESSURE RATIO EFFECT
ISO CONDITIONS
2400 PSIG /1000°F/1000°F STEAM CYCLE
TURBINE INLET TEMPERATURE = 1600°F

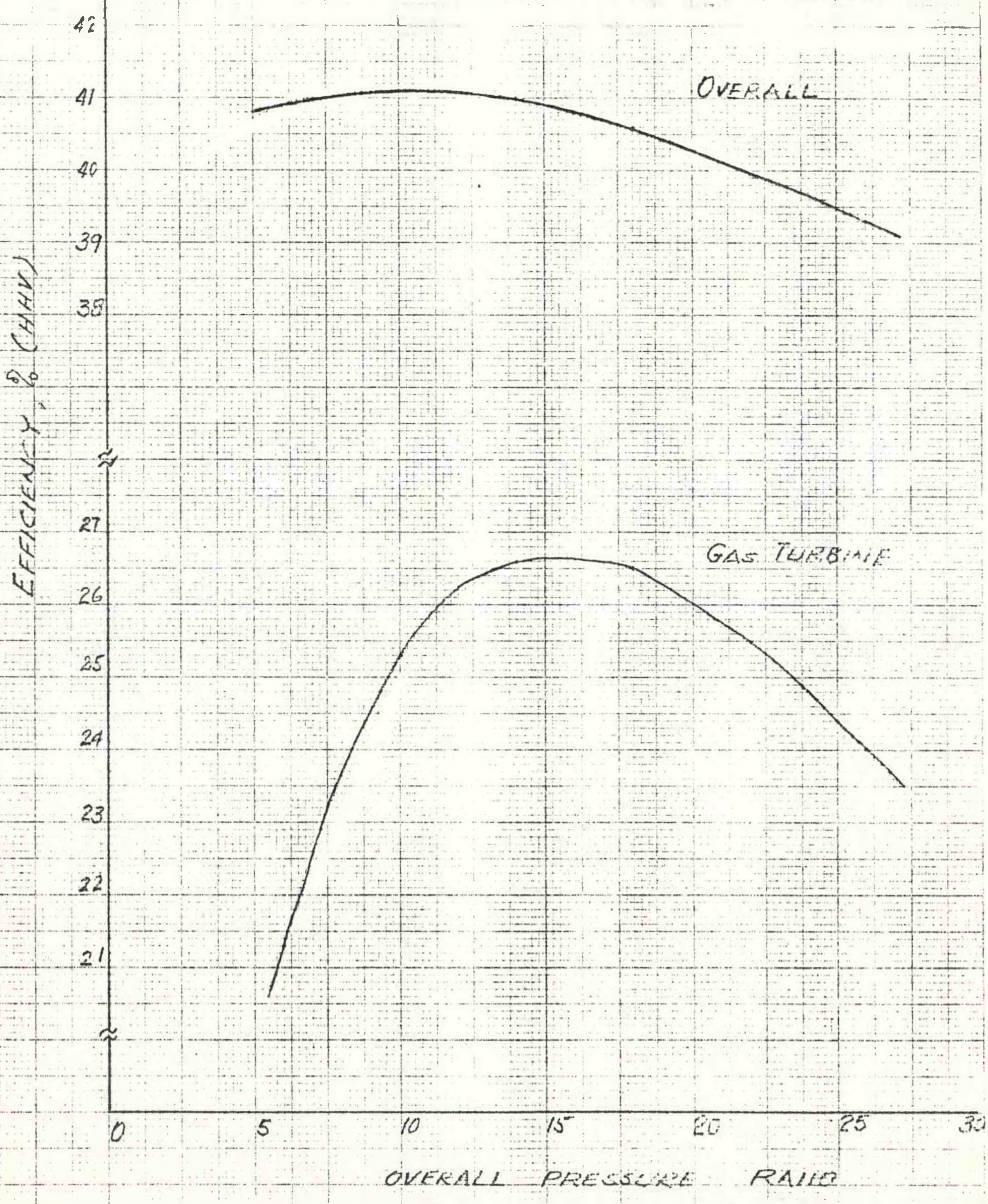


FIGURE: A-21

to date, multiple PFB/GT/WHRB modules must be used for plants over 100 MWe. Therefore, reductions in specific cost ($/kW) associated with the "economy of scale" are expected to be relatively small for Case A plant sizes over 100 MWe. However, in the small utility or industrial market, only one module would be required, and Case A should be more competitive with other alternatives which utilize a larger proportion of steam turbine to gas turbine power; therefore, a more detailed analysis of Case A is also warranted.

3.0 FABRICATION METHODS FOR PFB COMBUSTORS

The subtask 1.2 conceptual plant design utilizes two PFB combustor systems per gas turbine. Each PFB system consists of one coal feed system, one dolomite feed system, one PFB combustor, one PFB ash letdown system and one dust cleanup system which includes two Aerodyne 22000 SV-FBC Dust Collectors in series. Since the size of the combustor vessel prohibits rail shipment of fully assembled vessels alternate methods of fabricating, assembling and shipping the combustor vessel have been investigated. One method involves fabricating and assembling the PFB combustor vessel at Babcock & Wilcox's Mt. Vernon shops and shipping the complete vessel by barge to the jobsite. All necessary non-destructive testing of the vessel is done in the Mt. Vernon shops.

Another method consists of fabricating segments of the combustor vessel at Babcock & Wilcox's Barberton shops, shipping the segments by rail to the jobsite and assembling the segments in the field. Due to the size of the vessel, complete head and shell assemblies cannot be shipped by rail. Consequently, fairly elaborate field fabricating facilities are required for final machining and assembling of the vessel segments. Additional field facilities must be supplied for welding, stress relieving and non-destructive testing of the combustor vessel.

Both of the above methods of assembly call for installing the refractory and internals in the combustor at the jobsite. The estimated fabrication, transportation, and erection costs of the PFB system are shown in Table B-1 for both methods of assembly. Since the difference in cost between these assembly methods is only 1%, and since the costs themselves are only approximate, a thorough and more detailed cost estimate must be performed in order to justify the choosing of one assembly method over the other on the basis of economics. This estimate must consider such factors as the actual location of the jobsite, the modes of transportation available in the vicinity of the jobsite, the availability of local skilled craftsmen such as welders, pipe fitters and boilermakers, the wage scale of the local work force and the efficiency of the work force.

The third alternative is to use a larger number of smaller PFB combustors. Being able to ship an entire PFB combustor vessel by rail eliminates the need for costly barge unloading facilities and elaborate field fabrication facilities. Consequently, a smaller size PFB combustor has been designed in order to make use of the advantages of shipping an entire combustor vessel by rail. The arrangements and details of the smaller PFB combustor are shown in the following diagrams:

Figure B-1 Arrg't. Rail Shippable PFBC-Front View
Figure B-2 Arrg't. Rail Shippable PFBC-Side View
Figure B-3 Rail Shippable PFBC-Plan Sections Sheet 1
Figure B-4 Rail Shippable PFBC-Plan Sections Sheet 2

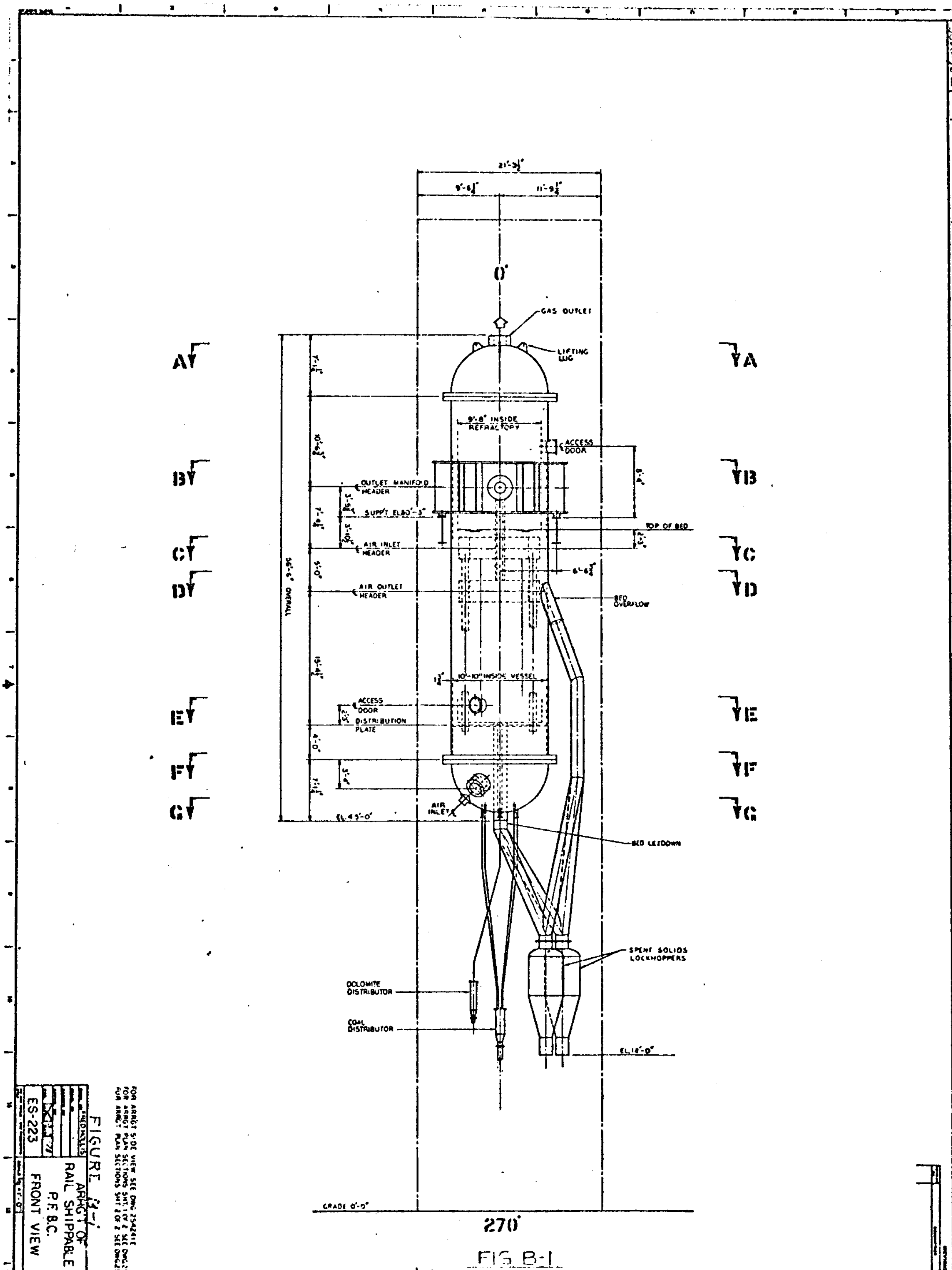


FIG B-1
ARRG'T OF RAIL SHIPPABLE P.F.B.C.
FRONT VIEW

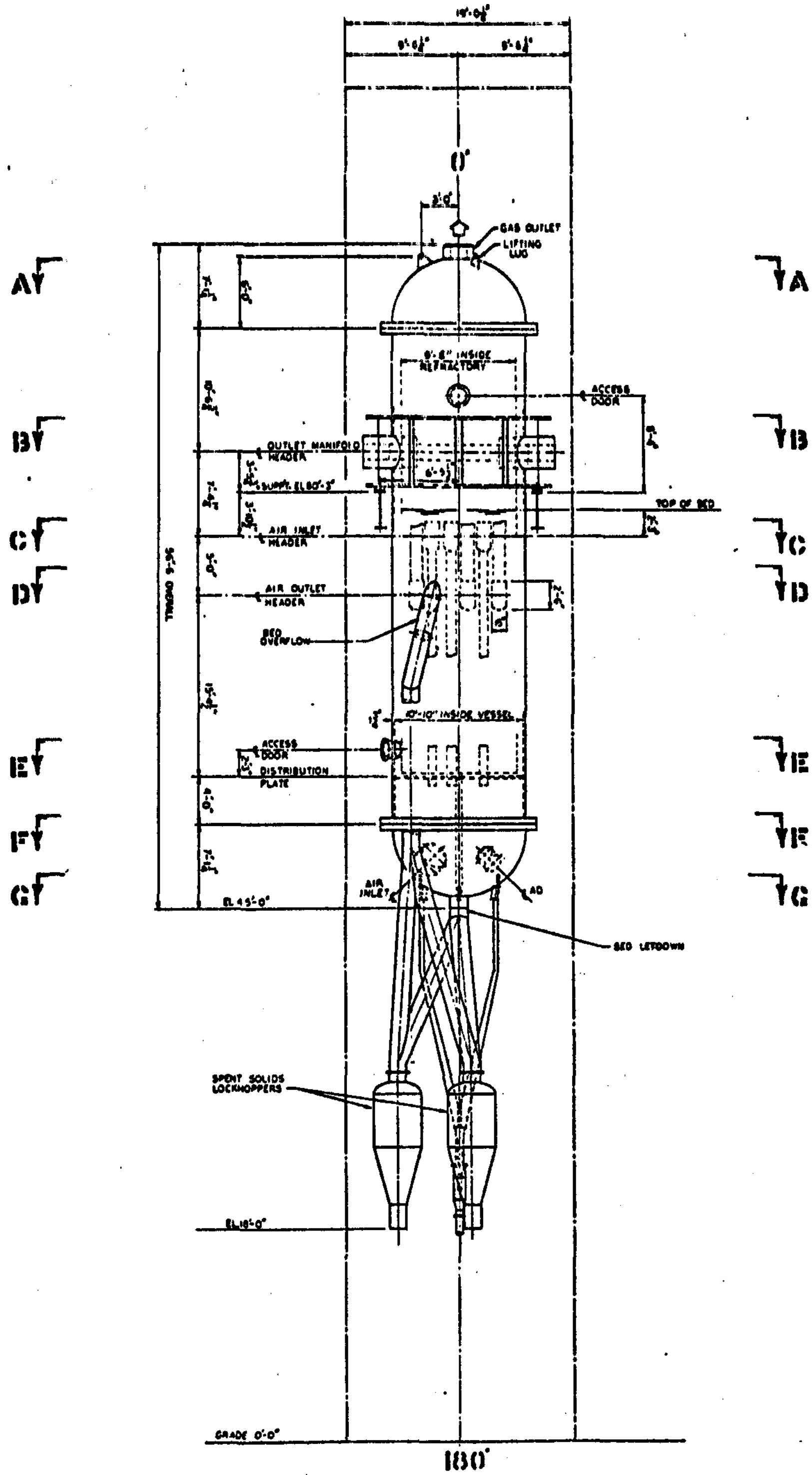
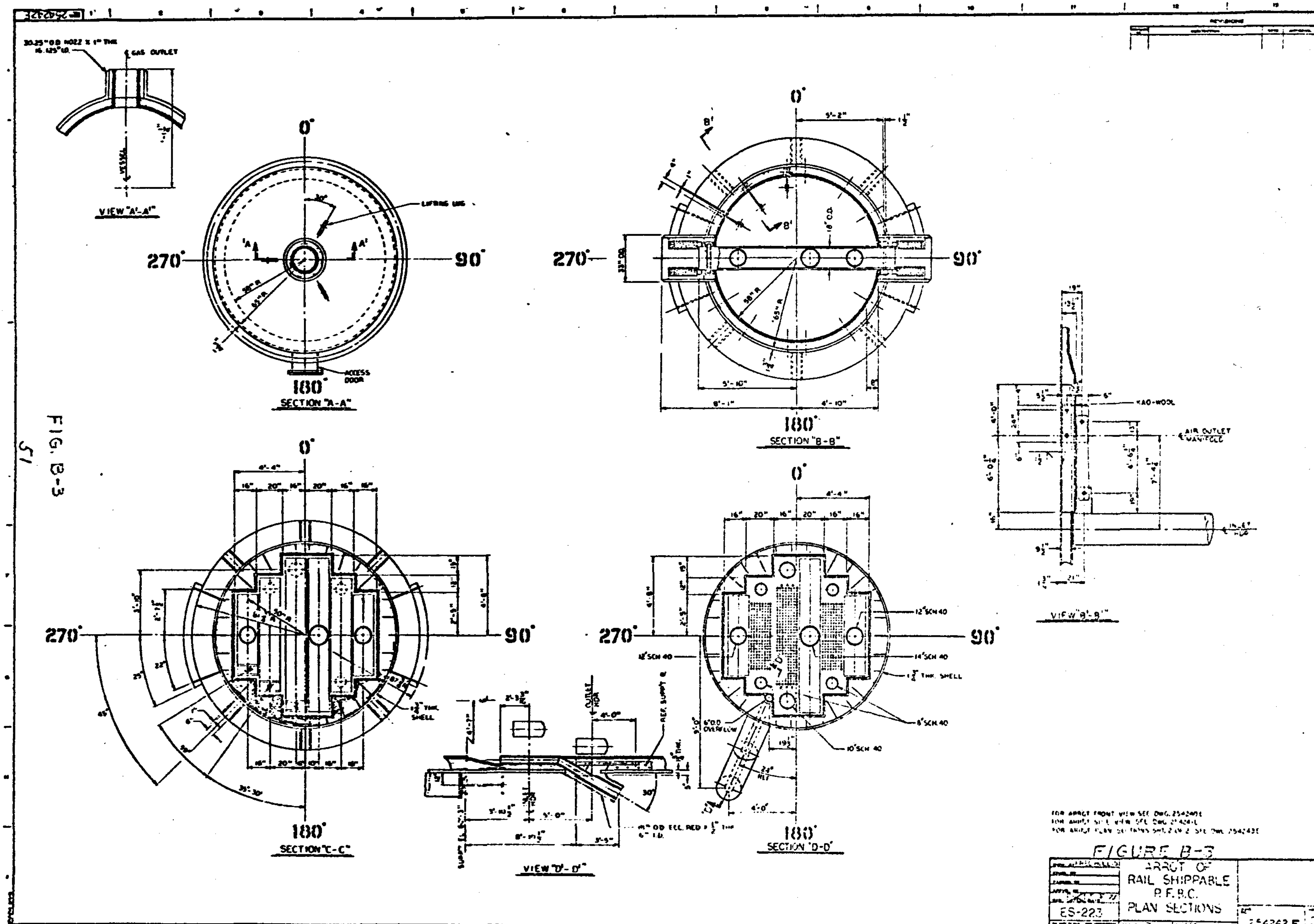


FIG. B-2
ARRG'T OF RAIL SHIPPABLE P.F.B.C.
SIDE VIEW

FIGURE B-2
ARRG'T OF RAIL SHIPPABLE P.F.B.C. SIDE VIEW
ES-223
254241 E 0



GAS OUTLET
VIEW "A'-A'"
LIFTING LUG
ACCESS DOOR
SECTION "A-A"
SECTION "B-B"
SECTION "C-C"
SECTION "D-D"
VIEW "D'-D'"
VIEW "B'-B'"
KAO-WOOL
AIR OUTLET MANIFOLD
12" SCH 40
14" SCH 40
$1\frac{3}{4}$" THK. SHELL
8" SCH 40
10" SCH 40
6" O.D. OVERFLOW
FIGURE B-3
ARRGT OF
RAIL SHIPPABLE
P.F.B.C.
PLAN SECTIONS
ES-223
254242 E

FIG. B-3

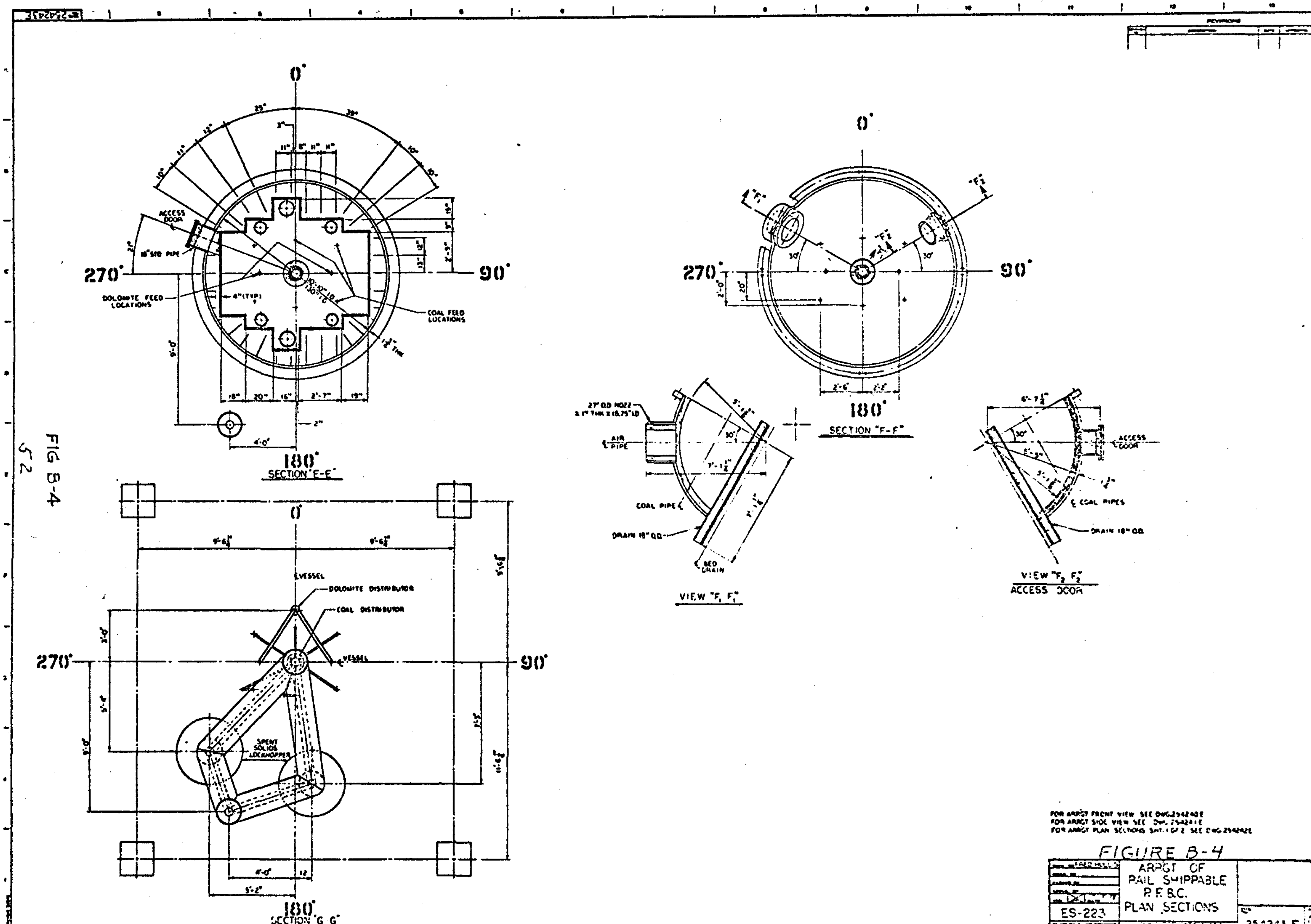

SECTION "E-E"
SECTION "F-F"
VIEW "F1 F1"
VIEW "F2 F2"
ACCESS DOOR
SECTION "G G"
0°
90°
180°
270°
ACCESS DOOR
18" STD PIPE
DOLOMITE FEED LOCATIONS
COAL FEED LOCATIONS
AIR PIPE
COAL PIPE
DRAIN 18" O.D.
BED DRAIN
COAL PIPES
VESSEL
DOLOMITE DISTRIBUTOR
COAL DISTRIBUTOR
SPENT SOLIDS LOCKHOPPER
FOR ARRGT FRONT VIEW SEE DWG 254240E
FOR ARRGT SIDE VIEW SEE DWG 254241E
FOR ARRGT PLAN SECTIONS SHT 1 OF 2 SEE DWG 254242E
FIGURE B-4
ARRGT OF RAIL SHIPPABLE P.F.B.C. PLAN SECTIONS
ES-223
254243 E
FIG B-4
52

TABLE B-1

Estimated Capital Costs Per PFB System*

PFB System	Fabrication & Transportation Costs ($)	Erection Cost ($)	Total Cost ($)
Barge Shipment	6,610,000	1,845,000	8,455,000
feed system, combustor, & ash letdown system	5,372,500	1,255,000	6,627,500
particulate removal system	1,237,500	590,000	1,827,500
Field Assembled	5,777,500	2,757,500	8,535,000
feed system, combustor ash letdown system	4,540,000	2,167,500	6,707,500
particulate removal system	1,237,500	590,000	1,827,500
Rail Shippable	10,836,600	4,352,200	15,188,800
feed system, combustors & ash letdown system	7,352,400	3,457,200	10,809,600
particulate removal system	3,484,200	895,000	4,379,200

* To obtain the estimated capital costs per plant multiply each of the cost in this table by 4.0.

The rail shippable combustor is a scaled down version of the base PFB combustor described in detail in Subtask 1.2, Section 4.3.1.5 and operates in exactly the same way (refer to Section 4.3.1.3). Because of its smaller size and the fact that it must satisfy the same design requirements as used in Subtask 1.2, ten rail shippable PFB combustors operating in parallel are needed to feed each gas turbine. It follows then that five rail shippable combustors are equivalent to one base PFB combustor. Consequently, it is possible to use the same coal feed system and dolomite feed system that fed each base PFB combustor in Subtask 1.2 to feed five of the rail shippable combustors. A schematic of the solids feed system is shown in Figure B-5. Control of the coal flow and dolomite flow to each combustor is accomplished by placing an air swept feeder valve in the feed lines as shown in Figure B-6. Placing the feeder valves as shown enables the solids feed flow of each combustor to be controlled independently of the other combustors. Consequently, a single combustor can be shutdown rather than having to shutdown all five combustors.

Particulates are removed from the gas streams exiting the rail shippable combustors by an Aerodyne SV-FBC Dust Collection System. Since the collection efficiency of the Aerodyne SV-FBC series of dust collectors is a function of gas flow rate and velocity, it follows that if all five of the rail shippable combustors are connected to a single Aerodyne dust collector system, the collection efficiency of the system will decrease if any one of the combustors is shutdown. Consequently, each combustor has its own particulate removal equipment in order to enable individual combustors to be shutdown without causing an increase in dust loading to the turbine.

Since the mass flow rate of the flue gas and particulates produced by each rail shippable combustor is 1/5 of that produced by the base PFB combustor, the dust cleanup system for each rail shippable combustor will be smaller in size and capacity than that of the base PFB combustor. As a result, the dust cleanup system for each rail shippable combustor includes two Aerodyne 4500 SV-FBC Dust Collectors operating in series. The design and operation of the Aerodyne 4500 SV-FBC Dust Collector (Figure B-7) is the same as that of the Aerodyne 22000 SV-FBC Dust Collector which is used to remove the particulates from the flue gas of the base PFB combustor in Subtask 1.2 (refer to Subtask 1.2 Report, Section 4.3.9). The predicted collection efficiency of the Aerodyne 4500 SV-FBC Dust Collector is shown in Figure B-8 and the predicted performance of two of these collectors operating in series is shown in Table B-2 for a Ca/S ratio of 1.0 and in Table B-3 for a Ca/S ratio of 3.0.

The rail shippable PFB system consists of one coal feed system, one dolomite feed system, five rail shippable combustors, five PFB ash letdown systems, and five dust cleanup systems which include 10 Aerodyne 4500 SV-FBC Dust Collectors. There are two PFB systems per gas turbine and the estimated fabrication, transportation, and erection costs of each system are shown in Table B-1. The rail shippable combustor vessels are fabricated at Babcock & Wilcox's Barberton shops where all necessary non-destructive testing is done. The refractory and internals are installed at the jobsite.

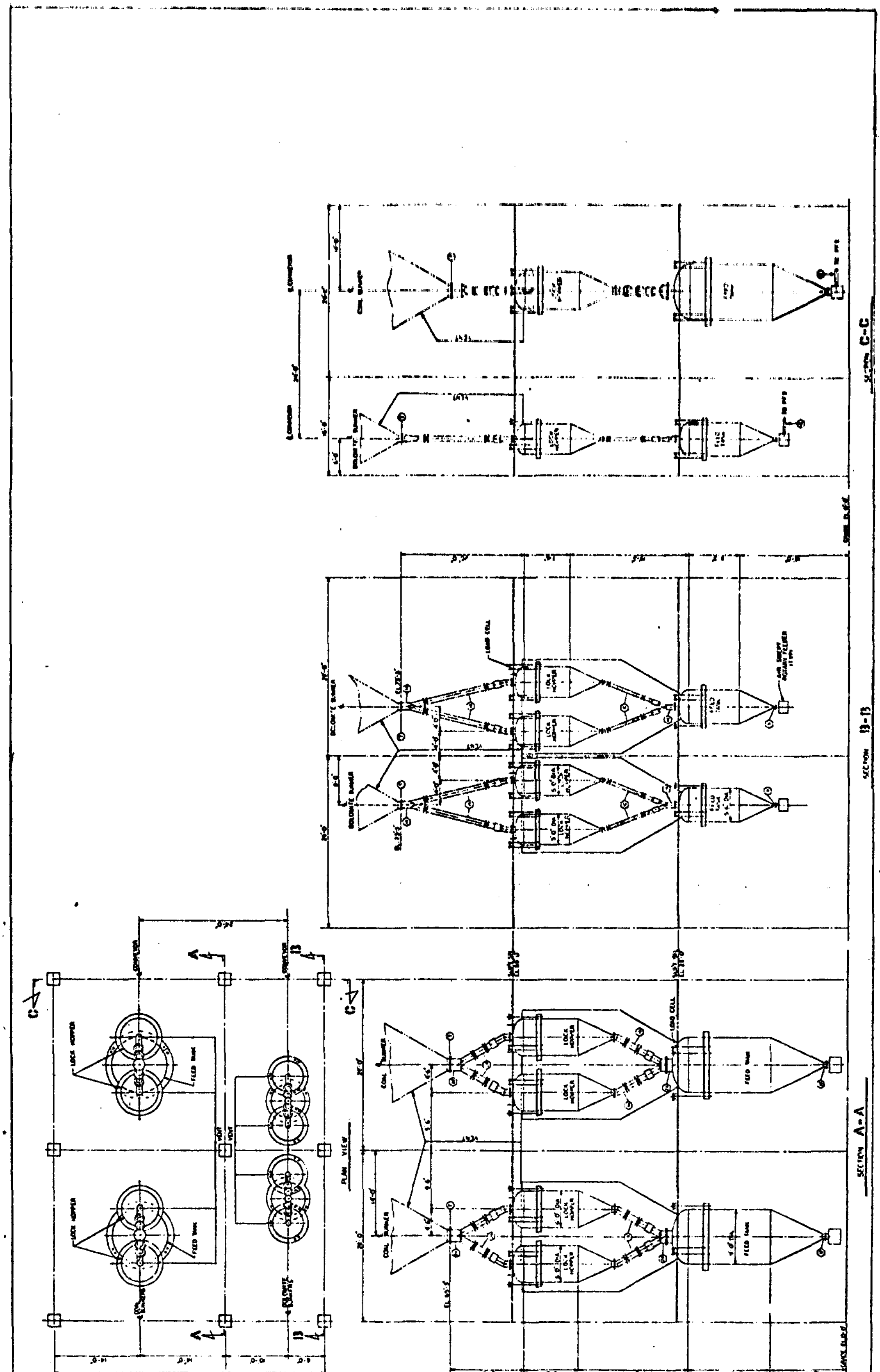
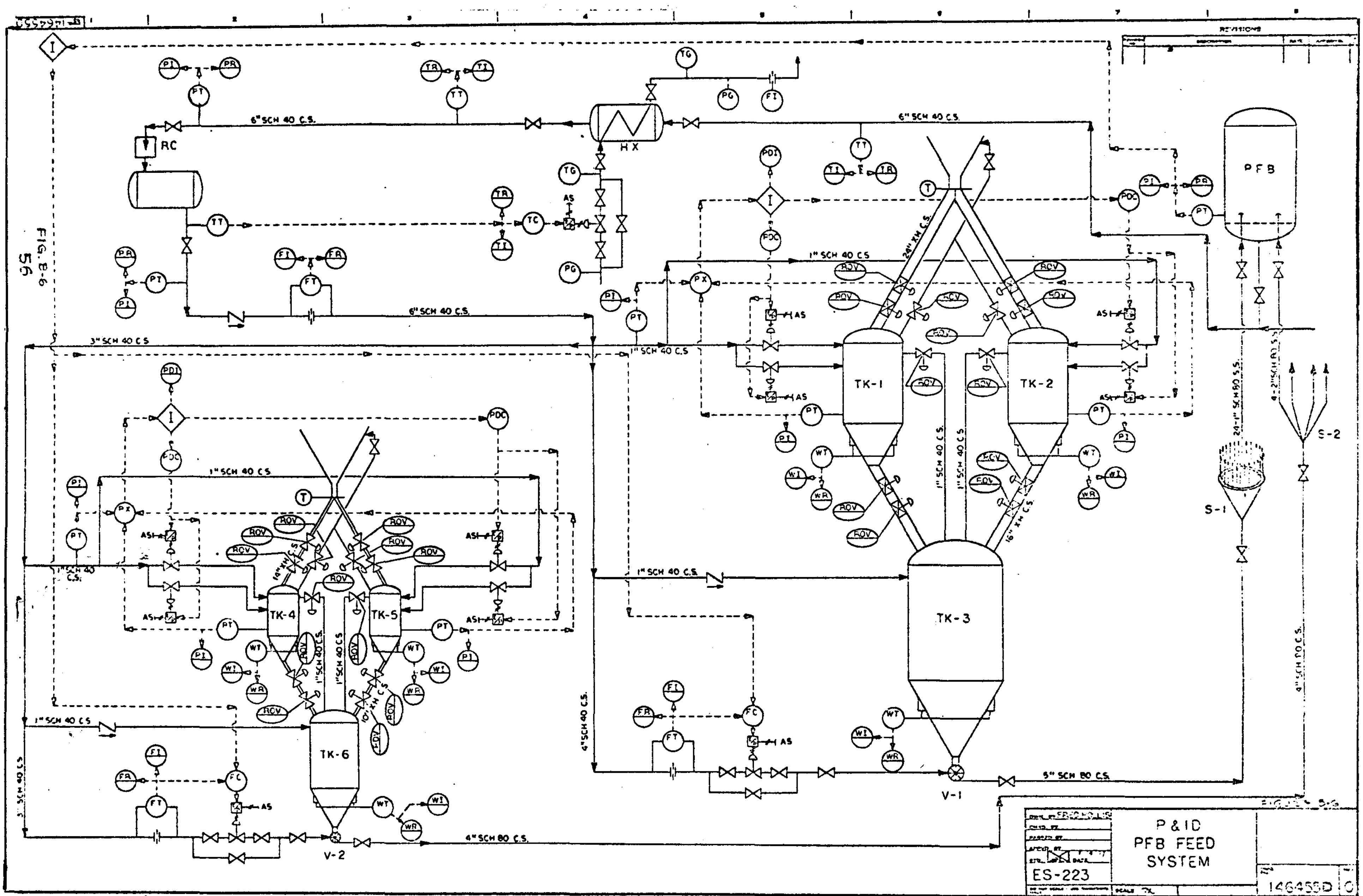
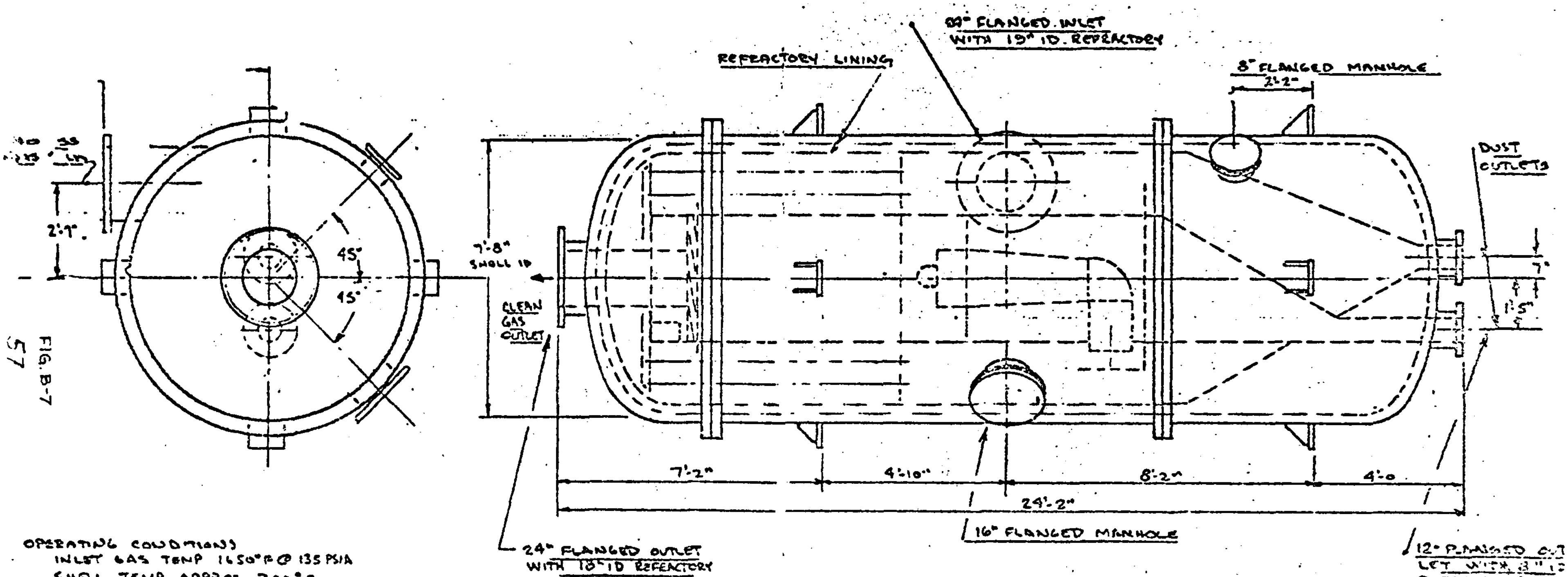

PLAN VIEW
LOCK HOPPER
FEED TANK
COAL BURNER
DOLOMITE BURNER
LOAD CELL
SECTION A-A
SECTION B-B
SECTION C-C
FIGURE B-5

FIG B-5
55


P & ID
PFB FEED
SYSTEM
ES-223
146455D
FIG. E-6
56
PFB
HX
RC
TK-1
TK-2
TK-3
TK-4
TK-5
TK-6
V-1
V-2
S-1
S-2
6" SCH 40 C.S.
3" SCH 40 C.S.
1" SCH 40 C.S.
4" SCH 40 C.S.
5" SCH 80 C.S.
4" SCH 80 C.S.
24" XH C.S.
16" XH C.S.
14" XH C.S.
10" XH C.S.
REVISIONS



OPERATING CONDITIONS
INLET GAS TEMP 1650°F @ 135 PSIA
SHELL TEMP APPROX 200°F

ALL DIMENSIONS ARE APPROXIMATE AND SUBJECT TO FINAL DESIGN CHANGES

APPROX. WEIGHT 46 TONS

FIGURE B-7

NO.	REVISION	DATE	BY

AERODYNE DEVELOPMENT CORP.
CLEVELAND, OHIO

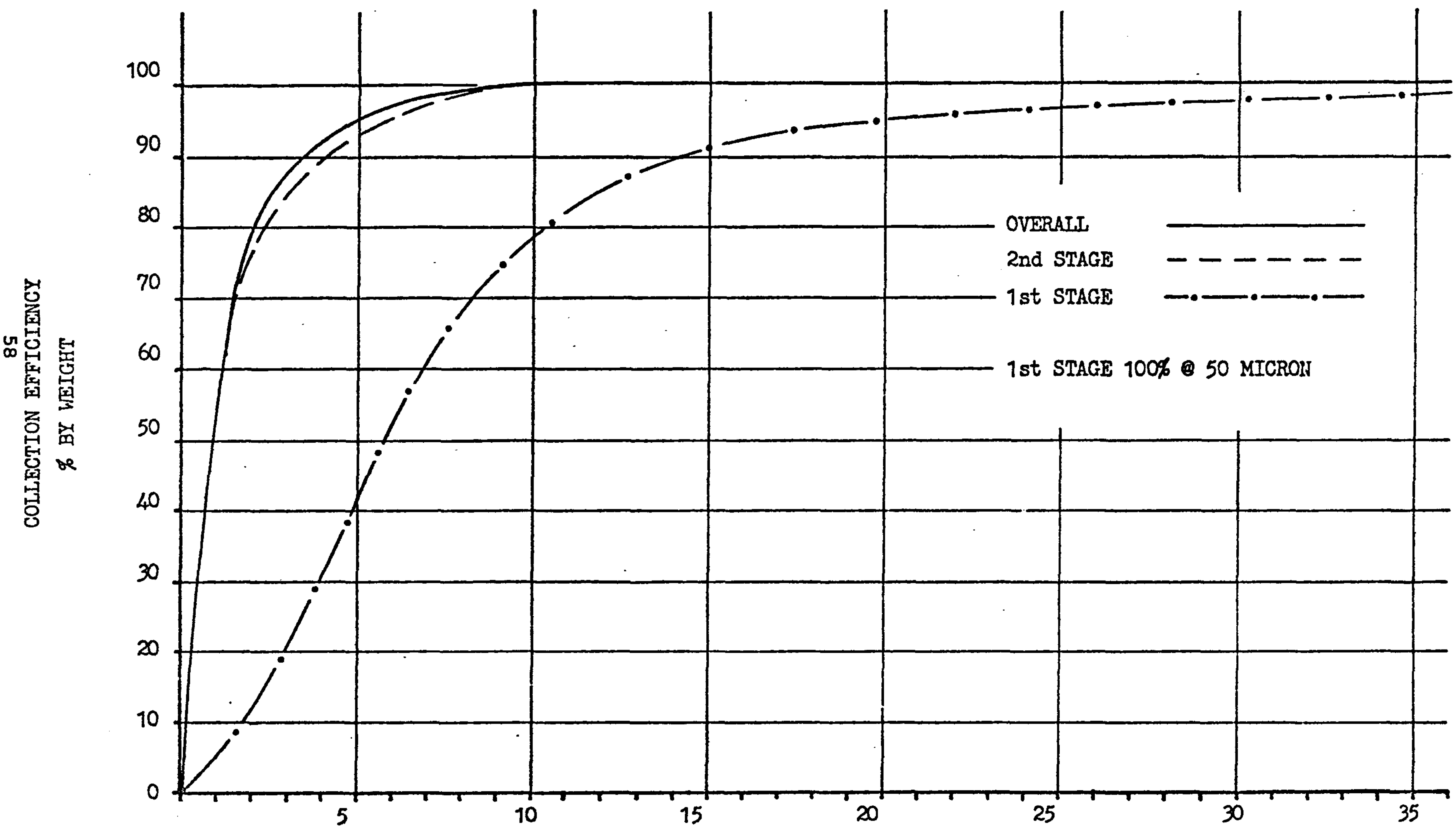
SCALE: 3/8"=1'-0" | APPROVED BY: | S. O. NO.: | DRAWN BY: WS
DATE: 2-16-78 | REVISED:

MODEL PV 4500 SV COLLECTOR

DRAWING NUMBER SC 1272

FEB 16 1978

AERODYNE SV-FBC SEPARATOR
EFFICIENCY CURVES



PARTICLE SIZE (MICRONS)

FIGURE: B-8

TABLE B-2

Particulates Removed in the Aerodyne 4500 SV-FBC System
For A Calcium/Surfur Ratio of 1.0

*Dust flow entering the first collector = 1294.8 lbm/hr

	First Collector	Second Collector
Particulates removed in first stage (lbm/hr)	1058.3	5.536
Particulates removed in second stages (lbm/hr)	198.28	21.542
Dust flow leaving the collector (lbm/hr)	38.22	11.142
Dust concentration entering the turbine (grains/SCF)	0.0710	0.0207

*Particle distribution entering the turbine

particle diameter, d (microns)	particle concentration (grains/SCF)
d<2.0	0.0176
2.0<d<10.0	0.0031
d>10.0	0.000

*These flow rates are for each PFB combustor.

TABLE B-3

Particulates Removed in the Aerodyne 4500 SV-FBC System
For a Calcium/Sulfur Ratio of 3.0

*Dust flow entering the first collector = 2074.47 lbm/hr

	First Collector	Second Collector
Particulates removed in first stage (lbm/hr)	1834.55	5.562
Particulates removed in second stage (lbm/hr)	201.668	21.548
Dust flow leaving the collector (lbm/hr)	38.252	11.142
Dust concentration entering the turbine (grains/SCF)	0.0710	0.0207

*Particle distribution entering the turbine

particle diameter, d (microns)	particle concentration (grains/SCF)
d<2.0	0.0176
2.0<d<10.0	0.0031
d>10.0	0.0000

*These flow rates are for each PFB combustor.

As can be observed in Table B-1, the cost of the rail shippable PFB system is considerably more expensive than the cost of the other PFB systems. This is primarily due to the increased number of combustors and cyclone dust collectors that are used in this system. Because of the economic disadvantages of the rail shippable PFB system, it was eliminated from further consideration.

To avoid installation of a costly barge unloading system it was decided to use field assembled PFB Combustors with shop fabricated and rail shipped sections.

4.0 PARTICULATE REMOVAL SYSTEMS

At present, empirical information regarding the particle size distribution of the solids elutriated from a PFB combustor is unavailable. Consequently, assumptions were made in order to establish the size distribution of the particulates entering the gas cleanup equipment.

The size distribution of the sulfur sorbent elutriated from the bed is based on the size distribution of the stone fed to the bed. To account for abrasion and thermal decrepitation in the bed, a 20% reduction in this size distribution is assumed; the resulting size distribution is shown in Figure C-1. A terminal settling velocity analysis indicates that particles less than 300 micron size are carried out of the PFB combustor. This results in an elutriation rate of 35% for the spent sorbent. It should be noted that based on these assumptions, less than 1/2% of the elutriated spent sorbent has a size of less than 10 microns.

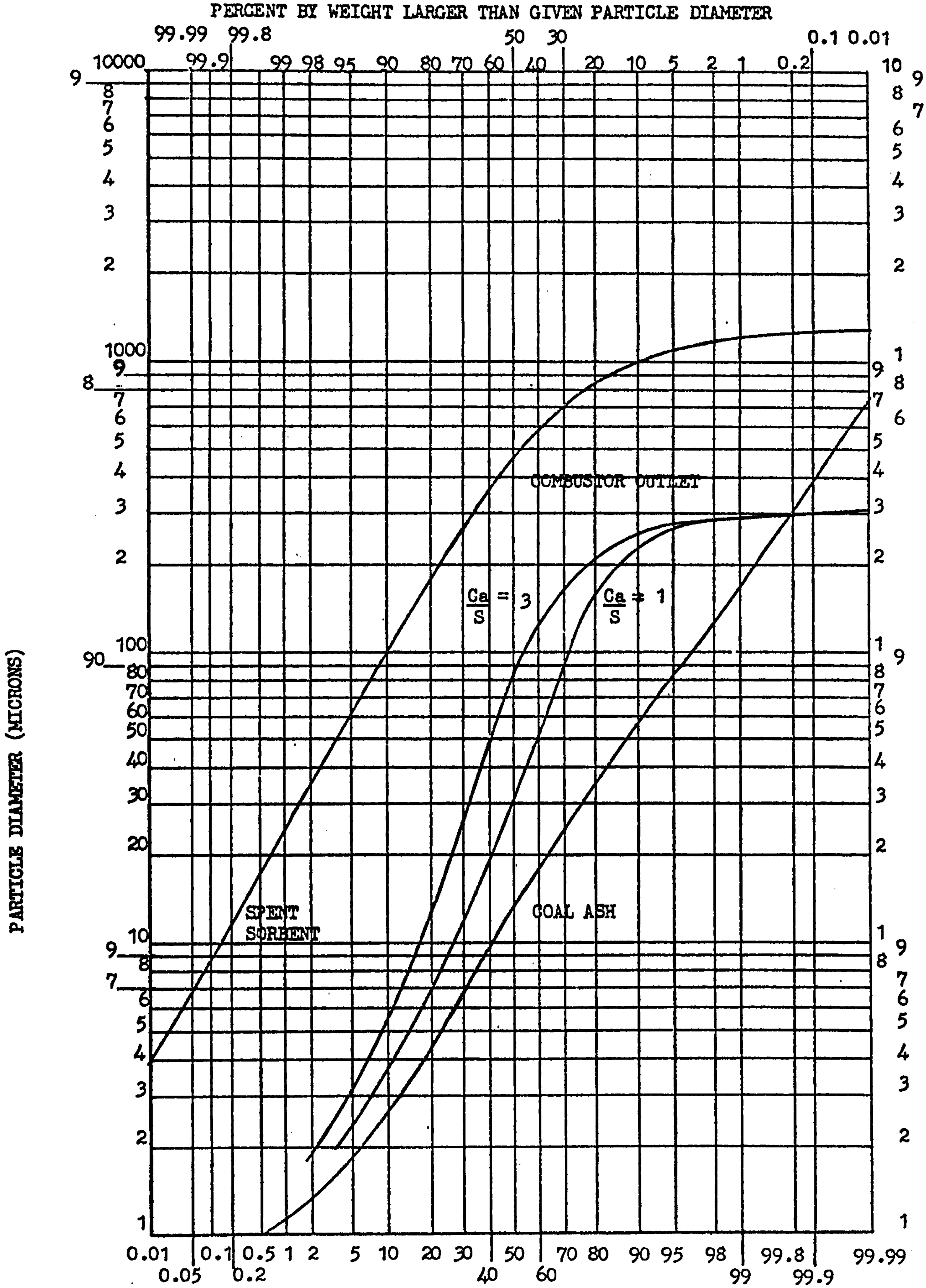
The size distribution of the coal ash is assumed to be the same as the fly ash size distribution leaving a pulverized coal or stoker fired boiler. This assumption is equivalent to essentially all the coal ash being elutriated from the bed with nearly 40% being less than 10 micron size.

In designing the particulate removal system two conditions are specified. The expected operating conditions are based on a Ca/S molar feed ratio of 1.0. The design conditions are based on a Ca/S ratio of 3.0. These conditions are shown in Tables C-1 and C-2 with the corresponding size distributions being shown in Figure C-1. Because of the assumptions for the size distribution of the spent sorbent, the dust in the less than 10 micron size range is essentially all coal ash and therefore the dust loading in this size range is the same for both conditions. The higher Ca/S ratio, then, primarily influences the design of the initial separator stages and the design of the ash let down system.

Because of a lack of actual operating experience with PFB exhaust gases in a gas turbine, the allowable level of particulate concentration in the gas entering the turbine has not been established. On the basis of limited data (Ref.3) an estimate of allowable gas turbine particulate loading has been made showing that particles greater than 10 microns in size give unsatisfactory turbine life, particles less than 2 microns in size have negligible effects, and that some limited amount of particulate in the 2-10 micron size can be tolerated within the gas turbine. Therefore, the performance requirements for the particulate removal system are based on the following dust loading entering the gas turbine:

particle diameter, d (microns)	max. particulate concentration (grains/SCF)
$d<2.0$	no limit
$2.0<d<10.0$	0.0100
$d>10.0$	0.0000

Additional design requirements for the equipment are shown in Table C-3.



PARTICLE SIZE DISTRIBUTION ENTERING THE GAS CLEANUP EQUIPMENT

FIGURE: C -1

TABLE C-1

Operating Conditions for Ca/S = 1.0

Collector Inlet Gas Analysis

Component	lbm/hr	moles/hr
O_2	11344.	354.35
N_2	282976.	10101.
Ar	5055.5	126.7
SO_2	475.5	7.45
CO_2	89678.	2038.
H_2O	16002.	888
Total	405531.	13515.5

Collector Inlet Gas Molecular Weight	30.005 lbm/mole
Collector Inlet Gas Temperature	1650°F
Collector Inlet Gas Pressure	136.0 psia
Collector Inlet Gas Density	.1803 lbm/ft
Collector Inlet Dust Flow	6474 lbm/hr
Collector Inlet Dust Concentration	.01596 lbm/dust/lbm wet gas

Collector Inlet Particle Size Distribution

particle diameter (microns)	% by weight >stated particulate diameter
100	27.46
80	31.53
60	36.18
40	43.27
20	57.75
10	73.02
8	76.90
6	82.06
4	88.48
2	96.16

Clean Compressed Air Flow	1085508 lbm/hr
Clean Compressed Air Temperature	1578°F
Clean Compressed Air Pressure	136.0 psia
Clean Compressed Air Density	.1803 lbm/ft

TABLE C-2

Design Conditions for Ca/S = 3.0

Collector Inlet Gas Analysis

Component	lbm/hr	moles/hr
O_2	11344.0	354.35
N_2	282976.0	10101.00
Ar	5055.5	126.7
SO_2	475.5	7.45
CO_2	92661.0	2105.5
H_2O	16144.0	896.1
Total	408656.0	13591.1

Collector Inlet Gas Molecular Weight	30.068 lbm/mole
Collector Inlet Gas Temperature	$1650^{o}F$
Collector Inlet Gas Pressure	136.0 psia
Collector Inlet Gas Density	.1807 lbm/ft
Collector Inlet Dust Flow	10372.35 lbm/hr
Collector Inlet Dust Concentration	.02538 lbm/dust/lbm wet gas

Collector Inlet Particle Size Distribution

particle diameter (microns)	% by weight > stated particulate diameter
100	43.77
80	49.33
60	54.94
40	61.78
20	72.90
10	83.01
8	85.50
6	88.76
4	92.80
2	97.61

Clean Compressed Air Flow	1085508 lbm/hr
Clean Compressed Air Temperature	$1578^{o}F$
Clean Compressed Air Pressure	136.0 psia
Clean Compressed Air Density	.1803 lbm/ft

TABLE C-3

Design Requirements

1. The maximum allowable unrecoverable pressure loss that can exist between the inlet and outlet of the dust collection equipment is 4.00 psi.

2. All insulation is to be located adjacent to the inside surface of the exterior walls of the dust collection equipment.

3. The metal temperature of the outside surface of the exterior walls of the dust collection equipment is to be maintained at 250^{o}F when the ambient air temperature is 80^{o}F and the flue gas temperature is 1650^{o}F.

4. Each pressurized fluidized bed combustor is to have its own dust collection system.

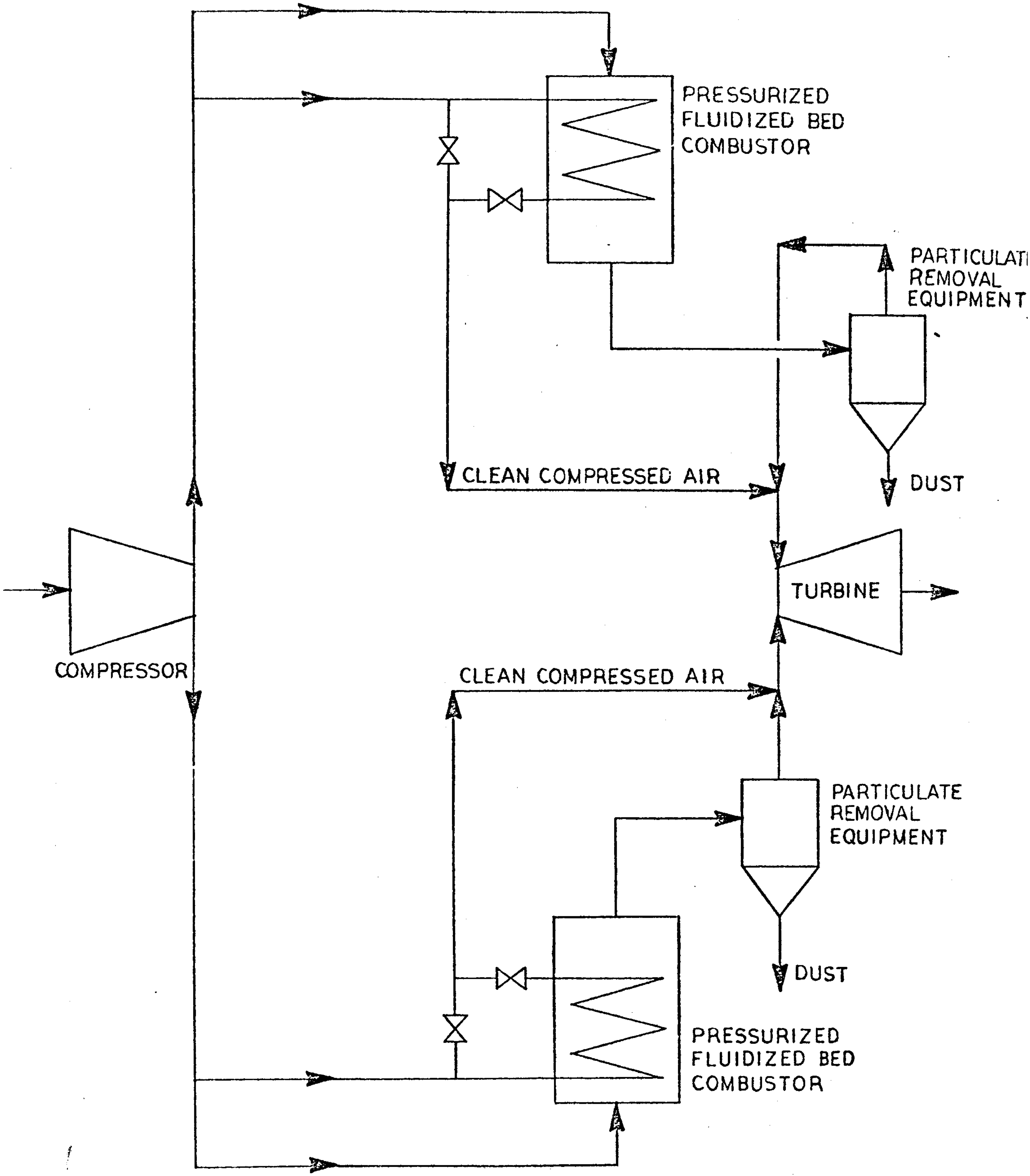
The location of the particulate removal equipment is shown schematically in Figure C-2. With the chosen concept for the PFB combustor, the particulate removal equipment need only accommodate approximately 25% of the total gas flow entering the gas turbine. Since the size and cost of this equipment is greatly influenced by the gas volume, this design concept helps minimize the equipment cost.

It should however be noted that the required particulate removal efficiency is only a function of the solids flow from the bed (i.e., a function of fuel flow) and the permissible solids flow to the gas turbine and is not a function of the proportion of the total gas flow that must be cleaned up. The air cooled PFB cycles consume less fuel per unit of turbine gas flow than do the steam cooled PFB cycles and hence require lower particulate removal efficiency for the same absolute turbine limits. The split flow air cooled cycle considered here requires the same removal efficiency as the excess air cooled cycle; but the former cycle has a smaller volume of gas to be cleaned.

Various particulate removal systems (conventional cyclones, high efficiency cyclones, and granular bed filters) have been investigated as part of the Trade Off Studies performed under Subtask 1.3. The results of this investigation indicate that conventional cyclones are incapable of removing all particulates greater than 10 micron diameter and are relatively inefficient in the range of 2 to 10 microns. Consequently, these devices are unable to meet the performance requirements for particulate removal equipment.

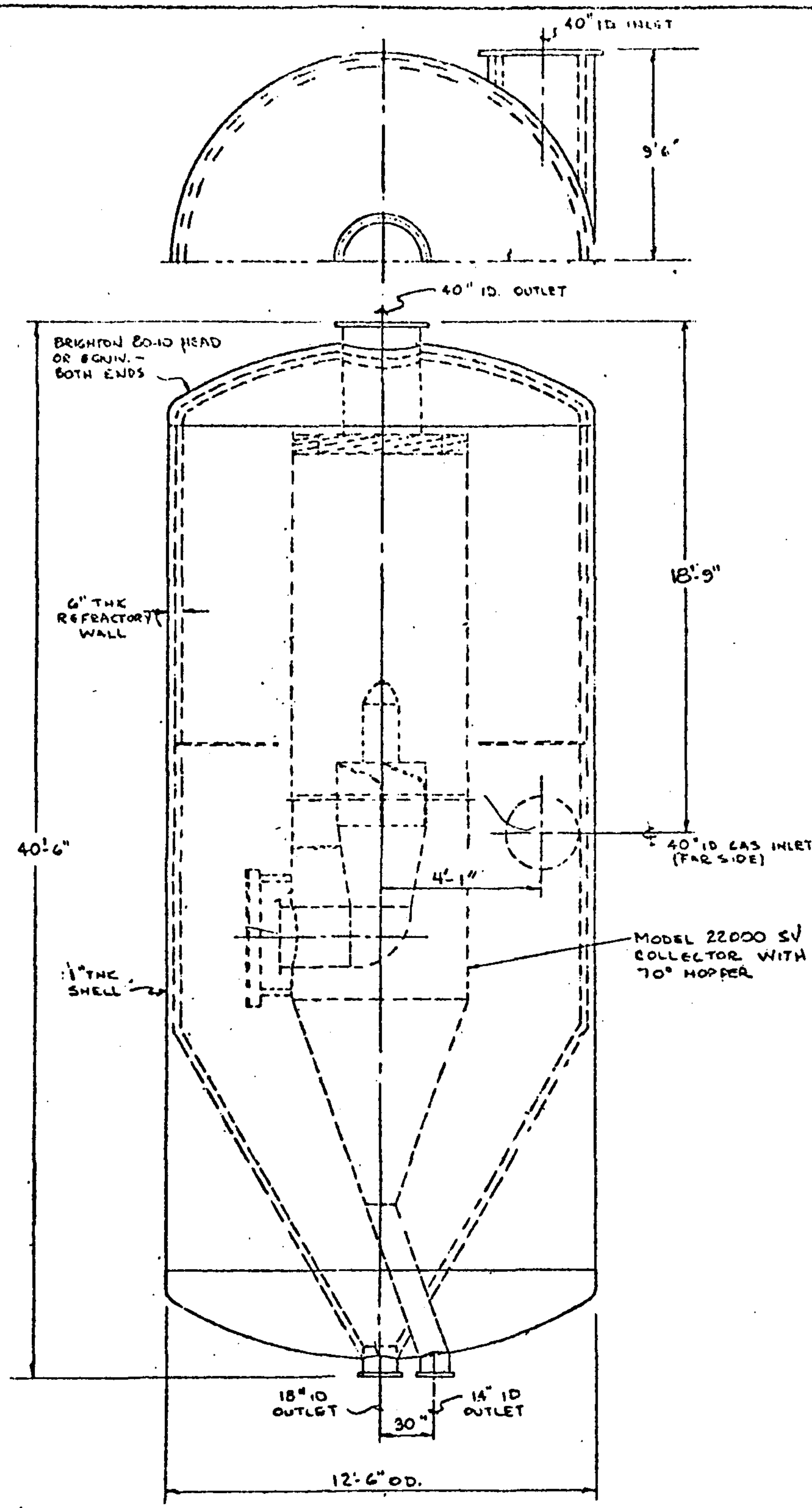
The high efficiency cyclones that were investigated are Aerodyne Development Corporation's "SV-FBC" Series of Dust Collectors and Donaldson Company's "Tan-Jet" System. The particular "SV-FBC" Dust Collector that has been studied is the Model 22000SV as shown in Figure C-3. Its design is an extension of the Aerodyne equipment presently used in low temperature, low pressure applications. The major modification consists of placing the collector inside a refractory lined pressure vessel which becomes an initial stage cyclone dust collector. This results in a single unit having two stages of collection; the second stage being based on the existing Aerodyne Series "SV" Dust Collector (License: System Siemens). The predicted collection efficiency of this two stage collector is shown in Figure C-4. Calculations indicate that two of these units operating in series would be required for each PFB combustor. The predicted performance is shown in Table C-4, for a Ca/S ratio of 1.0 and in Table C-5 for a Ca/S ratio of 3.0. The dust loading entering the gas turbine in the critical 2 to 10 micron size range is projected to be 1/3 of the allowable level.

The "Tan-Jet" system of Donaldson is essentially a high efficiency multiclone which employs a secondary stream of clean compressed air to generate a strong vortex in each of the multiclone tubes (Figures C-5, & C-6). The secondary air stream is supplied at a static pressure greater than that of the dirty gas stream by means of a booster compressor. Literature supplied by Donaldson indicates that the "Tan-Jet" system is capable of removing all particulates greater than 10 micron diameter. However, the predicted collection efficiency of the "Tan-Jet" system in the range of less than 4 microns



LOCATION OF DUST COLLECTION EQUIPMENT

FIGURE: C -2



VESSEL ELEVATION

SHELL THICKNESS = $\frac{PR}{5S - 0.6P} = \frac{(150)(6.25)(12)}{(17.5K)(.7) - 0.6(150)} = .93"$; USE 1"

SHELL MATL. - A515 GR 70 FLG. GR ST'L.

FLANGES - ASA 300# RATED

OPERATING CONDITIONS 150 PSIG 1650°F

DIMENSIONS SUBJECT TO CHANGE DURING FINAL DESIGN

Approx. Wt. = 185,000 lbs

FIGURE C-3

AERODYNE DEVELOPMENT CORP. CLEVELAND, OHIO			
SCALE None			W.A.M.
DATE 5-22-77			
MODEL 22000 SV PRESSURE VESSEL			
TYPE 3 ~ CW ROTATION			SV-22000-91H

NO.	REVISION	DATE	BY

FIG. C-3

AERODYNE SV-FBC SEPARATOR
EFFICIENCY CURVES

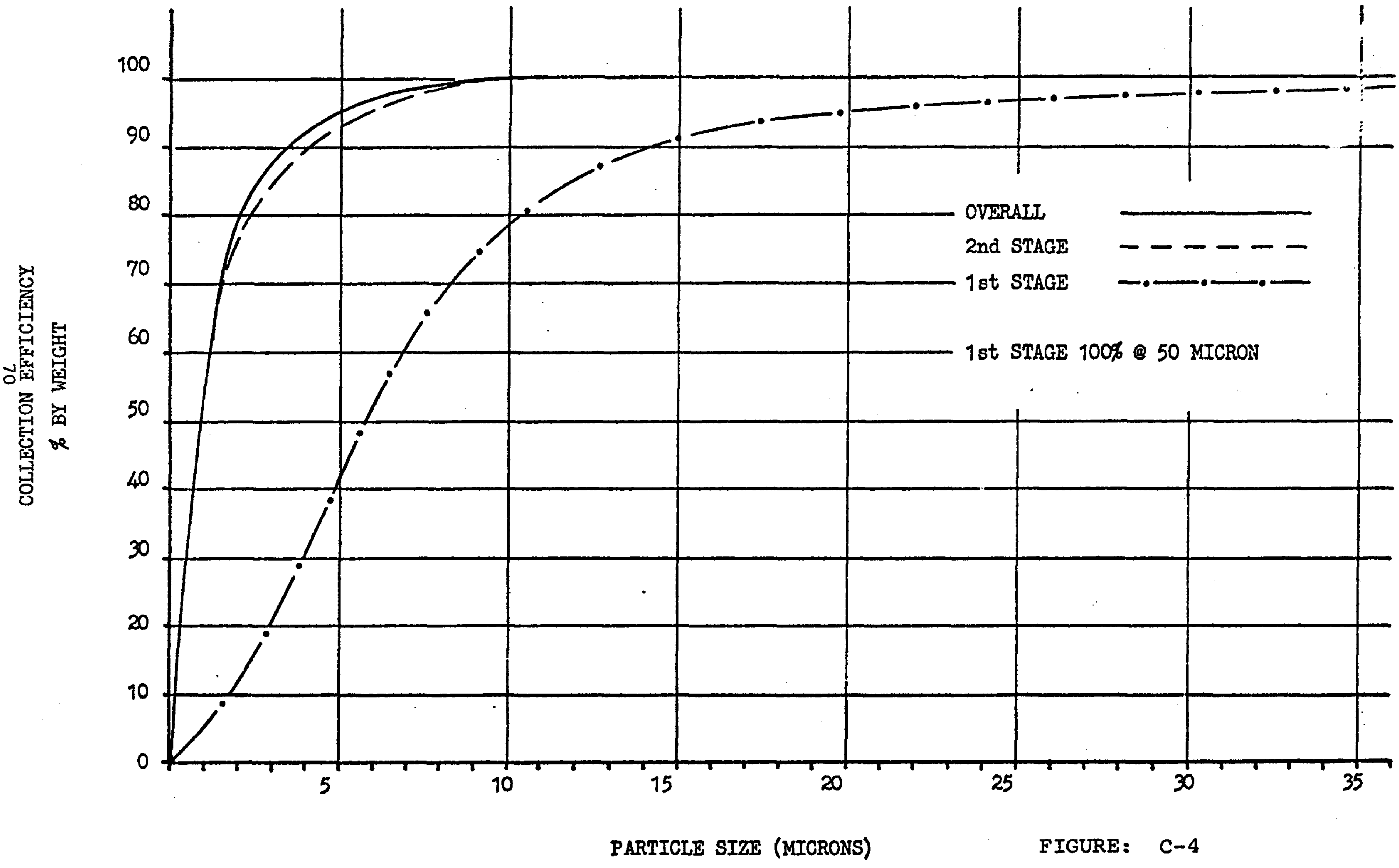


FIGURE: C-4

TABLE C-4

Projected Removal of Particulates in the Aerodyne 22000 SV FBC System for a Calcium/Sulfur ratio of 1.0

* Dust flow entering the first collector = 6474 lbm/h

	First Collector	Second Collector
Particulates removed in first stage (lbm/h)	5291.5	27.68
Particulates removed in second stage (lbm/h)	991.4	107.71
Dust flow leaving the collector (lbm/h)	191.1	55.71

	Concentration (grains/SCF)
Dust entering the turbine	0.0207
Particle distribution entering the turbine	
Particle diameter, d (microns)	
$d < 2.0$	0.0176
$2.0 < d < 10.0$	0.0031
$d > 10.0$	0.0000

*This flow rate is for each PFB Combustor.

TABLE C-5

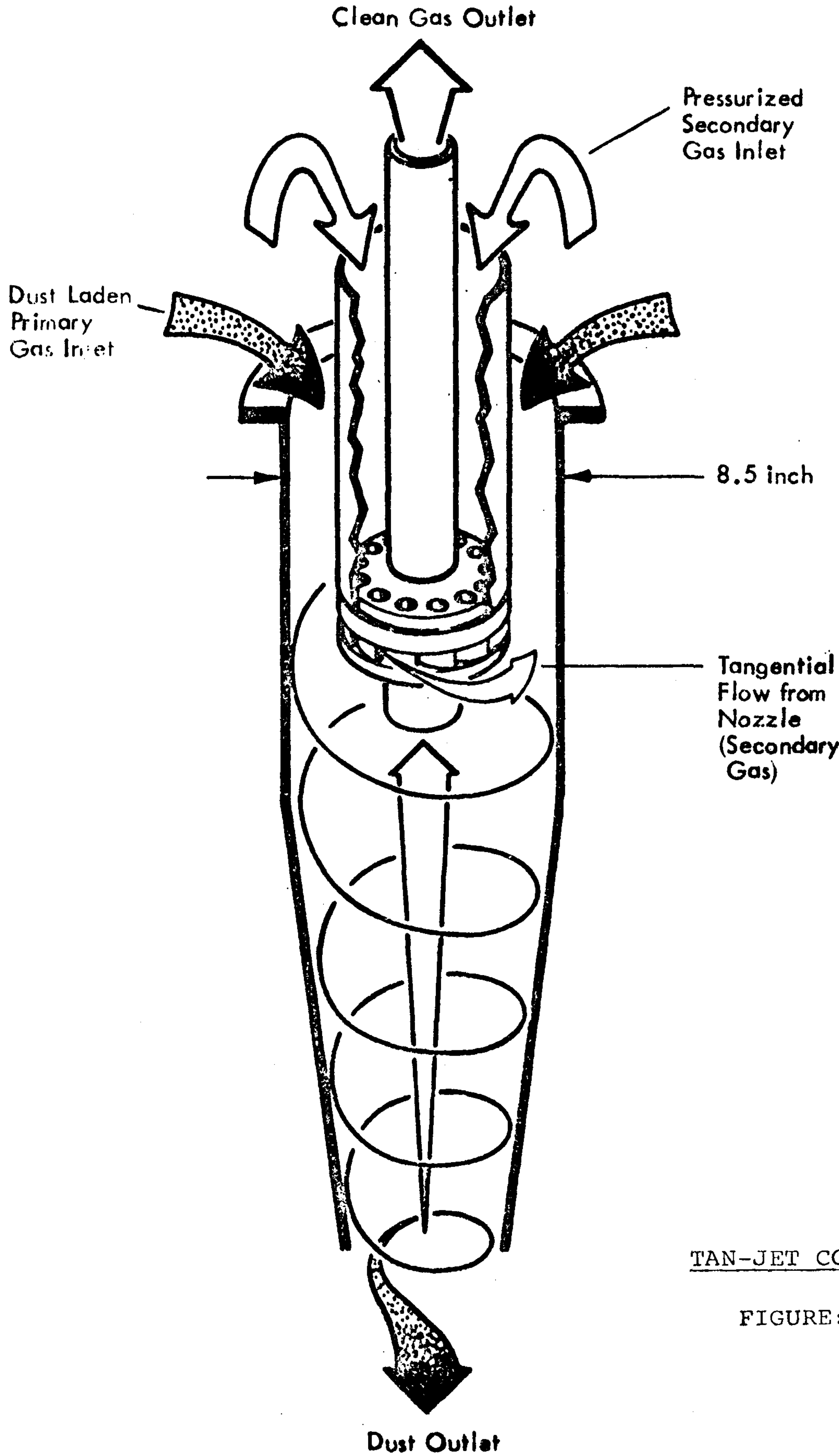
Projected Removal of Particulates in the Aerodyne 22000 SV FBC System for a Calcium/Sulfur Ratio of 3.0

*Dust flow entering the first collector = 10372.35 lbm/h

	First Collector	Second Collector
Particulates removed in first stage (lbm/h)	9172.75	27.81
Particulates removed in second stage (lbm/h)	1008.34	107.74
Dust flow leaving the collector (lbm/h)	191.26	55.71

	Concentration (grains/SCF)
Dust entering the turbine	0.0207
Particle distribution entering the turbine	
Particle diameter, d (microns)	
$d < 2.0$	0.0176
$2.0 < d < 10.0$	0.0031
$d > 10.0$	0.0000

*This flow rate is for each PFB Combustor.



TAN-JET COLLECTOR

FIGURE: C-5

TAN-JET COLLECTOR

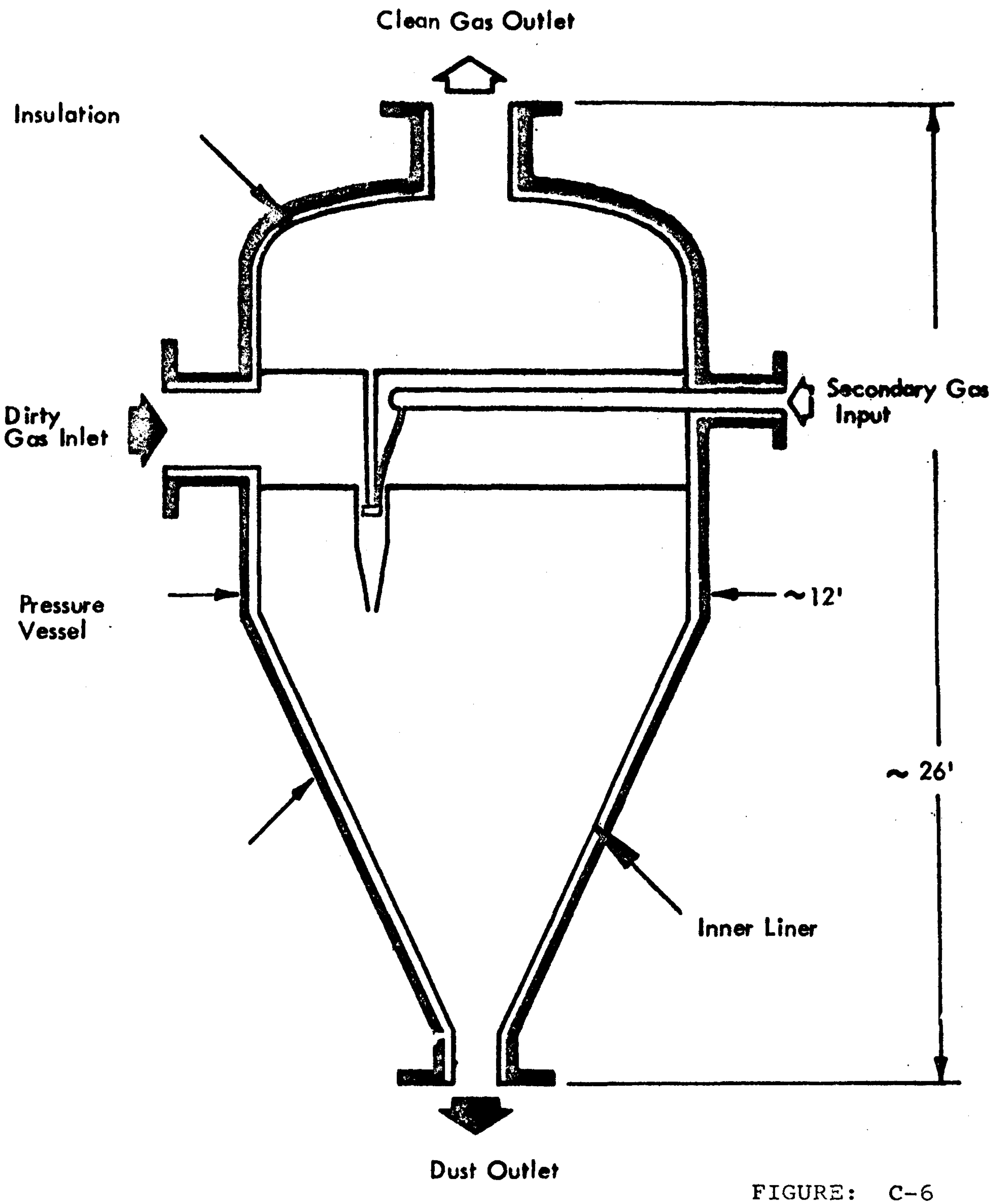


FIGURE: C-6

is lower than that of the Aerodyne Model 22000SV Dust Collector. Furthermore, the price of the Donaldson "Tan-Jet" system per PFB is significantly higher than that of the Aerodyne system. Consequently, Aerodyne's Model 22000SV Dust Collector has been chosen over Donaldson's "Tan-Jet" System because of its predicted performance and economic advantages.

Granular bed filters have been considered as an alternate method of removing particulates from the PFB gas stream. A fixed granular bed system design has been developed by the Ducon Company under subcontract to Babcock & Wilcox (B&W purchase order 717363DU). The study report by Ducon is contained in Appendix 8.1.

A conventional cyclone dust collector is located upstream of the granular bed filter in order to reduce the dust loading to the filter. This reduces the granular bed cleaning requirements and the overall system pressure drop. The predicted collection performance of the Ducon system is shown in Table C-6. Ducon predicts that no particles greater than 2 microns will enter the gas turbine. Comparison of Tables C-4 and C-6 indicates that the collection performance of the Ducon system is greatly superior to that of the Aerodyne system for particle diameters of less than 10 microns. However, the price of the Ducon system per PFB is significantly higher than that of the Aerodyne system. The installed costs of Ducon Granular Bed Filters and Aerodyne's Dust Collector Systems with all accessories for the whole plant are estimated to be \$14,930,000 and \$7,310,000, respectively. Furthermore, the anticipated operating availability of the Ducon system is less than that of the Aerodyne system. For example, the Aerodyne system has no moving parts and is continuously self cleaning whereas the Ducon system makes use of auxiliary equipment such as compressors and valves to periodically backwash (clean) individual filter elements. Consequently, Aerodyne's Model 22000SV Dust Collector has been selected for use in the conceptual plant design.

TABLE C-6

Projected Removal of Particulates in the Granular Bed Filter System for a Calcium/Sulfur Ratio of 1.0

*Dust flow entering the cyclone precleaner	6474 lbm/hr
Dust flow leaving the cyclone precleaner	1025.6 lbm/hr
Dust flow leaving the granular bed filter	25.64 lbm/hr
Dust concentration entering the gas turbine	0.0095 grains/SCF

Particle distribution entering the turbine

Particle diameter, d (microns)	Concentration (grains/SCF)
$d < 2.0$	0.0095
$2.0 < d < 10.0$	0.0000
$d > 10.0$	0.0000

*These flow rates are for each PFB combustor.

5.0 REGENERATION OF SULFATED SORBENTS FROM FLUIDIZED BEDS

5.1 INTRODUCTION

An engineering study of the regeneration of sulfated additives from a coal-fired fluidized bed power plant was performed.

The work has involved a review of the current literature, the selection of viable processes to be used for the regeneration, the preparation of conceptual flow diagrams, the identification of required equipment and supporting services for each case studied, and the development of order-of-magnitude capital and operating cost estimates for the complete sulfated additive processing and handling system. The system has been sized to service a 600 MWe combined cycle power plant utilizing a PFB/gas turbine topping cycle in conjunction with an AFB steam plant. Since this study has been performed in parallel with other subtasks, the performance parameters used herein are preliminary estimates and may differ from those discussed in reports on the other subtasks.

Several variations of the one-step regeneration process have been studied for the recovery of elemental sulfur. One has involved the purchase of hydrogen sulfide; a second has involved the manufacture of hydrogen sulfide; and a third variation has involved the use of anthracite coal for sulfur recovery. All three variations have been compared to a once-through additive system.

Wherever possible, commercially proven processes and equipment have been used. All supporting services and manning requirements for each variation have been considered.

Since cost estimates made in this study were only of an "order-of-magnitude", various sensitivity analyses were performed in order to test the validity of the various cost estimates.

5.2 CONCLUSIONS

The results of this engineering study on the regeneration of sulfated additive from a fluidized bed coal-fired power plant indicate the following:

1. From a technical viewpoint, sorbent regeneration appears feasible. However, more experimental data is required if a commercial plant is to be designed with significant confidence levels.

2. Sorbent regeneration utilizing sulfur recovery processes with commercial operating experience, such as the Claus system, cannot be economically justified unless sorbent costs approach $30 per ton.

3. Additional development efforts are required in order to achieve an economical sorbent regeneration system. These efforts must be focused on the development of an economical sulfur recovery system, such as RESOX, as well as on the regenerator itself. Development of one without the other will be of no use economically.

4. If the currently projected costs of a RESOX system prove realistic, sorbent regeneration utilizing this system for sulfur recovery may be more economical than a once-through sorbent system based on a sorbent cost of over $7 per ton. However, to our knowledge, there is no RESOX system in commercial operation today.

5. From an environmental viewpoint, the amount of solid wastes leaving a plant with regeneration is only 35-40% of the amount produced in a "once-through" sorbent system. Therefore, the environmental impact of the waste disposal is greatly reduced.

6. It is recommended that further development work be performed on the regeneration of spent additive using the one-step regeneration process at 2000^{o}F (1093^{o}C) and atmospheric pressure. In conjunction with this work, development effort should be expended upon an economical sulfur recovery system such as Foster Wheeler's RESOX system or an equivalent.

5.3 SCOPE OF REGENERATION STUDY

The scope of this engineering study is to:

1. Review the present state-of-the-art concerning the regeneration of spent additive from the fluidized bed combustion of coal using dolomite and/or limestone.

2. Select a viable process that could be used commercially.

3. Develop overall conceptual flow diagrams for all cases studied.

4. Identify all equipment and supporting services.

5. Develop order-of-magnitude material and heat balances, including supporting systems for a fully integrated plant complex and develop capital cost estimates.

6. Develop raw material and utility requirements, manpower and service needs to operate the entire additive regeneration and sulfur recovery complex on a continuing basis. Develop order-of-magnitude operating cost estimates.

7. Develop capital and operating cost data for the following cases:

a. Once-through system.

b. Regeneration of spent additive with recovery of elemental sulfur in a Claus Plant using purchased hydrogen sulfide.

c. Regeneration of spent additive with recovery of elemental sulfur in a Claus Plant using manufactured hydrogen sulfide.

d. Regeneration of spent additive with recovery of elemental sulfur, using the RESOX process developed by Foster Wheeler Corporation.

8. Conduct sensitivity analyses on the economics of the above cases.

5.4 PROCESSES FOR THE REGENERATION OF SULFATED SORBENTS

When coal is burned in a fluidized bed containing limestone and/or dolomite, SO_2 from the combustion of sulfur in the coal reacts with the calcium in the bed material and forms $CaSO_4$ which is retained in the bed.

The additive material may be either regenerated to a form suitable for SO_2 removal in the fluid bed system, or disposed of in its partially utilized form in a once-through system.

Two regeneration processes (designated the one-step and the two-step regeneration processes) were selected for study. Both consist primarily of heating the spent additive in the presence of reducing gases at relatively high temperatures to produce gaseous sulfur compounds and either calcium oxide or calcium carbonate.

5.4.1 One-Step Regeneration

The one-step dolomite or limestone regeneration process consists of a single fluidized bed reactor in which spent additive containing $CaSO_4$ from the coal-fired fluid bed system is reacted with a reducing gas, such as H_2 or CO, to produce CaO and SO_2. The endothermic reaction at 2000°F (1093°C) and 1 atmosphere pressure is:

$$CaSO_4 + \begin{matrix} H_2 \\ or \\ CO \end{matrix} \rightleftarrows CaO + SO_2 + \begin{matrix} H_2O \\ or \\ CO_2 \end{matrix} \qquad (1)$$

The rate of reaction between $CaSO_4$ and either H_2 or CO is quite high at 2000°F. It is desirable to produce high concentrations (10 to 15% by weight) of SO_2 in order to enhance the sulfur recovery. (Ref. 4). The SO_2 equilibrium concentration is favored by reduced pressure, being inversely proportional to the total pressure. The reduction of $CaSO_4$ to CaO is favored by high temperatures and mildly reducing conditions (one mole of either H_2 or CO for every mole of $CaSO_4$).

At lower temperatures, 1650°F (899°C) and more highly reducing conditions, the following reaction is favored (Ref. 5):

$$CaSO_4 + 4\ \begin{matrix} CO \\ \text{or} \\ H_2 \end{matrix} \longrightarrow CaS + 4\ \begin{matrix} CO_2 \\ \text{or} \\ H_2O \end{matrix} \qquad (2)$$

The formation of large amounts of CaS is undesirable since it prevents the reductive decomposition of $CaSO_4$ to CaO. Consequently, careful control of process conditions is required. If some CaS is formed along the way, it would eventually be eliminated (to some extent) by the following reaction at 2000°F (1093°C):

$$CaS + 3\ CaSO_4 \longrightarrow 4\ CaO + 4\ SO_2 \qquad (3)$$

To limit the formation of CaS, the concentration of reducing gases must be carefully controlled. Also, advantage can be taken from the fact that CO_2 and H_2O and high temperatures suppress the formation of CaS.

5.4.2 Two-Step Process

The two-step dolomite or limestone regeneration process involves, first, the reduction of $CaSO_4$ to CaS, and, second, the reaction of the CaS with CO_2 and H_2O to form $CaCO_3$ and H_2S. The first step at 1650°F and 1 atmosphere pressure is:

$$CaSO_4 + 4\ \begin{matrix} CO \\ \text{or} \\ H_2 \end{matrix} \longrightarrow CaS + 4\ \begin{matrix} CO_2 \\ \text{or} \\ H_2O \end{matrix} \qquad (4)$$

The second step at 1100°F and 10 atmosphere pressure is:

$$CaS + H_2O + CO_2 \longrightarrow CaCO_3 + H_2S \qquad (5)$$

In the first step, the reaction starts out reasonably fast, but then slows down quickly due to the tendency of the CaS to cover the pores of the remaining crystals of $CaSO_4$, thereby decreasing the available contact surface.

The regenerated additive from this two-step process must be recalcined to CaO. Furthermore, four times as much reducing gas must be used for the two-step process compared with the one-step process.

5.4.3 Process Variables

For either of the two processes discussed, the most important process variables from the standpoint of regeneration performance are temperature, partial pressure of the reducing gases, and space velocity or contact time of the $CaSO_4$ particles and gases (Refs. 5,6). An appropriate system for carrying out the additive regeneration process is a fluidized bed reactor which provides the necessary temperature uniformity, as well as efficient contact between the gases and solids involved.

When operating in the range of 2000°F (1093°C), deactivation of the CaO particles by sintering or deadburning occurs, especially when the solids must undergo repeated cycles of sulfur absorption and regeneration. In addition, a degree of solids attrition can be expected in the regeneration step. Additive recirculation rate through the regenerator and fresh additive make-up rates to the combustor are determined by the amounts of deactivation and attrition that occur in the overall process. Therefore, the overall economics of regeneration are greatly dependent upon the additive's resistance to these factors.

The composition of the ash component in the spent additive also limits the temperature of regeneration, since any low melting point materials could cause agglomeration of the particles in the fluidized bed.

With higher regeneration temperatures, there appears to be a definite trend toward higher SO_2 concentrations in the regenerator off-gas and greater $CaSO_4$ conversion to CaO.

As the mole fraction of SO_2 in the regenerator off-gas increases, fuel costs for regeneration decrease (Ref. 5). Hence, this provides an incentive to operate at low pressures, because the mole fraction of SO_2 at equilibrium rises as the total pressure of the system is reduced.

Varying the superficial gas velocity and settled bed height can affect the concentration of SO_2 in the regenerator off-gas by changing the contact time between gaseous and solid phases (Ref. 5). The less the contact time due to increasing fluidization gas velocity, the lower the percent of sulfur regeneration.

Since perfect mixing cannot be obtained in a commercial fluidized bed, a variable temperature distribution and reducing gas concentration can be expected throughout the bed, with significant proportions of CaS being formed instead of CaO. This problem can be sharply reduced by adding auxiliary air directly to the fluidized bed (Ref. 5). This creates adjacent reducing and oxidizing zones in the bed, thereby decreasing the tendency to form CaS.

5.4.4 Regeneration Process Selection

Table D-1 lists process conditions and end products for the two regeneration processes. In the one-step process, a high temperature reduction of calcium sulfate to calcium oxide yields sulfur dioxide for recovery. In the two-step process, a lower temperature reduction of calcium sulfate to calcium carbonate yields hydrogen sulfide for recovery.

The following tabulation lists some of the advantages and disadvantages of the two regeneration processes (Ref. 4):

TABLE D-1. REGENERATION PROCESS CONDITIONS

One-Step Regeneration

Conditions:	2000°F, one Atmospheric pressure, One mole of reducing gases required
End Products:	CaO for recycle to combustor. SO_2 for sulfur recovery

Two-Step Regeneration

First Step - Conditions:	1600°F, one Atmosphere pressure, Four moles of reducing gases required
End Products:	CaS for use in second step
Second Step - Conditions:	1100°F, ten Atmospheres pressure, CO_2 and H_2O gases required
End Products:	$CaCO_3$ for recycle to combustor H_2S for sulfur recovery

a. For the one-step regeneration process:

<u>Advantages</u>
- It is a single stage process
- Experimental data is available
- CaO is formed directly.

<u>Disadvantages</u>
- Reduction may produce undesirable CaS in a competing reaction
- High temperatures, 2000°F (1093°C), are required to avoid CaS formation and deactivation of the additive. Also, close temperature control is needed to avoid agglomeration of the coal ash in the bed.
- At equilibrium, the SO_2 concentration decreases with pressure.

b. For the two-step regeneration process:

<u>Advantages</u>
- Thermodynamics are favored by low temperature 1600°F (871°C)
- There is no thermodynamic disadvantage due to pressure
- Low temperature avoids solid sintering problems.
- Pressure favors H_2S production

<u>Disadvantages</u>
- Two stages are required to form $CaCO_3$
- Second step requires high pressure CO_2 and H_2O
- Little experimental data is available or publicized
- Competing reactions reduce sulfate to yield SO_2
- Carbonate is produced (rather than oxide) that must be recalcined for recycle
- Reaction rate of CaS conversion slows down drastically.

While it is recognized that no firm conclusion can be drawn from an evaluation of the above, the one-step process has been selected for economic evaluation in this study.

5.5 INTEGRATED REGENERATION AND SULFUR RECOVERY PROCESSES

Having selected the process to be used for regenerating the spent additive, the support processes required to achieve a complete integrated system were then selected. The philosophy adopted to guide the system design has been to utilize commercially proven processes where available in order to limit the time and cost required to commercialize the plant.

The one-step regeneration process is envisioned as a continuous process, in which $CaSO_4$ in the spent additive is reduced to CaO and SO_2 in a fluidized bed by reacting with a reducing gas at 2000°F (1093°C) and a pressure slightly above atmospheric (Ref. 7). The heat for the endothermic decomposition reaction is supplied by burning pulverized coal in the fluidized bed reactor, with the necessary reducing gases being supplied by gasifying additional coal either in the regenerator itself or in a separate outside source.

The fuel required to supply the necessary heat is added to the regenerator just above the fluidizing grid to produce a reducing zone, while supplementary excess air is added higher up in the bed to produce an oxidizing zone. The purpose of the oxidizing and reducing zones is to minimize the amount of CaS formed in the regeneration process (Ref. 5).

Wherever possible, commercially proven processes have been selected. This criterion led to the initial selection of a Claus sulfur recovering plant which requires a supply of H_2S for conversion of SO_2 to elemental sulfur.

For reasons which will be clarified later, a new sulfur recovery process, known as the RESOX(TM) process, currently under development by Foster Wheeler Corporation, has also been considered. This process does not require H_2S but does require a supply of anthracite coal.

The processes which have been integrated to form the three systems investigated in this study are as follows:

1.	Additive Regeneration	-	(Cases I, II and III)*
2.	Reducing Gas Generation	-	(Cases I and II)
3.	Claus Sulfur Recovery	-	(Cases I and II)
4.	Claus Tail-gas Cleanup	-	(Cases I and II)
5.	Hydrogen Generation	-	(Case II)
6.	Hydrogen Sulfide Generation	-	(Case II)
7.	RESOX Sulfur Recovery	-	(Case III)

---

* Case I involves the purchase of H_2S for a Claus Sulfur Recovery Unit. Case II involves in-plant manufacture of H_2S for a Claus Sulfur Recovery Unit. Case III involves sulfur recovery using the RESOX process.

Preliminary calculations indicated that it would not be economical to purchase H_2S for the Claus plant. In order to manufacture H_2S in-plant, the amount of reducing gas required would be four times that required for additive regeneration. This factor led to the decision to use a separate reducing gas plant to provide the raw gas needs for both a H_2 generation plant and for the additive regenerator. It was further decided that, in the interest of completeness, an estimate would still be prepared for the case involving the purchase of H_2S. However, to facilitate the design and cost estimating efforts, the production of reducing gas for the regeneration process has still been accomplished in a separate process, rather than directly in the regenerator vessel itself. While probably not the most economical approach for this case, the incremental costs involved would not significantly affect the conclusion regarding overall economics between Case I (Purchased H_2S) and Case II (In-plant H_2S manufacture).

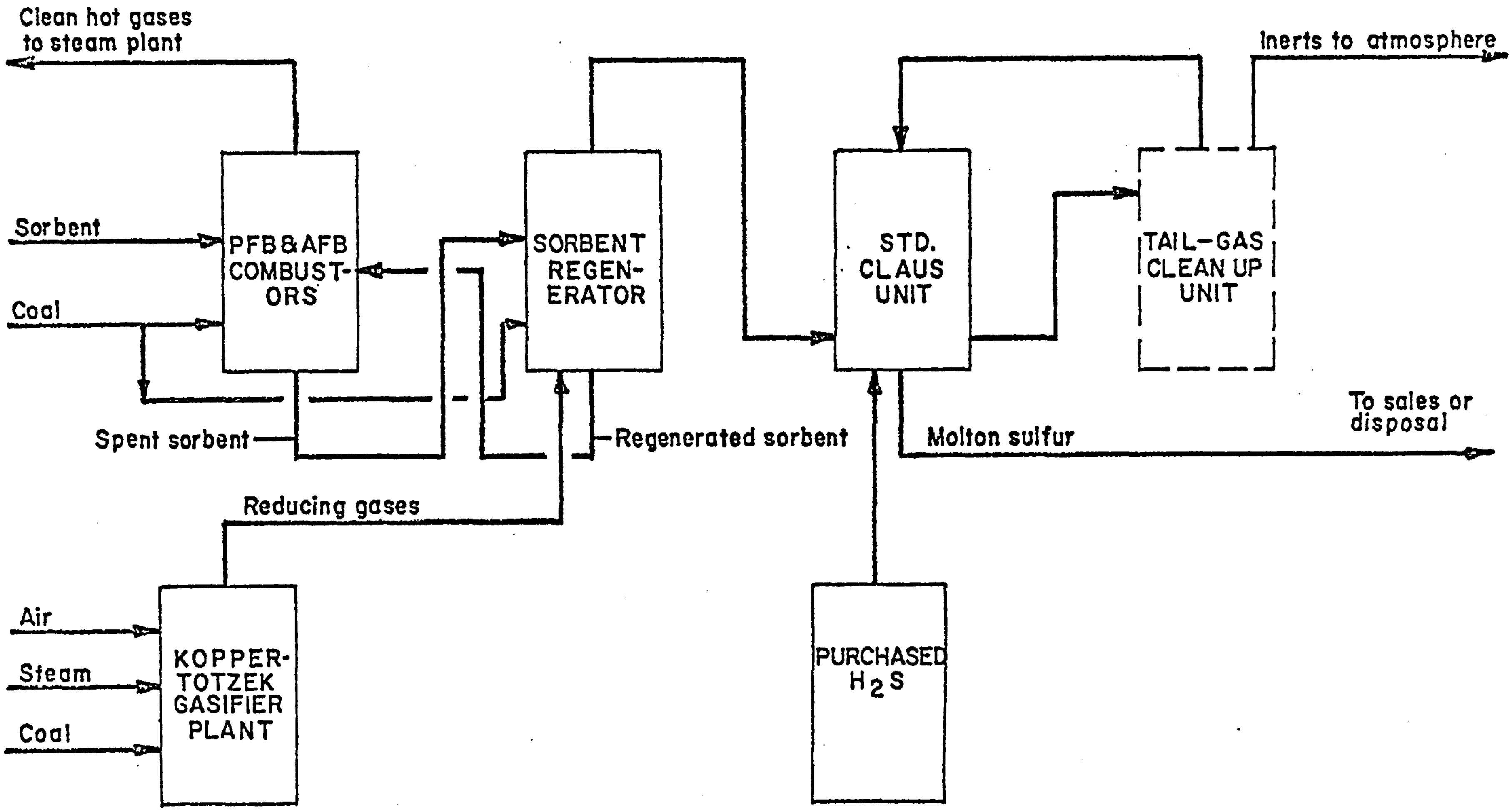
5.5.1 Process Flow Diagrams

Figures D-1 and D-2 show block diagrams of processes selected for Cases I and II, respectively. Reducing gases are separately generated in a Koppers-Totzek Coal Gasifier Package Unit. A standard Claus sulfur recovery unit and tail-gas treatment plant are also shown. It should be noted that the Claus tail-gas clean-up plant is only required if the recycle of tail-gas to the fluid bed combustors proved technically or economically impractical. While this is considered unlikely, the clean-up system is included here as a conservative measure. Again, the cost of this clean-up plant does not significantly affect the final conclusions. Other processes used only in Case II include a conventional water-gas shift reaction for the production of H_2 and the catalytic reaction of H_2 and sulfur vapors to produce H_2S gas.

On the basis of the block diagrams shown in Figures D-1 and D-2, a conceptual flow diagram and approximate heat and material balances were prepared for each case. Figure D-3 shows the overall flow diagram for both cases, with Case II the more complex of the two cases. Approximate sizes of all equipment and piping indicated on the flow diagrams were then established for both cases.

When it became apparent that the costs associated with Claus sulfur recovery system represented a major portion of the total annual operating costs for Cases I and II, a decision was made to investigate Foster Wheeler's RESOX process for this application. To our knowledge, there is no RESOX process in commercial operation as yet, and there is very little technical information available for study and evaluation. However, it appears that overall system costs with a RESOX unit could be lower than for an equivalent Claus-based system. Therefore, it was decided that for comparison purposes, a third case incorporating a RESOX system should be studied on the same basis as Cases I and II. The block diagram on Figure D-4 and conceptional flow diagram Figure D-5 were prepared based on our interpretation of the sparse information available on the RESOX system.

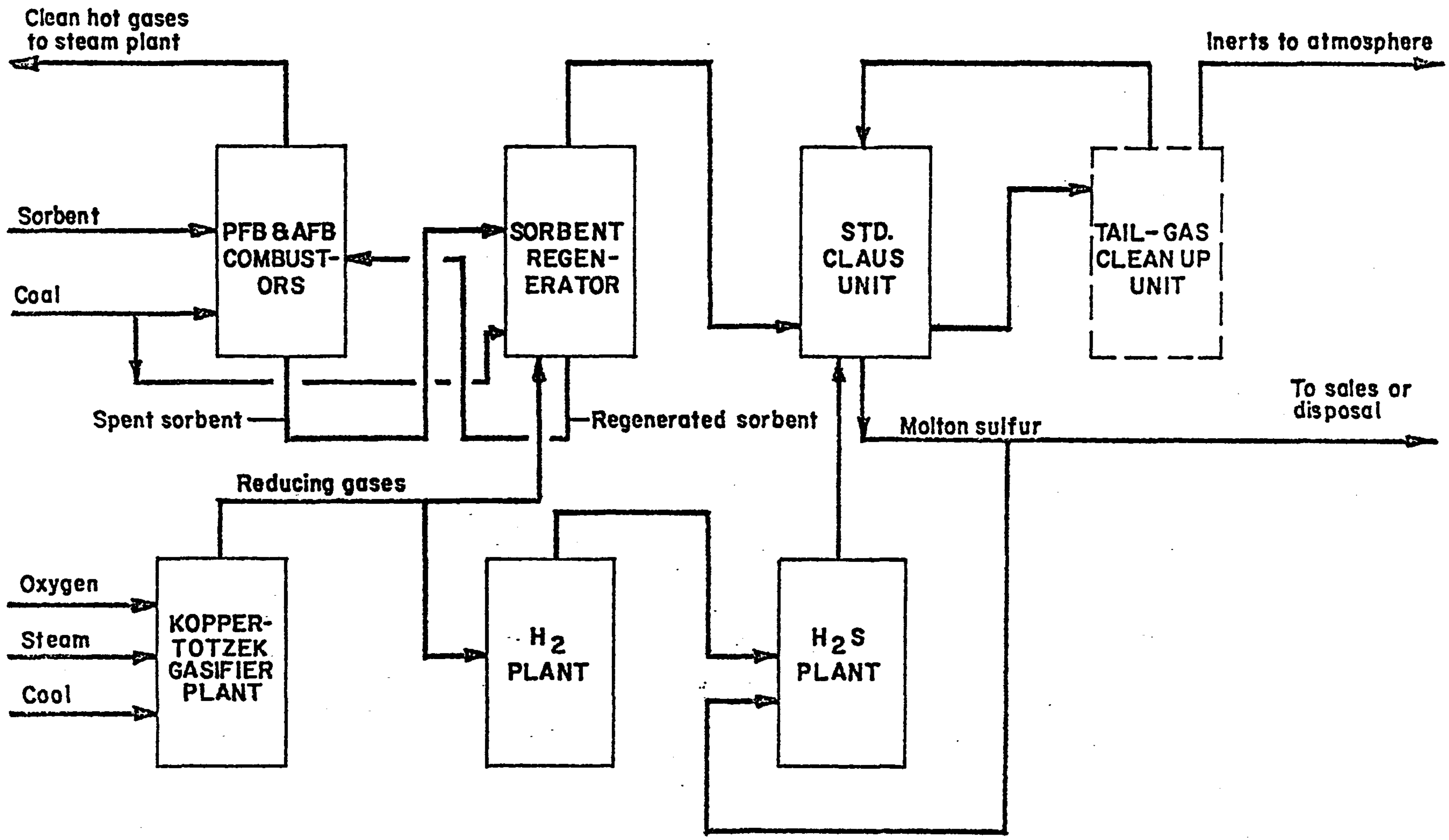
Case III differs from Cases I and II in that the RESOX tail-gas is recycled back to the AFBC boiler, so that no tail-gas clean-up system is used. Case III is therefore less conservative than Case II.



SORBENT REGENERATION WITH CLAUS PLANT AND PURCHASED H_2S

CASE I

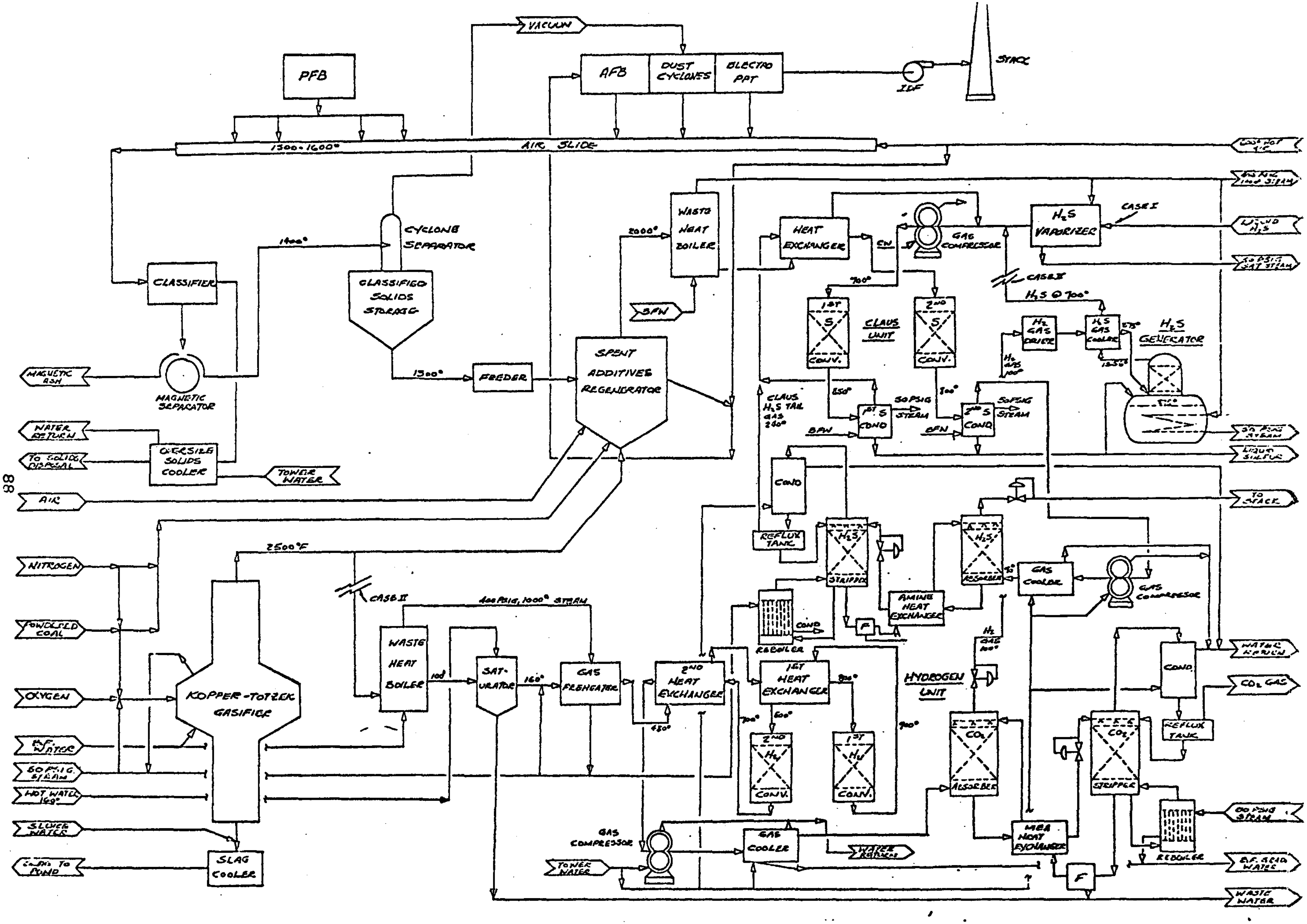
FIGURE: D-1



SORBENT REGENERATION WITH CLAUS PLANT AND MANUFACTURED H_2S

CASE II

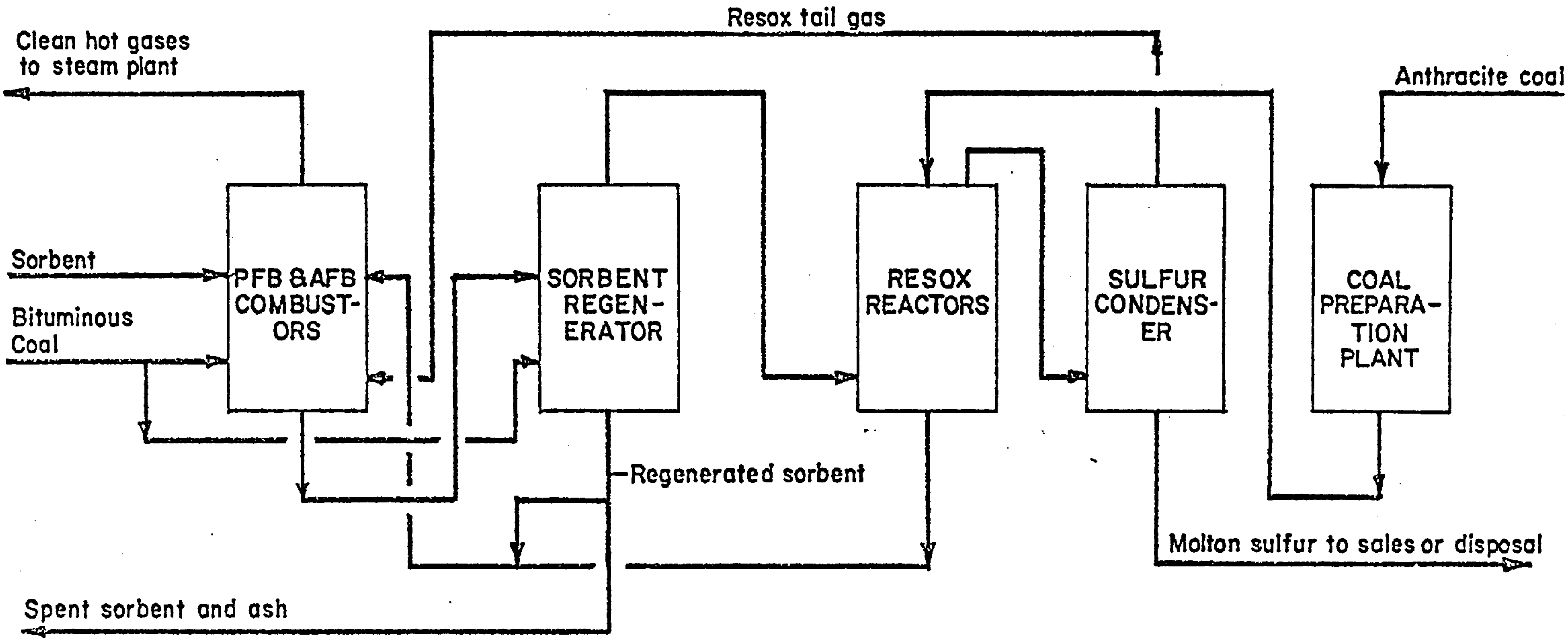
FIGURE: D-2



SORBENT REGENERATION WITH CLAUS PLANT

CASES I & II

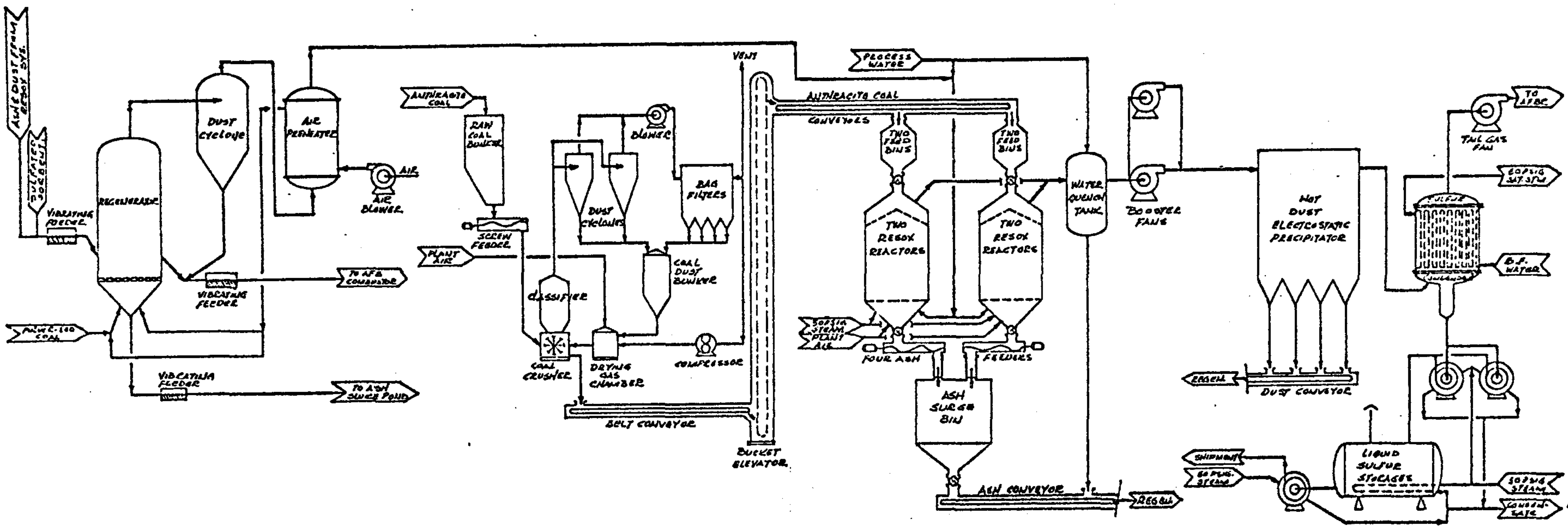
FIGURE: D-3



SORBENT REGENERATION WITH RESOX SULFUR RECOVERY PLANT

CASE III

FIGURE: D-4



SORBENT REGENERATION WITH
RESOX SULFUR RECOVERY
CASE III

FIGURE: D-5

In addition, all required reducing gases are produced by gasifying coal within the additive regenerator itself, rather than in a Koppers-Totzek gasifier.

5.5.2 Detailed Description of Integrated Regeneration Processes for Cases I, II, and III

Appendix 8.2 contains a complete description of the integrated regeneration processes and equipment for Cases I, II, and III. Also included is a discussion of the design criteria and detailed process flow sheets which have been developed for each case. This information has been used as the basis for the cost estimates which are discussed later in this section.

5.5.3 Description of Once-Through Process for the Base Case

Solid wastes from the PFBC combined cycle power plant are removed from the various terminal points located in different areas of the plant. These include sulfated dolomite from the PFBC, sulfated limestone from the AFB, fly ash from the AFB and PFB cyclones, and fly ash from the electrostatic precipitators.

The dry solid waste handling system includes waste cooling, pneumatic transporting, storage and unloading.

Solid wastes from the various terminal points flow by gravity to the coolers, where the waste is cooled from about 1600° F to 200° F. Air enters the bottom of the cooler to fluidize the waste to ensure agitation of the material and provide good contact with cooling coil surfaces, thus providing maximum heat transfer. Steam cycle condensate enters the cooling coils at 100° F and leaves at 200°F to be sent to deaerators. Each cooler is provided with a separate air blower. The heated fluidized air leaves the coolers at 600° F passes through multi-cyclone dust collectors and then to the electrostatic precipitators before venting to plant stack.

The solids from the coolers are dropped into transfer hoppers and then pneumatically conveyed to storage silos. The positive pressure pneumatic transporting system consists of nine independent conveying sections, designed to operate on a timed cycle operation. Each section is equipped with transfer hoppers, single stage rotary compressors and control panels.

The solid wastes are collected and stored in four concrete silos, each equipped with separate dust collection system, and a rotary unloader for transferring the solids from silos into closed railroad hopper cars or trucks. The unloaders can be remotely controlled from ground level. Provision has been provided to condition the dust wastes with sufficient water to prevent dusting during the loading, transporting, and dumping at the off-site disposal area.

Since the Commercial Plant Design (i.e. Subtask 1.2) task had not yet reached the point where cost data for this system was available, an order-of-magnitude capital cost was assumed for use in this trade-off study. Subsequently, when the estimate for Subtask 1.2 was completed, it was found that the solid waste disposal system cost was 25% higher than used here. This difference would not affect the conclusions drawn from the results presented here. Therefore, this analysis has not been updated.

5.6 SUMMARY OF MATERIAL FLOWS

Table D-2 contains a summary of the overall material flows for the entire plant. A summary of selected material flows for each individual process of the four cases considered (Cases I, II, III, and Base Case) are shown on the tables indicated below:

a. Additive Regeneration Process Flows for Cases I, II, and III - Table D-3

b. Reducing Gas Generation Process Flows for Cases I and II - Table D-4

c. Claus Sulfur Recovery Process Flows for Cases I and II - Table D-5

d. Claus Tail-gas Cleanup Process Flows for Cases I and II - Table D-6

e. Hydrogen Generation Process Flows for Case II - Table D-7

f. Hydrogen Sulfide Generation Process Flows for Case II - Table D-8

g. RESOX Sulfur Recovery Process Flows for Case III - Table D-9

h. Once-Through (No Regeneration) Sorbent System Flows for Base Case - Table D-10

5.7 ECONOMIC EVALUATION OF SORBENT REGENERATION SYSTEMS

5.7.1 Capital Cost Estimates

Using the order-of-magnitude material and heat balances described in previous sections, approximate sizes have been determined for each piece of equipment as a starting point for the development of individual installed capital costs.

The basis for the installed capital cost estimates for each piece of equipment is the "module" concept described in an article by K. M. Guthrie (Ref. 8). The FOB equipment costs, as given in the article, after considering operating conditions and materials of construction, have been scaled to 1977 costs by utilizing appropriate chemical plant equipment cost indices.

TABLE D-2. SUMMARY OF OVERALL MATERIAL FLOWS FOR 600 MWe COMBINED POWER PLANT

Basis: Tons/Day

		Regeneration Systems		
	Once Through	Case I	Case II	Case III
Entering:				
Coal: For Power	5,000	5,000	5,000	5,390
For Regeneration	-	245	509 (3)	255
For Sulfur	-	-	-	121
Oxygen	-	-	335	-
Liquid H_2S	-	458	-	-
Misc. Chemicals:				
Nitrogen	-	10	20	-
Amines	-	0.1	0.1	-
Additive Makeup	2,425	518	518	1,025 (5)
Additive Recirculated:	0	5,182	5,182	4,378
Leaving:				
Spent Additive & Ash	3,340	1,147	1,147	1,390
Recovered Sulfur	0	563 (1)	187 (2)	164 (4)

Notes:
(1) High sulfur recovery due to purchase of H_2S
(2) Includes sulfur in coal to regenerator and gasifier
(3) Total coal used for regenerator and gasifier for hydrogen generation
(4) Stoichiometric recovery at 90%
(5) As optimized by Argonne National Laboratory. However, according to Argonne data, the total costs are quite insensitive to additive feed rate over the range of these studies.

TABLE D-3

ADDITIVE REGENERATION PROCESS FLOWS

Basis: Tons/Day

	Case I	Case II	Case III
Entering:			
Coal to Combustors	5000	5000	5390
Coal for Regeneration	171	171	255
Air to Regenerator	1912	1912	1439
Fresh Additive	518	518	1025
Sulfated Additive Recirculation	5182	5182	4378
Leaving:			
Spent additive and Ash	1147	1147	1390
Off-gas	3449	2496	2159

TABLE D-4

REDUCING GAS GENERATION PROCESS FLOWS

Basis: Tons/Day

	Case I	Case II
Entering:		
Coal for Gasifier	74	338
Oxygen	-	335
Nitrogen	2	10
Air	354	-
Leaving:		
Reducing Gas	413	553

TABLE D-5

CLAUS SULFUR RECOVERY PROCESS FLOWS

Basis: Tons/Day

	Case I	Case II
Entering:		
Regenerator Off-gas	3449	2496
Hydrogen Sulfide	458	-
Recovered gases from Claus Tail Gas Cleanup	577	577
Leaving:		
Recovered Sulfur	563	187
Tail-gas	3158	2392

TABLE D-6

CLAUS TAIL-GAS CLEANUP PROCESS FLOWS

Basis: Tons/Day

	Case I	Case II
Entering:		
Tail-gas	3158	2392
Leaving:		
Tail-gas to Plant Stack	2006	1520
Recovered Gases to Claus Unit	577	577

TABLE D-7

HYDROGEN GENERATION PROCESS FLOWS

Basis: Tons/Day

	Case II
Entering:	
Kopper-Totzek Gas	553
Leaving:	
Hydrogen Gas to H_2S Generator	28
Stripper Vent Gases to Plant Stack	724

TABLE D-8

HYDROGEN SULFIDE GENERATION PROCESS FLOWS

Basis: Tons/Day

	Case II
Entering:	
Hydrogen Gas	28
Liquid Sulfur	433
Leaving:	
Hydrogen Sulfide	460

TABLE D-9

RESOX SULFUR RECOVERY PROCESS FLOWS

Basis: Tons/Day

Entering:	
Regenerator Gas	2159
Anthracite Coal	121
50 psig Steam	313
Process and Quench Water	710
Air	358
Leaving:	
Recovered Sulfur	164
Tail Gases to AFBC	3593
Coal Ash and Particulates to Regenerator	70

TABLE D-10

ONCE-THROUGH SORBENT SYSTEM FLOWS

Basis: Tons/Day

Entering:	
Bituminous Coal	5000
Additive	2425
Sluice Water	4546
Leaving:	
Spent Additive and Ash	3340
Sluice Water Content	4547

Guthrie's factors for materials (piping, concrete, steel, electrical, instrumentation, linings, insulation and paint) field labor, indirect costs (freight, insurance, taxes, construction overhead, engineering, contingencies, and contractor's fee), supporting services, and off-site facilities, have then been used with the 1977 FOB equipment costs. In addition, approximate cost estimates received from Ford, Bacon, and Davis (Claus Plants, Tail-gas Treatment Plant), Koppers (K.T. Gasifier), and Foster Wheeler Energy Corporation (RESOX), while not directly applicable in all cases, have been used as a guide in dev eloping costs for the various processes.

A breakdown of capital costs for each case by section is shown on Table D-11.

5.7.2 Operating Cost Estimates

Economic evaluations of the three regeneration cases relative to the once-through additive system has been based on operating cost estimates developed for each case.

For each case, the annual costs for raw material usages, utility requirements, waste and by-product generation, operating and maintenance labor, supervision, capital charges, and administrative and overhead expenses have been determined.

Table D-12 lists the basis for operating cost estimates.

Table D-13 lists the Regeneration Plant Manning Requirements.

Table D-14 lists utilities requirements.

Table D-15 lists operating cost comparisons for the various systems.

5.7.3 Economic Comparison of Regeneration Systems

Due to the high cost of the H_2S, Case II has been found to be more economical than Case I, even though the capital investment is twice as much as Case I (See Table D-15). However, both Cases I and II are uneconomical compared to once-through operation of the sorbent system.

As indicated on Table D-15, the regeneration system cost with RESOX sulfur recovery (Case III) is much lower than with a Claus sulfur recovery system. In addition, the operating cost for Case III is competitive with once-through operation, especially if a credit for sulfur sales is taken.

5.7.4 Cost Sensitivity Analysis

Since the cost estimates made in this study were only of an "order of magnitude", various sensitivity analyses were performed in order to test the validity of the comparisons made on Table D-15.

TABLE D-11

CAPITAL COST BREAKDOWN

SECTION	Once Through $	Case I $	Case II $	Case III $
No Regeneration	5,500,000	-	-	-
Regeneration	-	5,466,400	4,132,500	2,252,700
Reducing Gas Generation	-	6,507,500	14,646,800	-
Claus Sulfur Recovery	-	6,445,400	7,009,500	-
Claus Tail-Gas Cleanup	-	7,600,700	7,532,600	-
Hydrogen Gas Generation	-	-	15,065,300	-
Hydrogen Sulfide Generation	-	-	3,923,300	-
RESOX Sulfur Recovery	-	-	-	15,947,300
TOTAL	$5,500,000	$26,030,000	$52,310,000	$18,200,000

TABLE D-12

BASIS FOR OPERATING COST ESTIMATES

Materials

Coal

12,450 BTU/lb. HHV
4.5% S
$19.50 per ton delivered (Bituminous Coal)
$25.30 per ton delivered (Anthracite Coal)
Limestone & Dolomite $7/ton delivered

H_2S	$240/ton
O_2	$40/ton
Amines	$0.77/lb.
Electricity	$0.0225/KWH
Water	$.15/1000 gal avg. for all types
Waste Disposal	$3/ton (spent stone & sulfur)
Sulfur	$50/ton FOB plant (sales)

Labor

Operating labor at $20,000/man yr. incl. fringes
Operating superv. at $25,000/man yr. incl. fringes
Chemist, engineer, etc.

Maintenance

Including labor, supervision, supplies, materials and parts at 5% of capital cost.

Capital Charges

At 18% of capital cost

Admin. & Overhead

At 40% of labor and maintenance

Cases Studied

Base Case - Once-through Additive System

Case I - One-Step Additive Regeneration System Buying H_2S "Over the Fence"

Case II - One-Step Additive Regeneration System Making H_2S In-Plant

Case III - One-Step Additive Regeneration with RESOX Sulfur Recovery System

TABLE D-13

REGENERATION PLANT OVERALL MANNING REQUIREMENTS

Basis: 7 Days/Wk

	Once Through	Regeneration Systems Case I	Case II	Case III	Remarks
Overall Supv.	-	-	-	-	From Power Complex
Operation					
Operation Supv.	½	1	2	1	
Shift Supv.	-	5	6	5	
Operators	3	6	8	5	
Helpers	3	6	8	5	
Chemists	-	2	2	2	
Clerks	-	1	2	1	
Total	6½	21	28	19	
Maintenance					
Maintenance Supv.	½	1	1	1	
Electricians		1	2	1	
Helpers		1	2	1	
Millwrights	1	1	2	1	
Helpers	1	1	2	1	
Pipe Fitters	1	1	2	1	
Helpers	1	1	2	1	
Machinists		1	2	1	
Instrument Tech.		1	2	1	
Engineers		1	2	1	
Laborers		2	3	2	
Total	4½	12	22	12	
Overall Total	11	33	50	31	

Note: It is our considered opinion that after proper training all personnel listed above are interchangeable with power complex personnel.

TABLE D-14

SUMMARY OF OVERALL UTILITIES REQUIREMENTS

	Electricity (kwh/D)	Water (Gal./Min.)
Once Through	5,839	758
Case I	282,590	3,900
Case II	365,615	6,635
Case III	35,130	8,400

TABLE D-15

OPERATING COST COMPARISON FOR 600 MWe COMBINED POWER PLANT

	Once Through	Regeneration Systems Case I	Case II	Case III
Capital Cost	$5,500,000	$26,030,000	$52,310,000	$18,200,000
Operating Cost: $/yr.				
Direct Costs:				
(a) Raw Materials -				
Additive	$4,461,000	$ 953,000	$ 953,000	$ 1,886,000
Bituminous Coal		1,258,000	2,606,000	1,307,000
Anthracite Coal				805,000
Oxygen			3,522,000	
Liquid H_2S		28,912,000		
Misc. Chemicals		53,600	160,000	
Subtotal:	$4,461,000	$31,177,000	$ 7,241,000	$ 3,998,000
(b) Utilities -				
Electricity	$ 34,500	$ 1,671,000	$ 1,911,000	$ 208,000
Water	43,000	221,000	376,000	477,000
Subtotal:	$ 78,000	$ 1,892,000	$ 2,288,000	$ 685,000
(c) Stone & Ash Disposal	$2,633,000	$ 904,000	$ 904,000	$ 1,096,000
(d) Maintenance, etc.	$ 275,000	$ 1,302,000	$ 2,616,000	$ 910,000
(e) Operating Labor	$ 178,000	$ 411,000	$ 579,000	$ 425,000
Subtotal:	$3,086,000	$ 2,617,000	$ 4,099,000	$ 2,431,000
TOTAL DIRECT COSTS:	$7,625,000	$35,686,000	$13,628,000	$ 7,114,000
Indirect Costs:				
(a) Capital Charges	$ 900,000	$ 4,685,000	$ 9,416,000	$ 3,276,000
(b) Admin. & Overhead	181,000	685,000	1,278,000	534,000
TOTAL INDIRECT COSTS:	$1,171,000	$ 5,370,000	$10,694,000	$ 3,810,000
TOTAL ALL COSTS: Without Sulfur Disposal	$8,796,000	$41,056,000	$24,322,000	$10,924,000
Credit for Sulfur Sales	$ 0	$ 7,398,000	$ 2,457,000	$ 2,158,000
Net Cost With Sulfur Sales	$8,796,000	$33,658,000	$21,865,000	$ 8,766,000
Net Cost Without Sulfur Sales	$8,796,000	$41,500,000	$24,469,000	$11,054,000
SO_2 Removal Cost Per KWH:				
With Sulfur Sales		$ 8.9 mil.	$ 5.8 mil.	$ 2.3 mil.
Without Sulfur Sales	$ 2.3 mil.	$ 11.0 mil.	$ 6.5 mil.	$ 2.9 mil.

Figure D-6 shows that the systems utilizing Claus plants (Cases I and II) are not economical even if the capital costs are 40% less than the estimated value. However, even a slight reduction in the capital cost of Case III (taking credit for sulfur sales) relative to that estimated makes it more economical than the once-through operation. Within the accuracy of the estimate, Case III with sulfur sales has a comparable cost to once-through operation.

In comparing direct operating cost, Figure D-7 indicates that even if the actual value is 40% less than the estimated value on Table D-15, Cases I and II are uneconomical compared with the once-through case. However, within the accuracy of this estimate, the cost of Case III when credited with sulfur sales is comparable to costs for a once-through system.

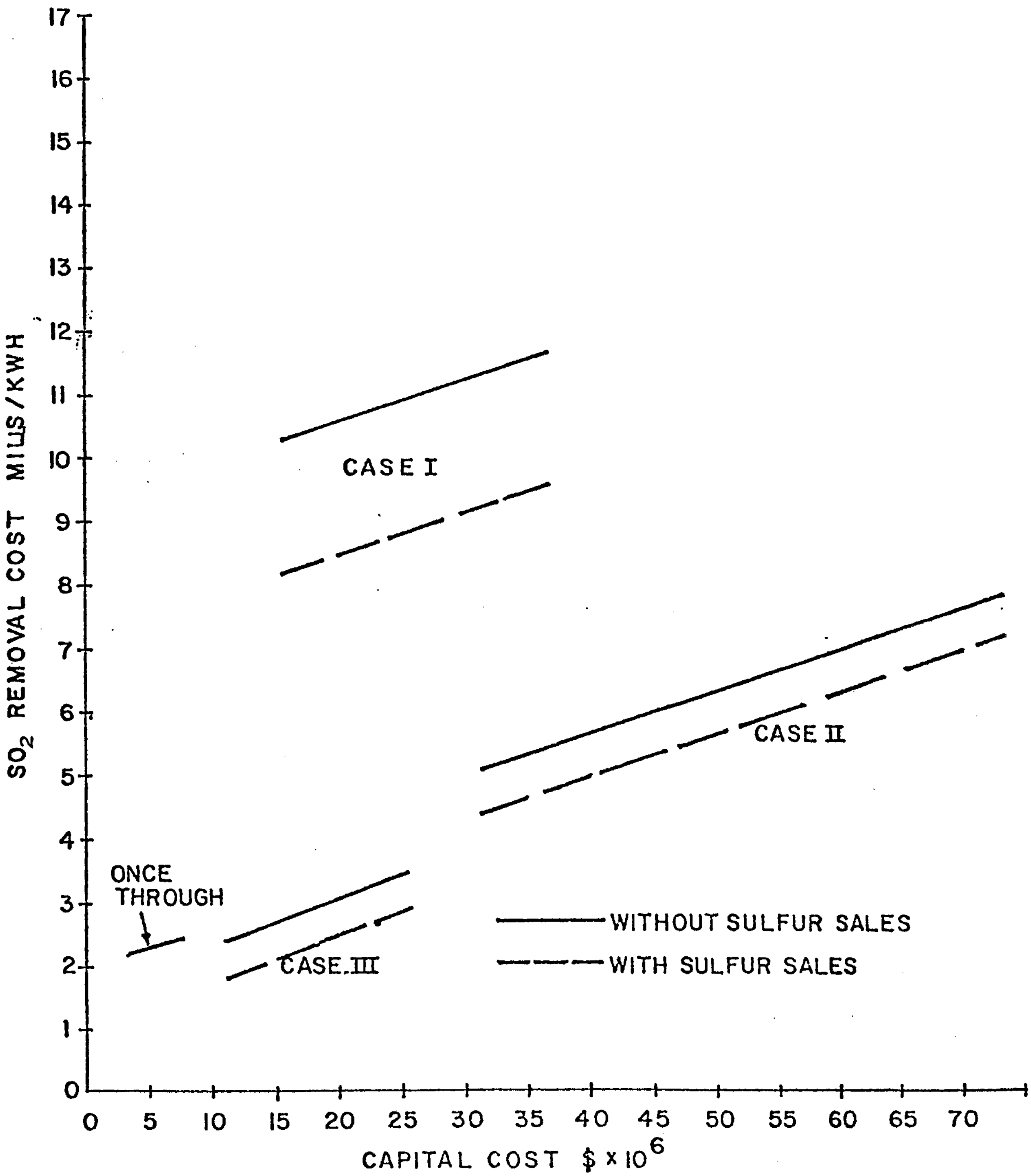
Figure D-8 shows the effect of total operating cost on the annual SO_2 removal costs for the various cases.

Figure D-9 shows the additive cost that would be required in order for the various sorbent regeneration cases to be equivalent economically to the once-through operation case. When taking credit for sulfur sales, these breakeven sorbent costs are as follows: Case I - $58/ton; Case II - $34/ton; and Case III - $7/ton.

5.8 ADDITIONAL REGENERATION PLANT DATA AND COMMENTS

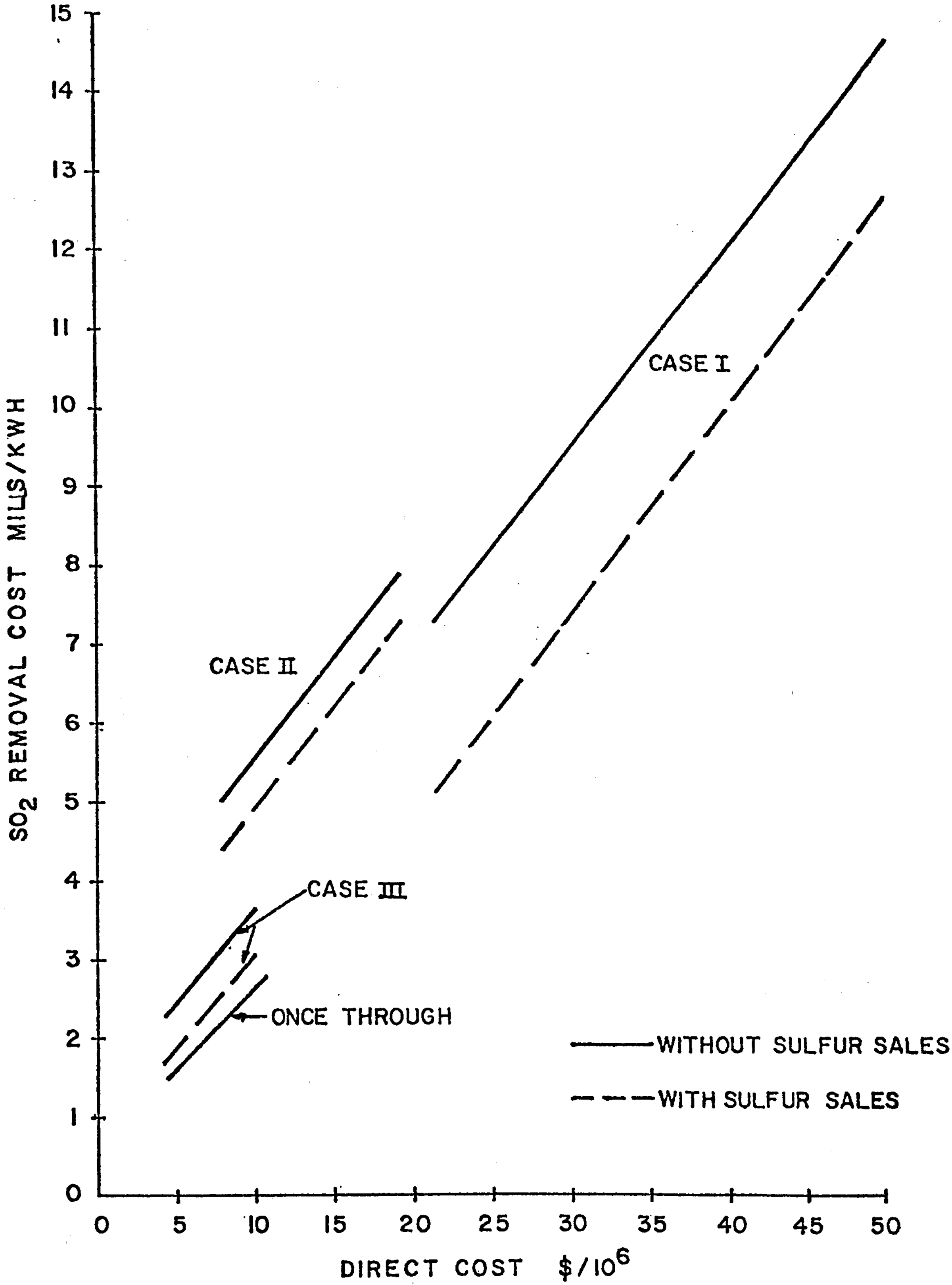
Table D-13 details the regeneration complex manning requirements for the various cases studied. With proper training all personnel listed are considered interchangeable with power complex personnel.

Table D-16 shows a breakdown of the space requirements for the various processes used in Cases I, II, and III.



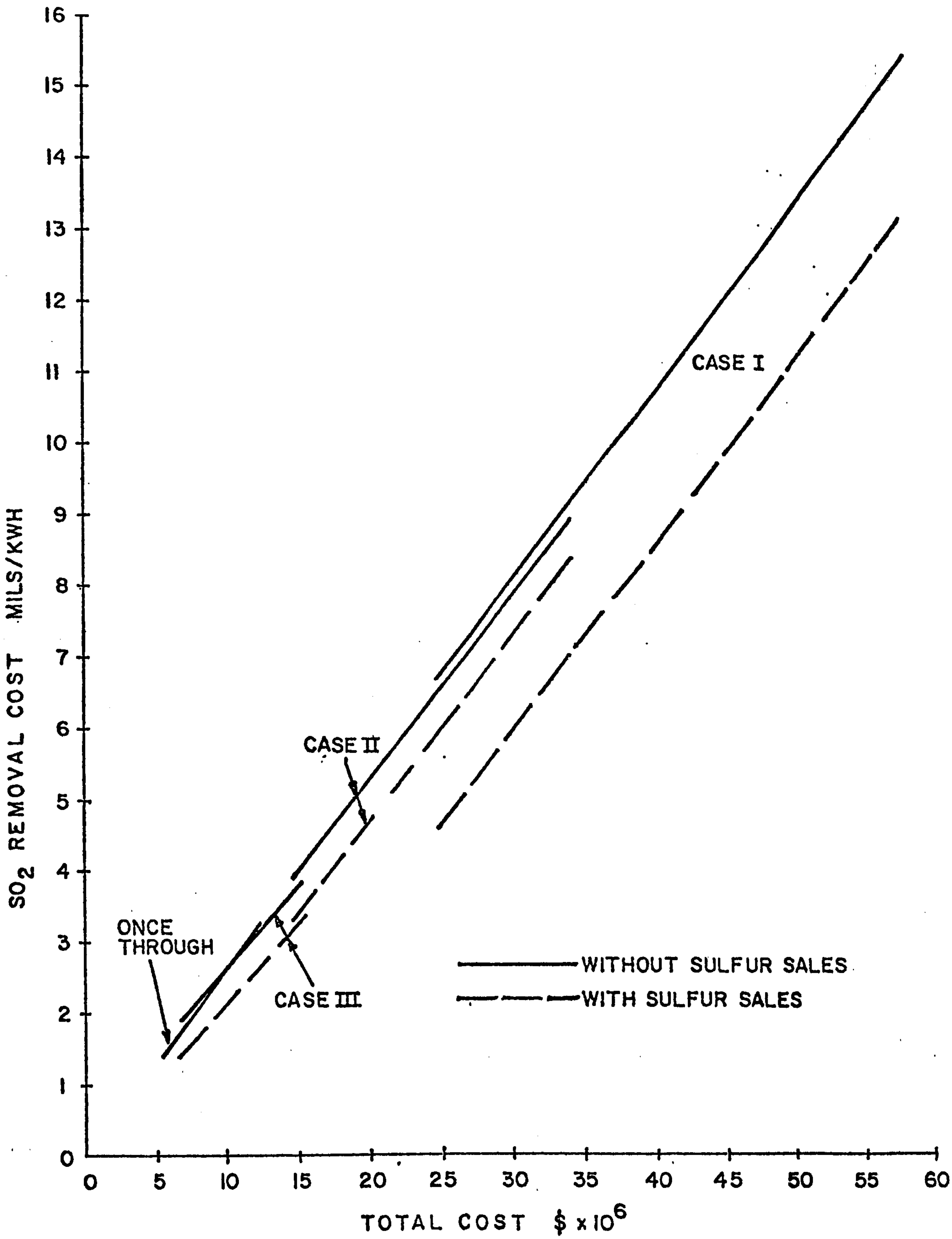
Capital Cost Sensitivity

FIGURE: D- 6



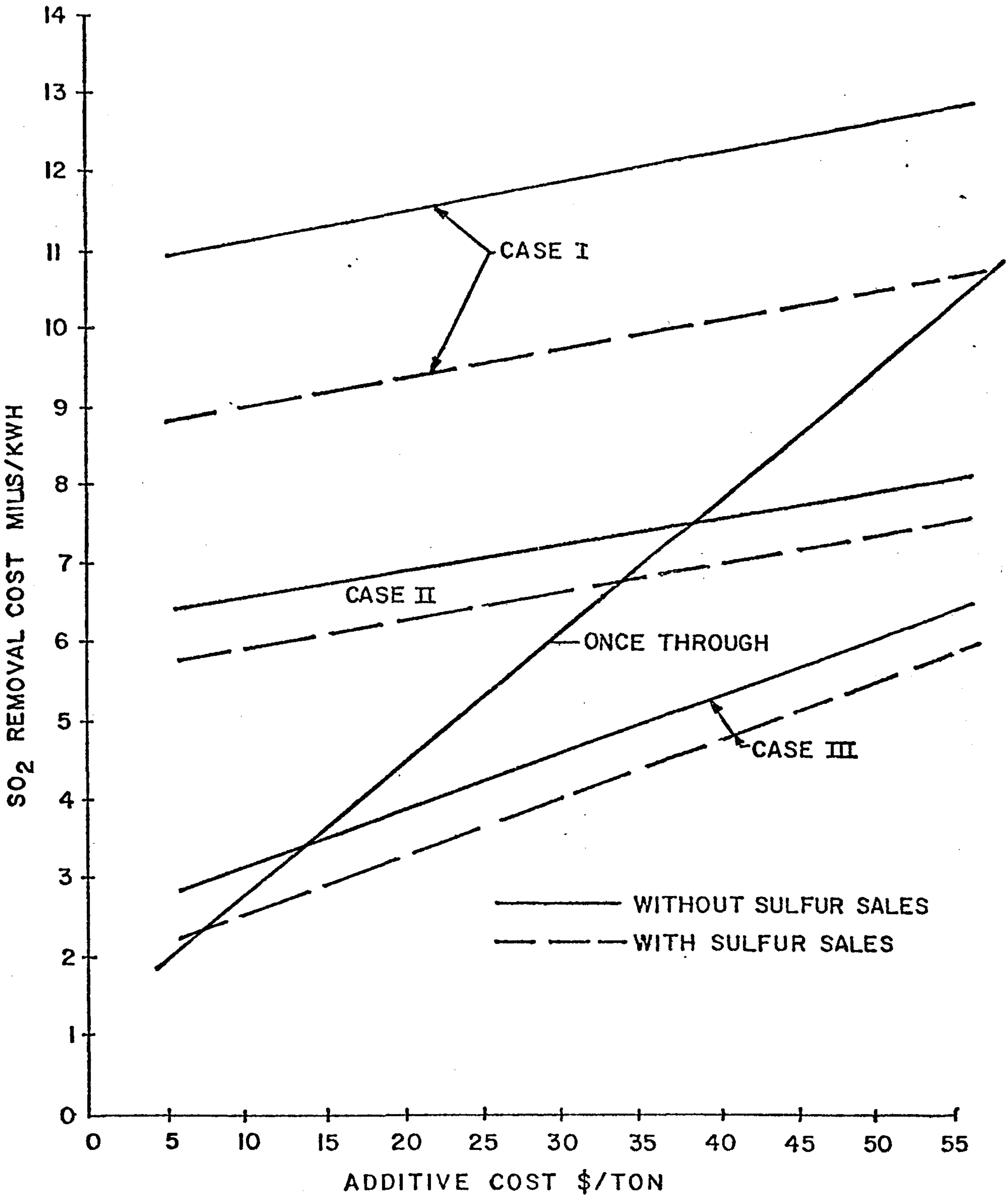
Direct Operating Cost Sensitivity

FIGURE: D-7



Total Operating Cost Sensitivity

FIGURE: D-8



Additive Cost Sensitivity

FIGURE D-9

TABLE D-16

APPROXIMATE CHEMICAL COMPLEX SPACE REQUIREMENTS

	Case I	Case II	Case III
Regeneration	75' x 75'	75' x 75'	75' x 75'
Koppers-Totzek Gasifier	75' x 75'	75' x 75'	-
Claus Sulfur Recovery	75' x 75'	75' x 75'	-
Claus Tail-Gas Cleanup	75' x 75'	75' x 75'	-
RESOX Sulfur Recovery	-	-	75' x 75'
Anthracity Coal Prep.	-	-	75' x 75'
Electrostatic Precipitator	-	-	75' x 25'
Anthracity Coal Storage	-	-	75' x 50'
Hydrogen Plant	-	75' x 50'	-
Hydrogen Sulfide Plant	-	50' x 25'	-
Cooling Tower, Waste Treatment and Liquid Storage	75' x 25'	75' x 25'	75' x 25'
Office, Control Room and Maintenance Shops	75' x 25'	75' x 25'	75' x 25'
Overall Area	200' x 200'	200' x 300'	200' x 250'

ACKNOWLEDGMENT

The assistance and cooperation of Argonne National Laboratory is acknowledged in developing the input information for Case III.

6.0 ADDITIONAL GUIDES FOR FURTHER DEVELOPMENT EFFORTS ON THE COMMERCIAL PLANT DESIGN

Various modifications may be made to the selected PFB/AFB commercial power plant. Each change has its own advantages and disadvantages which must be evaluated in order to determine if there is a net benefit.

In this section some of these modifications are presented along with a discussion of the pros and cons of each.

6.1 SUBSTITUTE BAGHOUSE FOR ELECTROSTATIC PRECIPITATOR (ESP)

A baghouse is capable of 99.4% dust capture efficiency which is more than adequate to meet the design requirements of the commercial plant. For the same capacity, the capital cost of a baghouse is generally lower than the cost of an ESP. However, differential maintenance and power costs must also be considered along with factors related to plant availability. Use of a baghouse would eliminate the uncertainty concerning the performance of ESP's used in FBC applications. However, since the low level economizer would have to be relocated upstream of the baghouse, its cost would increase due to the design changes required to accommodate the dirtier gas stream.

6.2 REDUCED STACK GAS TEMPERATURES

In a conventional power plant, a substantial amount of heat is lost through the stack. "Cold-end" corrosion of economizers, stacks, etc., due to the SO_3 content of the flue gases has been instrumental in keeping the stack temperature near 300° F in conventional plants. Fluidized bed combustion of coal in the presence of a sulfur dioxide sorbent promises to relax this constraint significantly. To date, the experimental results have shown very little SO_3 concentration in the flue gases leaving fluid bed units. If further experimentation confirms this low SO_3 concentration in the flue gas, the stack gas temperature may be lowered to perhaps 200° F and the feedwater entering the economizer to 160° F (REF. 13) without causing any appreciable corrosion problem. In addition to efficiency improvements, the gas volume passing through fans, stacks, etc., would be reduced. Extraction feedwater heaters would be eliminated, but economizer surface would increase. Furthermore, since the extraction steam quantities would be reduced, the size of the final turbine stages, condenser, cooling towers, etc., would have to increase. An evaluation of all of these factors is necessary to determine whether a reduction in stack temperature would be cost effective, regardless of any relaxation in acid dew point criteria which may be justified through further FBC testing.

6.3 DOUBLE-ENDED GENERATOR DRIVE

Costs can be reduced in most cases by applying gas turbine power input to both ends of the electric generator rotor. In the FT4 TWIN-PAC, the power turbines are "mirror images" of each other, so that they can drive the generator rotor from both ends. For a direct drive gas turbine, it is probably most economical to drive the generator from the front of one gas turbine and from the rear of the other. In the case of the selected cycle, this arrangement would simplify the selection of generator type. At a rating of 63.5 MW, the appropriate generator would be air cooled. Doubling the size to 127 MW would allow economical use of a hydrogen cooled generator with a reduction in specific cost and a slight increase in overall efficiency.

Further cost savings could result from commonality of some control systems, lubrication systems and starting systems. The enclosure and maintenance provisions can also be accomplished at lower cost in a TWIN-PAC arrangement. Problems associated with plant layout may cause increased costs which, if they exist, must be evaluated against the benefits.

6.4 VERY HIGH PRESSURE RATIO SYSTEMS

It is possible that the size of the PFB and associated cleanup equipment could be markedly reduced if the pressure ratio of the gas turbine was raised beyond the level considered in the cycle selection study. Pressures of 40 to 100 atmospheres could be investigated. The higher pressure could be obtained by cooling the compressor discharge air (at 10 to 16 atmospheres) and supplying the cooled pressurized air to the inlet of a high pressure compressor with a pressure ratio of 4 to 6 to 1. After combustion in the PFB, the air would pass through a high pressure turbine and be reheated one or more times in the process of expanding through turbines to atmospheric pressure. The system would be more complex than the selected cycle, but offers a possibility of improved efficiency and lower cost.

7.0 REFERENCES

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8.0 APPENDIX

8.1 REPORT ON DUCON GRANULAR BED FILTER

ENGINEERING EVALUATION STUDY

TO

DEVELOP AN OPTIMUM GRANULAR BED FILTER SYSTEM

FOR

A COMMERCIAL POWER PLANT UTILIZING

FOUR (4) PRESSURIZED FLUIDIZED BED COMBUSTORS

EACH WITH ITS OWN COLLECTION SYSTEM

PREPARED BY: THE DUCON COMPANY, INC.
147 E. Second Street
Mineola, N. Y. 11501

PROJECT MANAGER: ROBERT A. PIRRMANN
CHIEF ENGINEER

CONTRACT NO.: C77-903

B & W PURCHASE ORDER NO.: 717363DU

INDEX

SECTION A

STATEMENT OF WORK

The Ducon Company, Inc. has entered into a contractual agreement with the Babcock & Wilcox Company to perform an engineering study and cost evaluation of a granular bed type particulate removal system to be used in conjunction with a pressurized fluidized bed combustor.

The overall conceptual design for the commercial pressurized fluidized bed plant calls for a total of two gas turbines and four pressurized fluidized bed combustors. In this scheme, two combustors operate in parallel to feed a gas turbine and each combustor would have its own dust collection system.

The work performed by Ducon under this contractual agreement is essentially divided into two tasks:

Task I is an optimization study to determine physical equipment sizing and operating characteristics and/or parameters.

Task II is a finalization of equipment sizing and design culminating in a detailed cost analysis and formal report.

SECTION B

DESIGN CONDITIONS AND REQUIREMENTS

Each granular bed filter system is based upon handling the gas stream exiting a single pressurized fluidized bed combustor. The inlet gas conditions and composition are as follows:

Inlet Gas Volume	37416 ACFM
Inlet Gas Temperature	1650^{o}F
Inlet Gas Pressure	136.0 PSIA
Inlet Gas Density	0.1803 Lbs/FT^3

Inlet Gas Composition:

Component	Lbs/Hr.	Lb-Moles/Hr.
O_2	11344	354.35
N_2	282976	10101
Ar	5055.5	126.7
SO_2	475.5	7.45
CO_2	89678	2038
H_2O	16002	888
TOTAL	405531	13515.5

The particulate matter to be collected consists of fly ash and a mixture of dolomite, half-calcined dolomite and half calcined, partially sulfated dolomite with the following size distribution.

Particle Diameter (Microns)	Percent (%) By Weight $\geq$ Stated Particle Diameter
100	27.46
80	31.53
60	36.18
40	43.27
20	57.75
10	73.02
8	76.90
6	82.06
4	88.48
2	96.16

Inlet Dust Loading	6474 Lbs/Hr. 20.2 Grains/ACF
Particle Density (Assumed)	125 Lbs/CF

Specific Design Requirements

1. The maximum allowable unrecoverable pressure loss that can exist between the inlet and outlet of the dust collection equipment is 4.00 PSI.

2. All insulation is to be located adjacent to the inside surface of the exterior walls of the dust collection equipment.

3. The metal temperature of the outside surface of the exterior walls of the dust collection equipment is to be maintained at 250° F when the ambient air temperature is 80° F and the flue gas temperature is 1650° F.

4. Each pressurized fluidized bed combustor is to have its own dust collection system.

5. Each dust collection system must be capable of being cleaned without causing a decrease in the power output of either turbine.

SECTION C

GRANULAR BED FILTER BACKGROUND

Granular or gravel bed filtration is not a new concept. Patents for the filtration of gases and liquids by means of a so called gravel bed date back to the late 1800's. The Ducon Company's Granular Bed Filter has been under development for approximately twelve years.

Ducon's original developmental design concept was to provide a device capable of filtering particulate matter from a gas stream at high temperature and/or pressure. In addition, the development program was to incorporate a means whereby the filtration media could be cleaned without resorting to moving parts within the gas stream or the removal of filtration media during operation. Original conceptual work resulted in a patented system of blow-back gas whereby the principle of operation would allow an individual filtration element to be taken out of service momentarily and the filter beds fluidized. The fluidization of the filter beds entrain the collected solids and deposit them in a collection hopper for withdrawal from the filter housing. The original element design incorporated a series of cones with inner and outer screens. The filtration principle with this design was proven. In addition, the blow-back principle was also determined to be acceptable. The limitations of this design were that due to the conical cross section of the media retainers, the height of the bed varied from the inner diameter to the outer diameters. Although the cleaning principle was proven with this configuration, upon fluidizing the granular material, the fluidized gas became maldistributed through the bed due to the lower pressure drop in the shallower section. This configuration also resulted in a churning of the filter media bed so that effective cleaning of the backwash cycle over an extended period of operation would be questionable. The second limitation to this design was that by utilizing inner and outer screens, the conical retainers for the filtering media could not be positively attached to the screens. This design configuration was then abandoned, and a configuration utilizing a flat bed or donut cross section was adapted.

It was determined through laboratory testing in order to achieve proper distribution in a donut cross section on the backwash cycle that the bed was limited to an annular width of approximately 1-1/2 with inner and outer diameters of the donut being 5" and 8" respectively, established as maximums. This limitation necessitated stacked beds resulting in a design requiring a great deal of assembly labor. In addition, outer inlet screens were required since the configuration limited the height of each filter compartment. Additional developmental work was performed to overcome these limitations and to remove the inherent quality control problems associated with these small elements. The results of this program led to a rectangular cross section design which was tested in Ducon's laboratory. The rectangular cross section has a number of advantages. The mechanical constraints in the fabrication of elements was now effectively removed. In addition, 3.6 times the filter area could be supplied at approximately the same cost per element. Continuing development has established a design whereby each element is presently designed with 12 filtering compartments. Each compartment is 6" wide x 36" long with a sand bed approximately 2 inches deep, resulting in a net filtering area of 18 ft^2 per element. The compartments are supplied with a clean gas discharge between two rows of six compartments each.

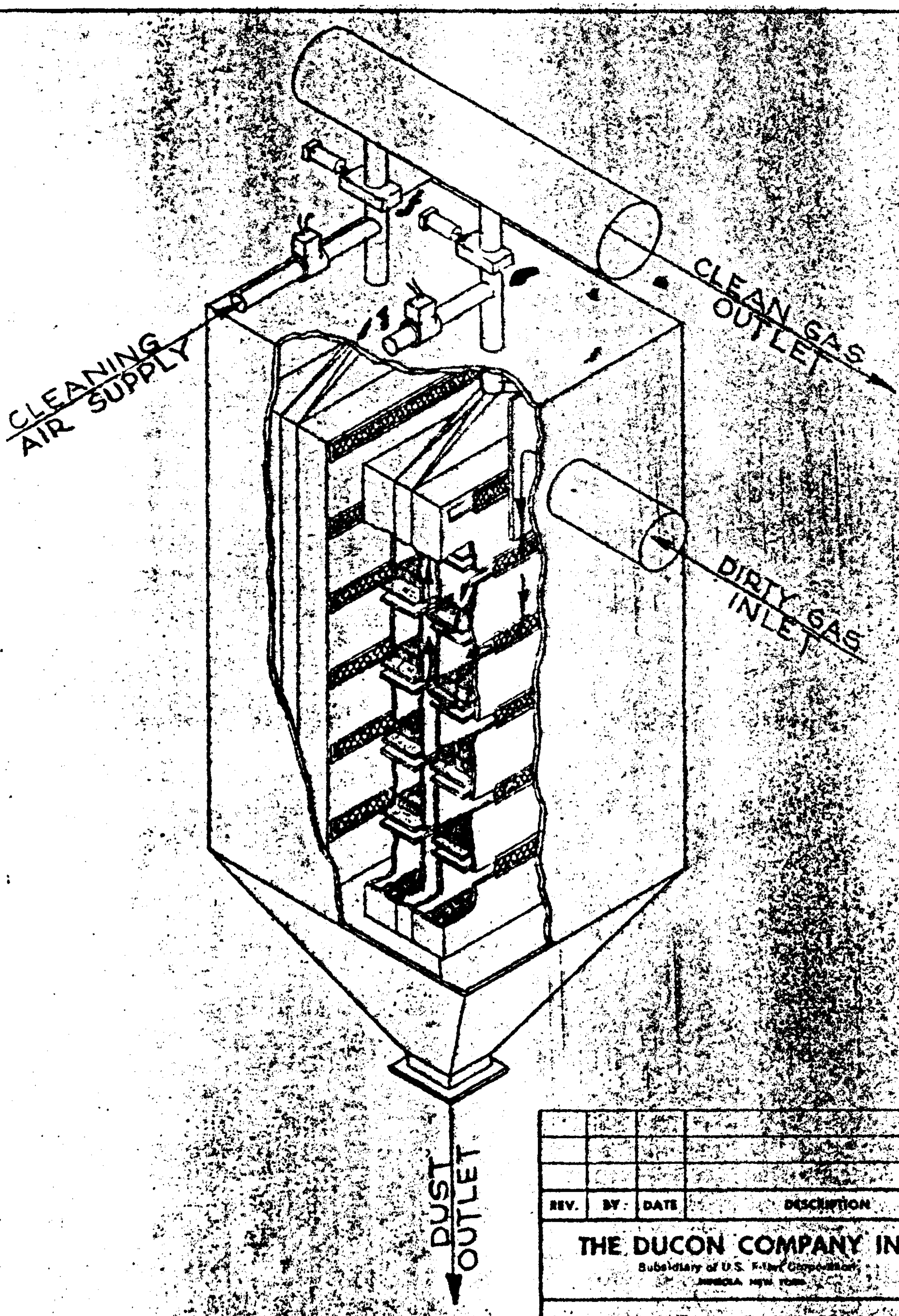
However, with all the design changes made to date, the principle of operation has not changed since its original conception 12 years ago.

SECTION D

PRINCIPLES OF OPERATION

Dirty gas enters the Ducon Granular Bed Filter through the inlet duct. The filter elements are suspended within a vessel. Drawing No. S-4195 attached shows a simplified arrangement of two elements within a single vessel. Particulate matter and gases pass through the inlet slots and through parallel 1-1/2" beds of sand. The sand is between 250 and 595 microns in diameter. The dust particles are entrapped in the interstices of the granular material as the cleaned gas passes through to the clean gas plenum chamber of the element. The clean gas streams from all the elements are joined together in the vessel plenum and a main exhaust duct.

When the accumulation of collected particulates causes the pressure drop to reach a specified level, an individual element is isolated from the other elements so that it can be backwashed. The primary cleaning and backwash sequence is shown in Drawing No. S-4196 (attached) where a single element is shown. Individual elements are isolated from the clean gas outlet by inducing sufficient volume by means of compressed motive gas to overcome and reverse the normal flow of air. This introduction of both motive and induced gas is sufficient to expand and fluidize the beds. The fluidization of the filtering sand releases the fine dust particles, entraining them in an upward gas flow. The agglomerated dust settles from this flow into the dust collection hopper. This sequence is then repeated with each element until all the elements have been cleaned. The cleaning cycle will depend on the characteristics of the dust being filtered, the dust load and the ratio of filtering gas rate to bed surface area.

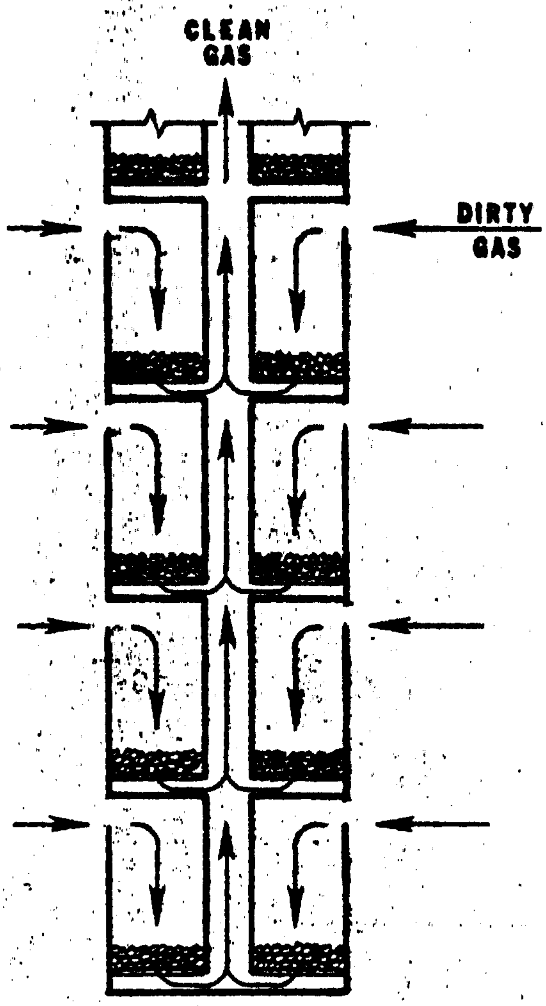


REV.	BY	DATE	DESCRIPTION

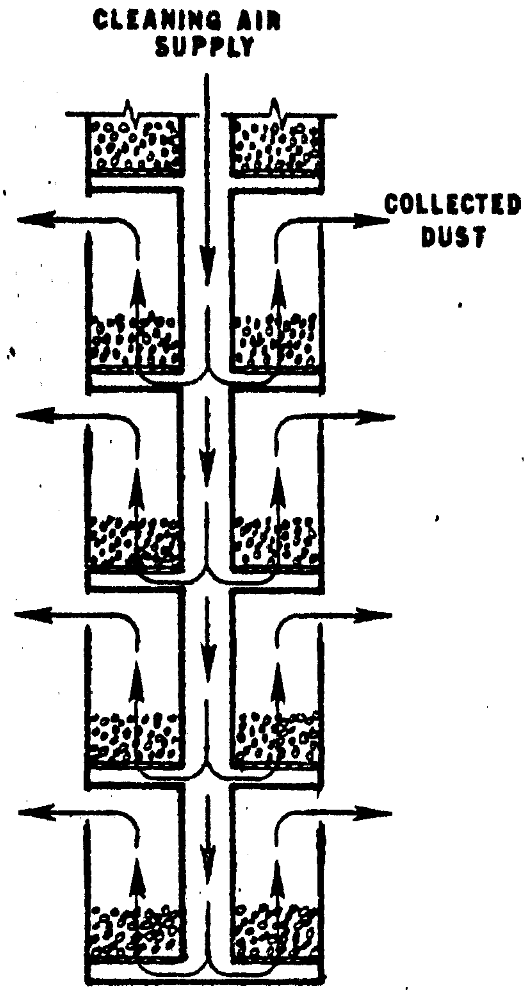
THE DUCON COMPANY INC.
Subsidiary of U.S. Filter Corporation

DUCON
GRANULAR BED FILTER

DR. BY	K.F.	[redacted]	REV.
SCALE			



COLLECTION CYCLE
(FIG. 2)



CLEANING CYCLE
(FIG. 3)

REV.	BY	DATE	DESCRIPTION

THE DUCON COMPANY INC.
Subsidiary of U.S. Filter Corporation
MINEOLA, NEW YORK

DUCON
GRANULAR BED FILTER

DR. BY	KF	S-4196	REV.
SCALE			

SECTION E

DESIGN AND SELECTION PROCEDURE

Ducon has conducted preliminary engineering studies in an attempt to optimize a Granular Bed Filter arrangement for utilization in a power plant with fluidized bed combustors.

Two approaches or alternate selections were investigated. Firstly, selection of a single vessel which would provide effective filtration of the total gas stream from one pressurized fluidized bed combustor. Secondly, selection of multiple shop fabricated vessels which would be connected in parallel to handle the gas stream from one combustor.

As the Ducon Granular Bed Filter is presently a developmental item, the selection procedure to determine the amount of filter area required to treat any given process gas stream is limited to the gas-to-filter ratios which have been established through laboratory testing. While the literature may report ranges anywhere from 40 to 100 ACFM/Sq. Ft. of filter bed surface area, it was decided that, at its present state of development, the optimum gas-to-filter ratio would be established as approximately 50 ACFM per square foot of filter area for study purposes.

Based upon a per combustor gas flow of 37416 ACFM as outlined under "Design Conditions and Requirements" a filter bed surface area of approximately 748 ft.^{2} would be required for effective filtration. Depending upon the cleaning cycle frequency, up to approximately 10% of the filtering area could be off-steam at any given time, thus increasing the face velocity through the remaining elements. As a result, anywhere from 5-10% is added to the original design filtering area so as to provide the optimum filtering rate at steady state-equilibrium conditions with continuous cleaning.

An additional 5% of filter area would give a total filter area requirement of 785 ft.^{2}. Each filter element contains a net filter area of 18 ft.^{2}. Thus, the specified gas volume would require a total of 44 elements (42.6) for effective filtration.

Alternate 1 - A single field erected pressure vessel containing 44 filtering elements would provide the proper filtering ratio as described above. Previous optimization studies of vessel size versus number of filtering elements developed a design standard arrangement of a twenty-five (25) foot diameter vessel containing forty-four (44) elements which was selected for the Alternate 1 arrangement.

Alternate II - Multiple shop fabricated pressure vessels. It has been established that the maximum diameter which can be conveniently shipped from any fabricating facility to any plant site would be approximately thirteen (13) feet. The optimization study of vessel size versus number of filtering elements developed a design standard arrangement of a thirteen (13) foot diameter vessel containing ten (10) elements. In order to approximate the required total filtering area of 785 Ft^2, five (5) vessels would be required. As five (5) vessels with the (10) elements each would provide a total filtering area of 900 Ft^2, the number of elements per vessel was reduced to nine (9), thus providing 810 Ft^2 of filtering area. It was decided that this would provide a more equitable cost comparison between the two arrangements.

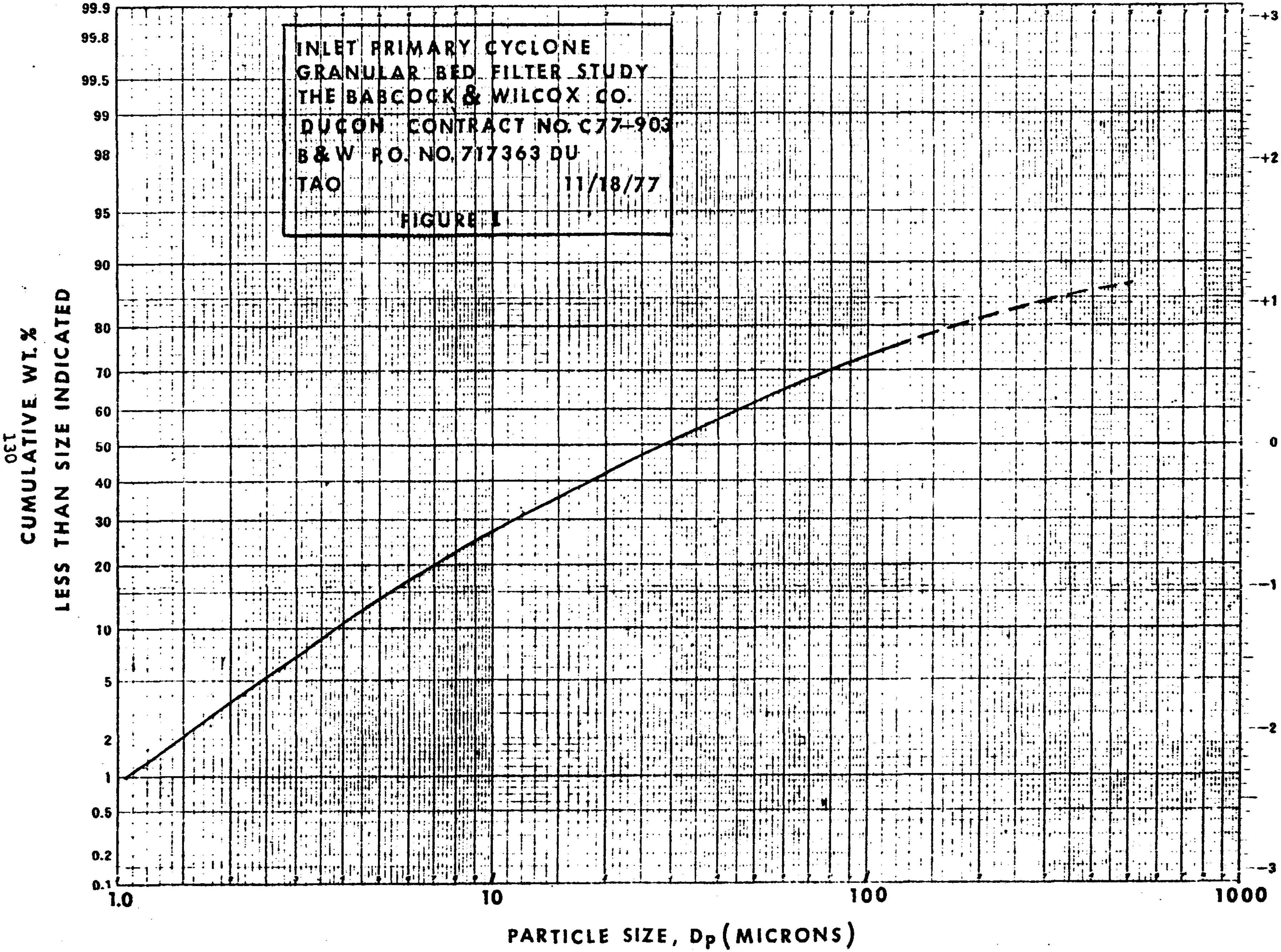
Precleaner:

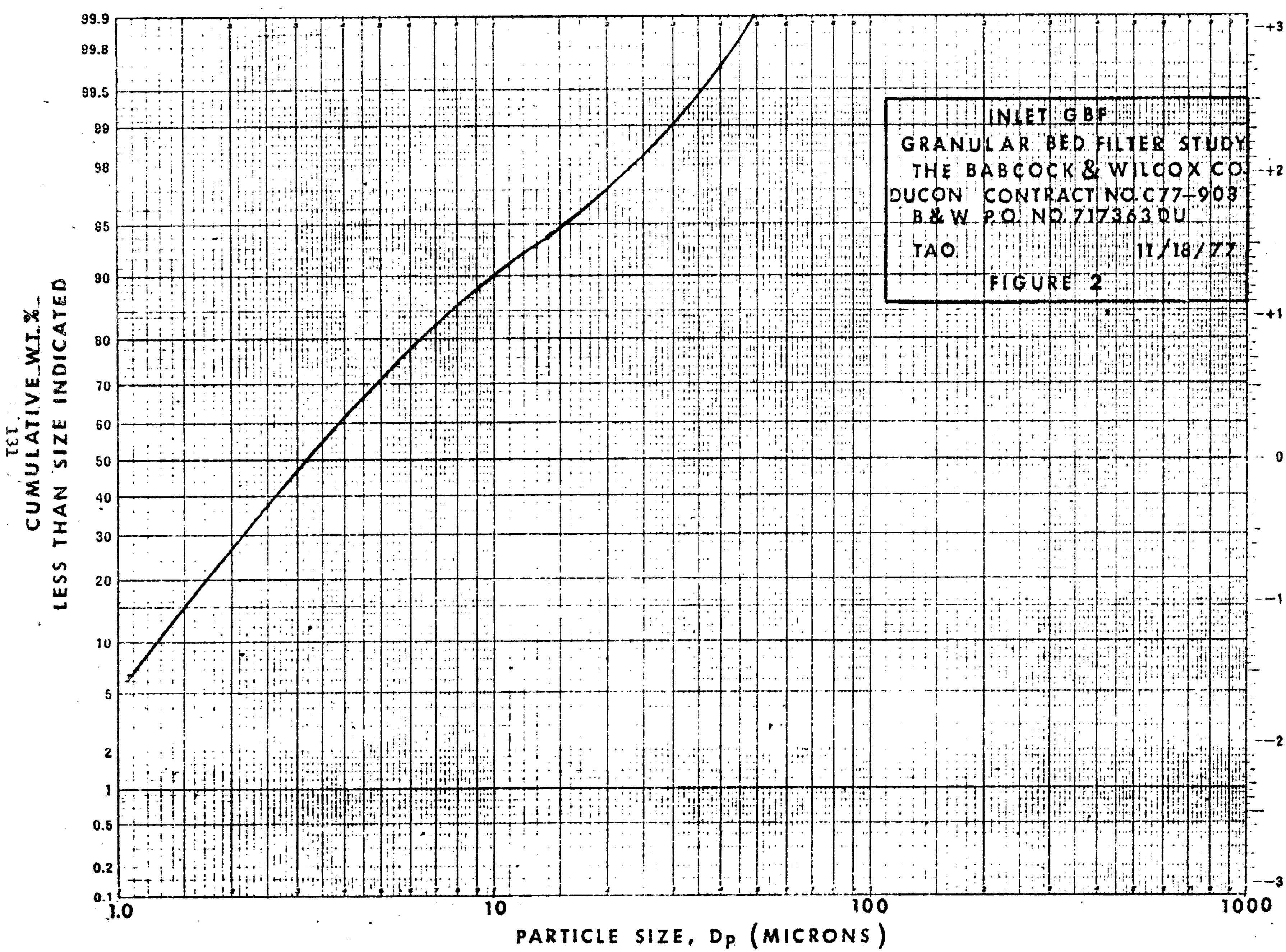
The anticipated dust loading of 20.2 grains/ACF as specified under "Design Conditions and Requirements" would generally be considered high enough to consider a precleaner ahead of the Granular Bed Filter. It was decided to include a cyclone precleaner as a part of the overall dust collection system and in so doing, reduce the dust load to the Granular Bed Filter.

A single, large diameter cyclone was selected to handle the per combustor volume of 37416 ACFM. The cyclone design was based upon an inlet velocity of approximately 3800 fpm and an operating differential pressure drop of 1 PSI.

Through the use of a cyclone precleaner, the dust load to the Granular Bed Filter can be reduced from 20.2 grains/ACF to approximately 3.2 grains/ACF. The Granular Bed Filter is capable of handling either dust load but at a lower rate less frequent cleaning would be necessary and Granular Bed Filter ΔP is reduced considerably.

The anticipated particle size distribution of solids to the dust collection system is shown in Figure 1. In the system where a precleaner is provided, this curve represents the solids to the cyclone. In Figure 2, the cyclone outlet distribution is shown.


INLET PRIMARY CYCLONE
GRANULAR BED FILTER STUDY
THE BABCOCK & WILCOX CO.
DUCON CONTRACT NO. C77-903
B & W P.O. NO. 717363 DU
TAO 11/18/77
FIGURE 1
CUMULATIVE WT. %
LESS THAN SIZE INDICATED
PARTICLE SIZE, Dp (MICRONS)
99.9
99.8
99.5
99
98
95
90
80
70
60
50
40
30
20
10
5
2
1
0.5
0.2
0.1
1.0
10
100
1000
+3
+2
+1
0
−1
−2
−3


INLET GBF
GRANULAR BED FILTER STUDY
THE BABCOCK & WILCOX CO.
DUCON CONTRACT NO. C77-903
B & W P.O. NO. 717363 DU
TAO 11/18/77
FIGURE 2
CUMULATIVE W.T.%
LESS THAN SIZE INDICATED
PARTICLE SIZE, Dp (MICRONS)
99.9
99.8
99.5
99
98
95
90
80
70
60
50
40
30
20
10
5
2
1
0.5
0.2
0.1
1.0
10
100
1000
+3
+2
+1
0
−1
−2
−3
131

SECTION F

CYCLONE PRECLEANER

A total of four (4) cyclones 10 feet in diameter by approximately 42 feet high would be required for the commercial power plant with four (4) pressurized fluidized bed combustors, each with its own dust collection system.

Discussion will be limited to a per combustor basis. As previously stated, it is not absolutely necessary for performance that a cyclone precleaner be utilized. The main advantage of considering a cyclone precleaner ahead of the Granular Bed filter is that system equilibrium pressure drop is lower and the incremental cost of including a cyclone is reasonable.

Construction:

The cyclone precleaner selected is a Duclone, Size 975, Type M, Model 1200/90, See Figure 3. The cyclone would be shop fabricated of Carbon Steel SA515 Grade 70, outlet tube would be Hasteloy "X" and flanges would be class 175 slip on pressure vessel flanges. Plate thicknesses would be as follows:

Eliptical Head ----------------	3/4"
Cylindrical Section -----------	3/4"
Inlet Section -----------------	7/8"
Cone Section ------------------	7/8"
Outlet Tube -------------------	1/4"

In order to meet the external skin temperature limitation of 250^{o} F at design operating conditions with an ambient temperature of 80^{o} F as well as provide adequate abrasion protection, a dual layer lining was selected. Three inches of RESCO RS 3-35 insulating castable and three inches of RESCO RA-22 abrasion resistant castable provide the dual purpose protection. Anchoring tynes of Type 304 Stainless Steel will be located on 9" triangular centers.

Weight:

The Size 975 Ducon Cyclone complete with installed lining, as shown on Figure 5 will weigh approximately 82,000 Lbs. This weight represents the sum of the following components:

Shell		41,540 Lbs.
Lining		40,460 Lbs.
	Total	82,000 Lbs.

The cyclone will operate at a differential pressure drop of 1.0 PSI at design conditions. See cyclone design sheet, TABLE 1.

Cyclone performance is approximately in accordance with Fractional Efficiency Curve No. 24A.

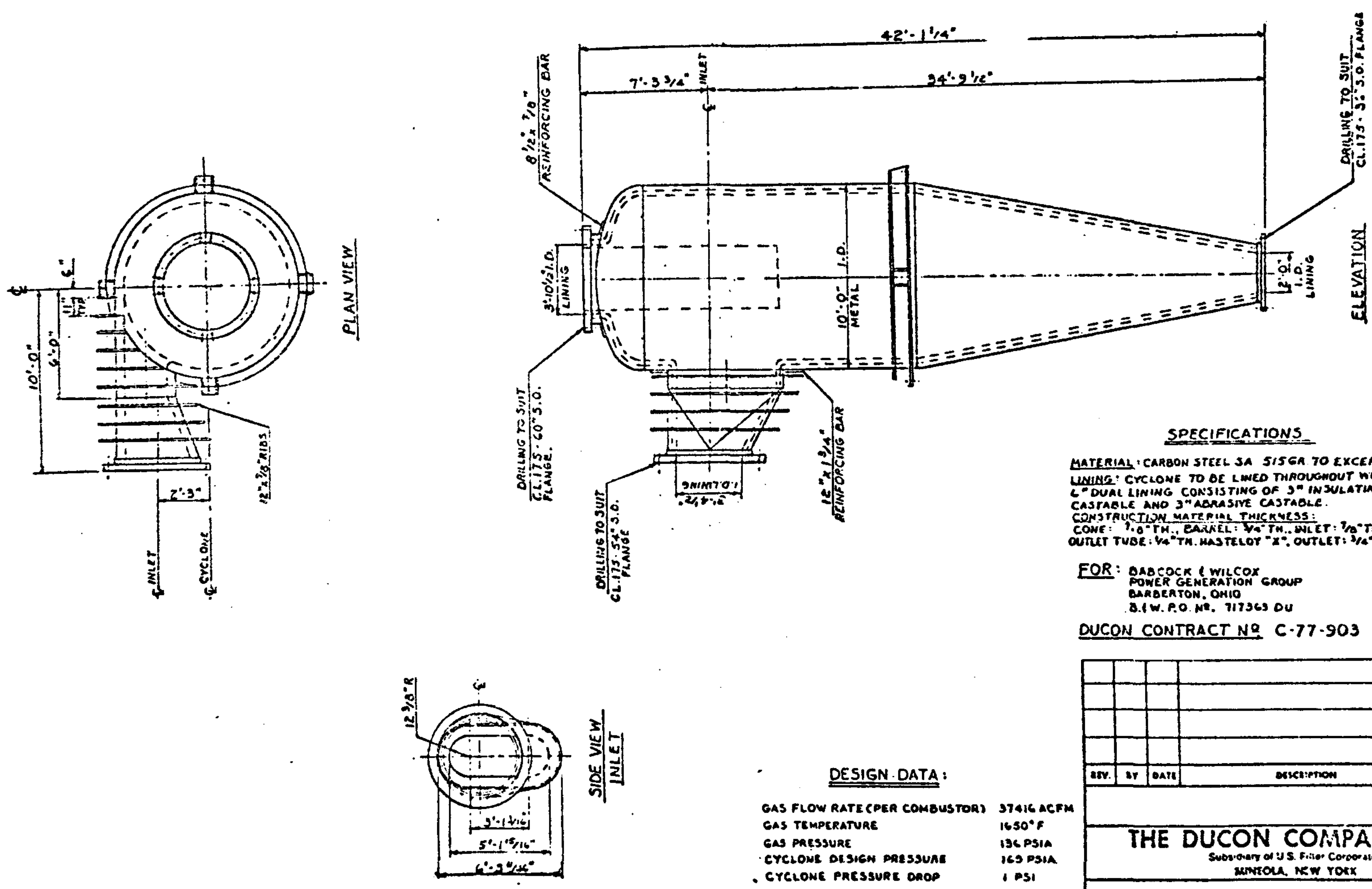


FIG. 3
CYCLONE COLLECTOR
SIZE 975 TYPE M MODEL 1200/90

SPECIFICATIONS

MATERIAL: CARBON STEEL SA 515GR 70 EXCEPT AS NOTED.
LINING: CYCLONE TO BE LINED THROUGHOUT WITH 6" DUAL LINING CONSISTING OF 3" INSULATING CASTABLE AND 3" ABRASIVE CASTABLE.
CONSTRUCTION MATERIAL THICKNESS:
CONE: 7/8" TH., BARREL: 3/4" TH., INLET: 7/8" TH.,
OUTLET TUBE: 1/4" TH. HASTELOY "X", OUTLET: 3/4" TH.

FOR: BABCOCK & WILCOX
POWER GENERATION GROUP
BARBERTON, OHIO
B.&W. P.O. Nº. 717363 DU

DUCON CONTRACT Nº C-77-903

DESIGN DATA:

GAS FLOW RATE (PER COMBUSTOR)	37416 ACFM
GAS TEMPERATURE	1650°F
GAS PRESSURE	136 PSIA
CYCLONE DESIGN PRESSURE	165 PSIA
CYCLONE PRESSURE DROP	1 PSI

APPROX WT. LINED - 82,000#

REV.	BY	DATE	DESCRIPTION	CH. BY	DATE

THE DUCON COMPANY INC.
Subsidiary of U.S. Filter Corporation
MINEOLA, NEW YORK

CYCLONE COLLECTOR
SIZE 975 TYPE M MODEL 1200/90

DR. BY	B.C. 11/16/77	APP. BY		DRAWING NO.	REV.
CH. BY		SCALE		K-77-903-3	

CYCLONE FRACTIONAL EFFICIENCY CURVE.
CURVE #24A

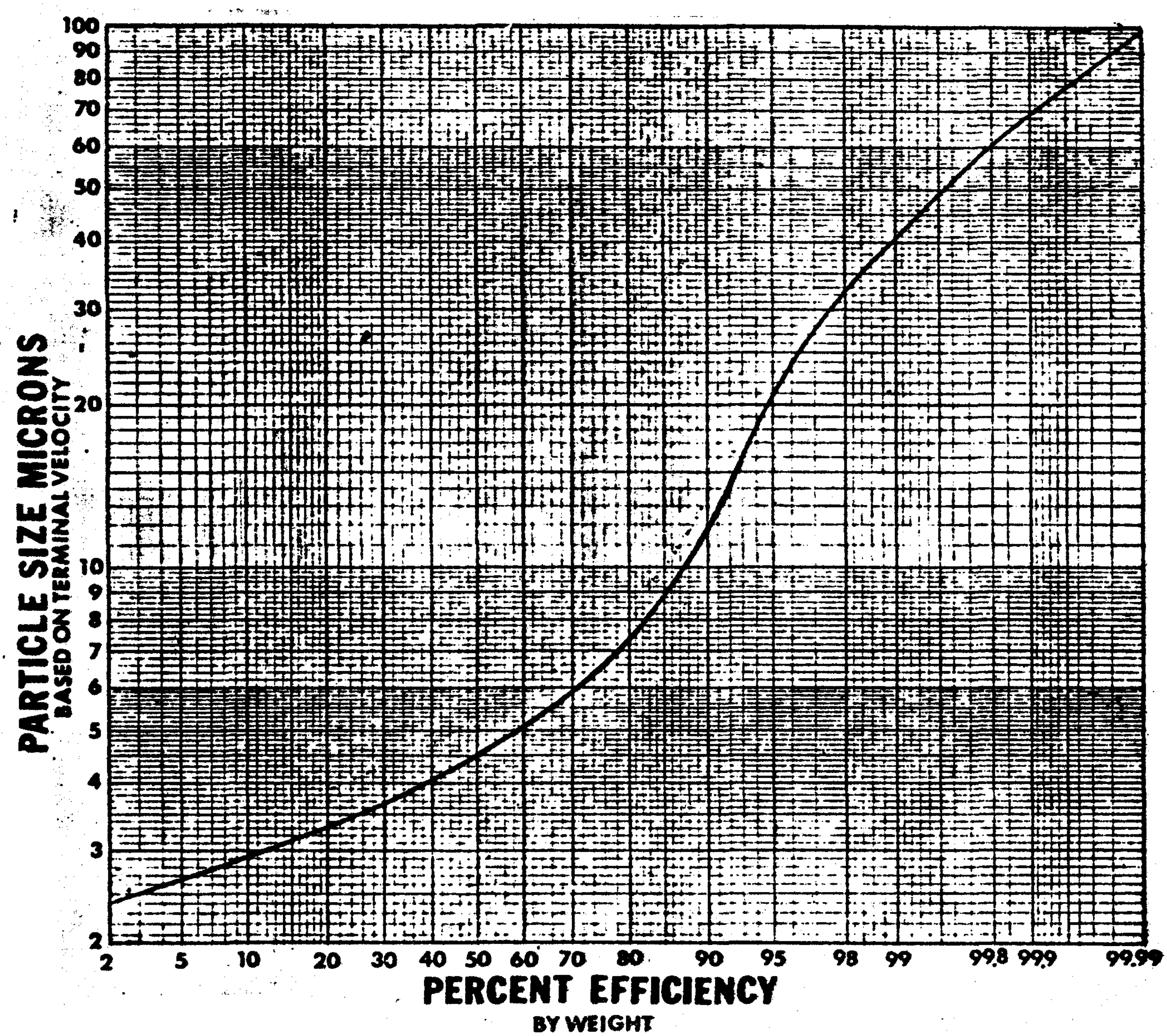


TABLE 1

DUCON CYCLONE OPERATING AND DESIGN DATA SHEET

PROGRAM: CYCLONE 1 DATE: 11/11/77

THE DUCON CO., INC.
147 E. SECOND STREET
MINEOLA, N.Y. 11501

CUSTOMER:

CUSTOMER REFERENCE NUMBER:

DUCON REFERENCE NUMBER:

DUCON CYCLONE OPERATING AND DESIGN DATA SHEET

CYCLONE SELECTION: NO. OF STAGES	1
TOTAL GAS FLOW	404766 LBS/HR
TOTAL INLET GAS FLOW AT OPERATING CONDITION	37416 ACFM
OPERATING INLET GAS DENSITY & COND.	0.1803 LBS/CF
OPERATING VISCOSITY & COND.	0.0460 CP
OPERATING GAS TEMPERATURE	1650 DEG.F.
OPERATING GAS PRESSURE	136.00 PSIA
PARTICLE DENSITY	125.0 LBS/CF
OPERATING SOLIDS INLET LOADING	6478 LBS/HR
OPERATING SOLIDS INLET LOADING	20.20 GRAINS/ACF
INDIVIDUAL HOPER CYCLONE SYSTEM	

CYCLONE STAGES	FIRST
CYCLONE SIZE	975
CYCLONE TYPE	M
CYCLONE MODEL	1200
CYCLONE DIAMETER	108.17 INCHES
NO. OF CYCLONES	1
OPERATING GAS FLOW PER CYCLONE	37416 ACFM
CYCLONE INLET VELOCITY	64.0 FT/SEC
CYCLONE SALTATION VELOCITY	41.0 FT/SEC
CYCLONE FRACTIONAL EFFICIENCY CURVE NO.	23.1
CYCLONE OUTLET TO INLET RATIO	0.91
CYCLONE PRESSURE DROP	28.0 IN. WG.
CYCLONE PRESSURE DROP	1.011 PSI.
CYCLONE DIP-LEG SUCTION	20.0 IN. WG.
CYCLONE DIP-LEG DIAMETER	18 IN.
DIP-LEG SOLIDS RATE	0.86 LBS/SEC-SQ. FT.
CYCLONE SOLIDS LOSS RATE	1023.48 LBS/HR
CYCLONE SOLIDS EMISSION	3.191 GRAINS/ACF
ACC. CYCLONE EFFICIENCY	84.20%

TABLE 1 - Continued

CYCLONE SYSTEM PARTICLE DISTRIBUTION

PARTICLE SIZE LESS THAN (MICRONS)	INLET ACC. WGT. PERCENT	LOSS ACC. WGT. PERCENT	COLLECTED ACC. WGT. PERCENT
1.07	1.00	6.330	0.00
2.52	6.00	37.895	0.02
4.16	11.00	62.374	1.36
5.56	16.00	75.398	4.86
7.15	21.00	83.563	9.26
9.30	26.00	88.804	14.22
11.97	31.00	92.315	19.50
15.29	36.00	94.921	24.95
19.38	41.00	96.840	30.52
24.42	46.00	98.224	36.20
30.59	51.00	99.126	41.97
38.12	56.00	99.662	47.81
49.23	61.00	99.904	53.70
66.16	66.00	99.988	59.62
92.29	71.00	100.000	65.56
123.74	75.00	100.000	70.31

SECTION G

GRANULAR BED FILTER

ALTERNATE I

A total of four (4) vessels, 25 feet in diameter would be required for the commercial power plant with four (4) pressurized fluidized bed combustors, each with its own dust collection system.

Discussion will be limited to a per vessel or per combustor basis.

Construction:

The Ducon Granular Bed Filter selected is a Size 44/792, see Figure 4. The 25 Ft. O.D. filter vessel would be field fabricated of Carbon Steel SA515 Grade 70 and flanges would be 350# or 400# ASTM 105 or 181 Grade I or II. Plate thicknesses would be as follows:

Eliptical Head ---------------	1-1/4"
Cylindrical Section ----------	1-1/4"
Cone Section -----------------	1-3/8"

Granular Bed Filter elements, ejectors, plenum, pulse air manifold and piping to be Hasteloy "X". Plate thicknesses would be as follows:

Filter Elements --------------	10 Ga.
Ejectors ---------------------	10 Ga.
Plenum -----------------------	1/2"
Manifold & Piping ------------	Schedule 40 Pipe

In order to meet the external skin temperature limitation of 250° F at design operating conditions with an ambient temperature of 80° F a 4" gunnited lining of insulating castable, RESCO RS 3-35 has been selected. Anchoring tynes of type 304 Stainless Steel will be located on 9" triangular centers.

Weight:

The 25 Ft. O.D. vessel including lining, plenum, contained filter elements (loaded with sand filter media), ejectors and blowback air lines extending to the limit of all flanges as shown on Figure 4 will weigh approximately 375,000 lbs.

This weight represents the sum of the following components:

Vessel	243,800 Lbs.
Lining	38,000
Plenum	20,000
Elements	55,000
Sand Filter Media	11,000
Ejectors	4,800
Pulse Air Piping	2,400
TOTAL	375,000 Lbs.

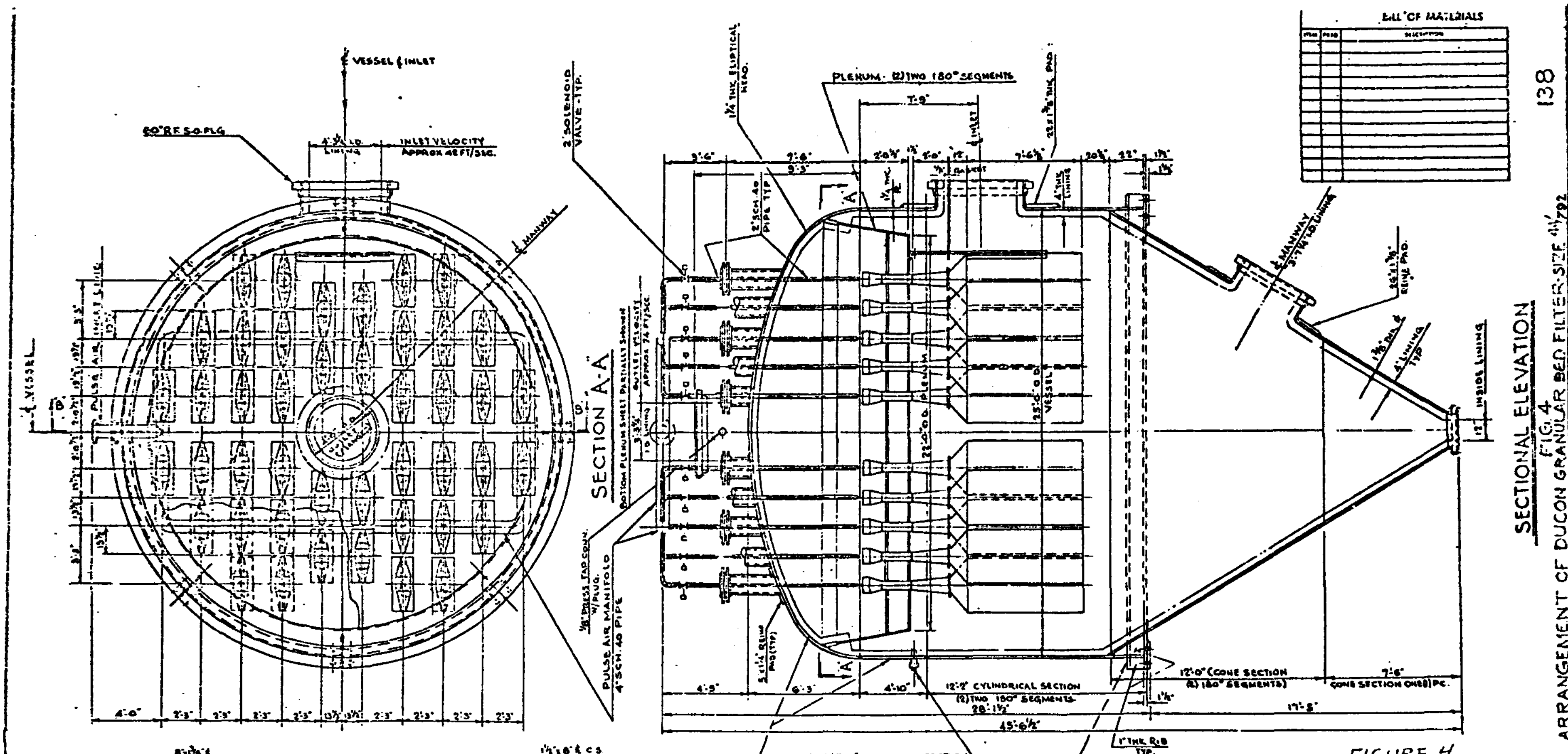


FIG. 4
ARRANGEMENT OF DUCON GRANULAR BED FILTER-SIZE 44/792

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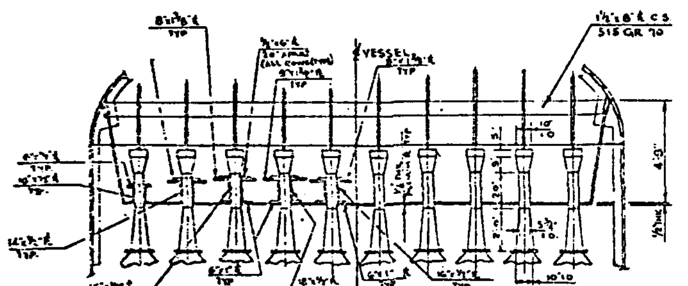


SECTION 'B-B'

FIGURE 4

SPECIFICATIONS

(1) VESSEL TO BE FABRICATED FROM C.S. SA 515 GR-70
(2) VESSEL FLANGES TO BE 350# OR 400# ASTM-105 OR 181 GR. I OR II
(3) GRANULAR BED FILTER ELEMENTS: EJECTORS: PLENUM, PULSE AIR MANIFOLDING AND PIPING TO BE HASTELOY "X"
(4) GRANULAR BED FILTER ELEMENTS #10GA
(5) EJECTORS #10GA

DESIGN DATA:

GAS FLOW RATE (PER COMBUSTOR)	37416 ACFM
GAS FLOW RATE (PER VESSEL)	37416 ACFM/VESSEL
GAS TEMPERATURE	1650°F
GAS PRESSURE	136 P.S.I.A.
VESSEL DESIGN PRESSURE	165 P.S.I.A.
APPROX. WT.	375,000#

FOR
BABCOCK & WILCOX
POWER GENERATION GROUP
BARBERTON, OHIO
DUCON CONTRACT NO. C-77903
B&W. P.O. NO. 717363 DU

REV	BY	DATE	DESCRIPTION	CH	APP
A			ADJUSTED DESIGN STANDARD DWG FOR GB FILTER SIZE 44/792 TO SUIT B&W SPEC. AND B&W P.O. NO. 717363 DU		

THE DUCON COMPANY INC.
MINEOLA, NEW YORK

ARRANGEMENT OF
DUCON GRANULAR BED FILTER
SIZE 44/792
ONE (1) REQ'D PER COMBUSTOR

DRAWING NO. K-77903-1 REV. A

SECTION H

GRANULAR BED FILTER

ALTERNATE II

A total of twenty (20) vessels, 13 feet in diameter would be required for the total commercial power plant with four (4) pressurized fluidized bed combusters, each with its own dust collection system.

Each combustor would have a dust collection system consisting of five (5) shop fabricated vessels, 13 feet in diameter.

Construction:

The Ducon Granular Bed Filter selected is a Size 9/162. See Figure 5. The 13 Ft. O.D. filter vessel would be fabricated of Carbon Steel SA515 Grade 70 and flanges would be 350# or 400# ASTM 105 or 181 Grade I or II. Plate thicknesses would be as follows:

Eliptical Head----------------1"
Cylindrical Section----------1"
Cone Section-----------------1"

Granular Bed Filter elements, ejectors, plenum, pulse air manifold and piping to be Hasteloy "X". Plate thicknesses would be as follows:

Filter Elements----------------10 Ga.
Ejectors-----------------------10 Ga.
Plenum-------------------------1/2"
Manifold & Piping--------------Schedule 40 Pipe

In order to meet the external skin temperature limitation of 250^{o}F at design operating conditions with an ambient temperature of 80^{o}F a 4" gunnited lining of insulating castable, RESCO RS 3-35 has been selected. Anchoring tynes of type 304 Stainless Steel will be located on 9" triangular centers.

Weight:

The 13 Ft. O.D. vessel including lining, plenum, contained filter elements (loaded with sand filter media), ejectors and blowback air lines extending to the limit of all flanges as shown on Figure 5 will weigh approximately 77,000 Lbs.

This weight represents the sum of the following components:

	Per Vessel Basis	Five (5) Vessels Per Combustor Basis
Vessel	43,235 Lbs.	216,175 Lbs.
Lining	15,660	78,300
Plenum	3,050	15,250
Elements	11,250	56,250
Sand Filter Media	2,250	11,250
Ejectors	1,000	5,000
Pulse Air Piping	555	2,775
TOTAL	77,000 Lbs.	385,000 Lbs.

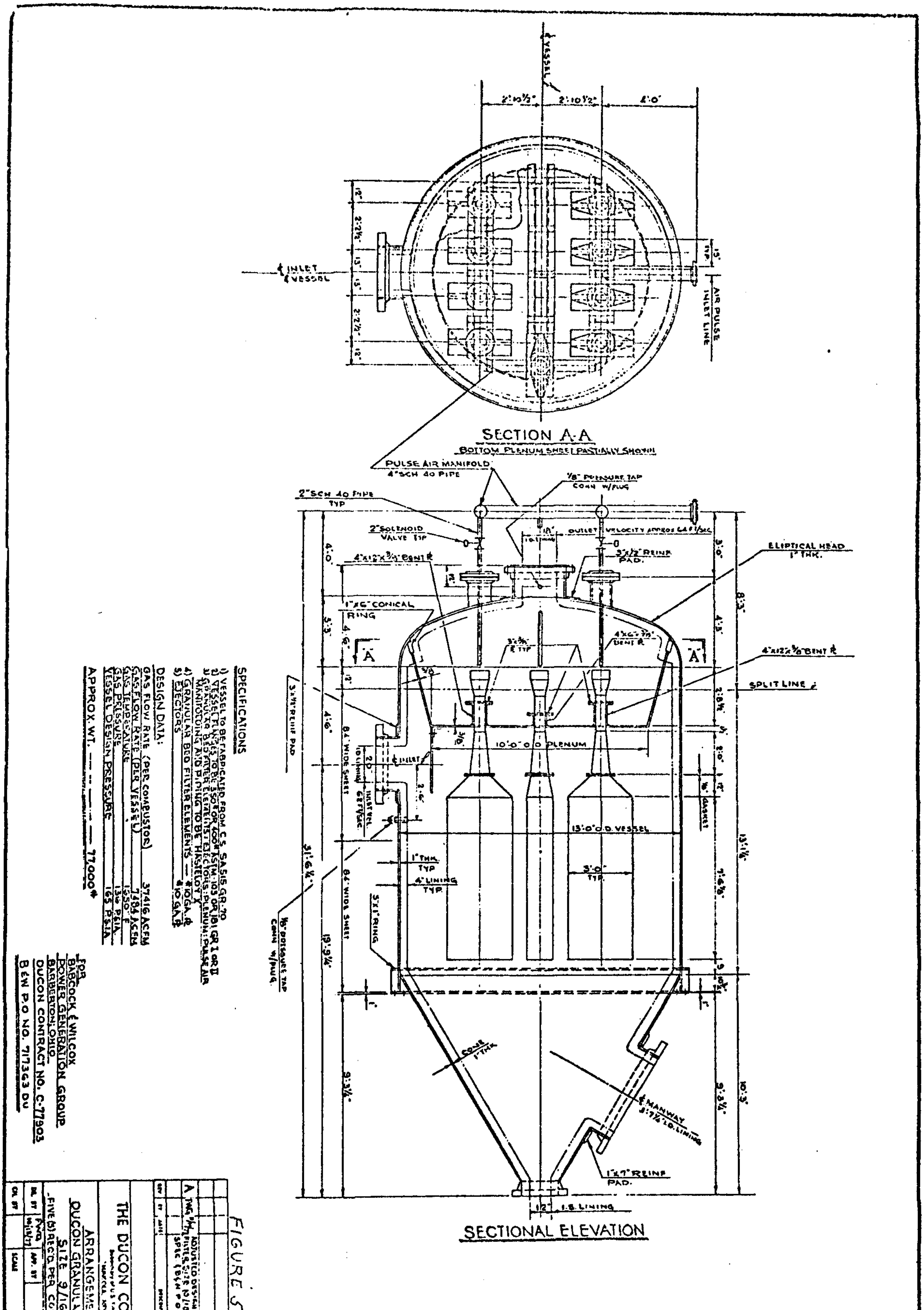


FIG. 5
ARRANGEMENT OF
DUCON GRANULAR BED FILTER
SIZE 9/162

SECTION I

OPERATIONAL DISCUSSION

Number of Elements

The specified design is based on one (1) 25 Ft. vessel containing forty four (44 elements. An alternate design is based on five (5) 13 Ft. filter vessels, each containing nine (9) elements or a total of forty five (45) elements.

Bed Configuration:

Each "sand" bed is 6" x 36" in cross section and approximately 1-1/2" in depth. The beds are arranged in stacks six high so that an element consists of two stacks back to back. Each bed is within a 15" high compartment so that the overall dimensions of an element becomes 36" x 15 and 7' -61/8" in length. The face velocity across 42 elements* is approximately 49 Ft/minute. In test programs during the development of the filter, it has been successfully operated at face velocities of 40 to 100 Ft/Min. The 49 Ft/Min. in this instance is based primarily on pressure drop limitation. The development work has encompasses a vast array of loadings, blowback cycles, face velocities, dust characteristics, sand particle size, compartment configuration, blowback modes and associated variables.

Screen Grid:

The bed support screens consist of 24 - 110 screening on a 13 guage backing strip, all fabricated from Hastelloy "X" metal.

Sand Size:

The development program carried out by Ducon, as well as others, pertaining to filtration through sand beds, converged quite rapidly to an optimum "sand" size in the range of 250 to 600 microns. Larger sizes require deeper beds to achieve reasonable efficiency, but then also require enormous blowback rates in order to fluidize the solids. Smaller "sand" sizes create excessive pressure drop or curtail face velocities to an uneconomical level. The distribution of the sand between 250 and 600 microns is of the conventional log probability or sieved form. It has been simply obtained from conventional bank "sand" by sieving between 30 and 60 mesh U.S. Standard Screens.

*42 elements assumes a design point whereby two elements are continually in a cleaning mode (see later discussion).

Pressure Drop:

Under steady state operation, with a fixed number of elements being blown back simultaneously in each dust collection system, the overall pressure drop across the Granular Bed Filter under the design conditions will be essentially constant. The equilibrium pressure level will depend upon the number of elements being cleaned simultaneously, which in turn establishes how frequently they are cleaned. The number of elements cleaned simultaneously will depend upon the quantity of motive blowback gas available to the ejectors. Figures 6 and 7 show the relationships between the parameters discussed above.

In Figure 6 the motive gas rate vs. Granular Bed Filter ΔP is shown. As previously discussed, two cases were considered during this study: Granular Bed Filterdust system with precleaner and without the precleaner. The equilibrium pressure drop without the precleaner is approximately three (3) times the pressure drop with the precleaner. The precleaner is a one-stage cyclone with a pressure drop of 1.0 psi.

Operational limits for the Granular Bed Filter are indicated by the recommended design range. The lower limit is based on cleaning only one (1) element at a time. The upper design limit is based on cleaning five (5) elements simultaneously, which is approximately 10% of the total elements. The 10% upper limit has been established as the optimum number of elements being cleaned simultaneously without a detrimental effect on Granular Bed Filter efficiency. Operation within these limits would guarantee efficient operation of the Granular Bed Filter.

Figure 6 shows that increasing the motive gas rate increases the number of elements being cleaned simultaneously while decreasing the equilibrium pressure drop. From Figure 7 it can be observed that decreasing the equilibrium ΔP also decreases cleaning cycle period and increases cleaning frequency.

In order to establish the most economical design point, a detailed economic study considering capital costs, compression costs and other related variables would be required. For the purpose of this study, a design point has been considered which meets pressure drop limitations and is within the recommended design range.

The design point chosen would yield a Granular Bed Filter ΔP of 2.28 psi in addition to 1 psi across the cyclone for a total ΔP of 3.28 psi. The cleaning cycle period would be three (3) minutes long, with two elements being cleaned simultaneously for eight (8) seconds. The motive gas requirement is 9072 lb/hr.

Back-Flushing:

The back-flush gas consists of compressed clean gas admitted through ejectors, each of which serves a single element or 12 "sand" beds. The compressed recycle motive gas induces approximately five (5) times its own rate from the clean gas plenum to supply the required volumetric fluidizing flow.

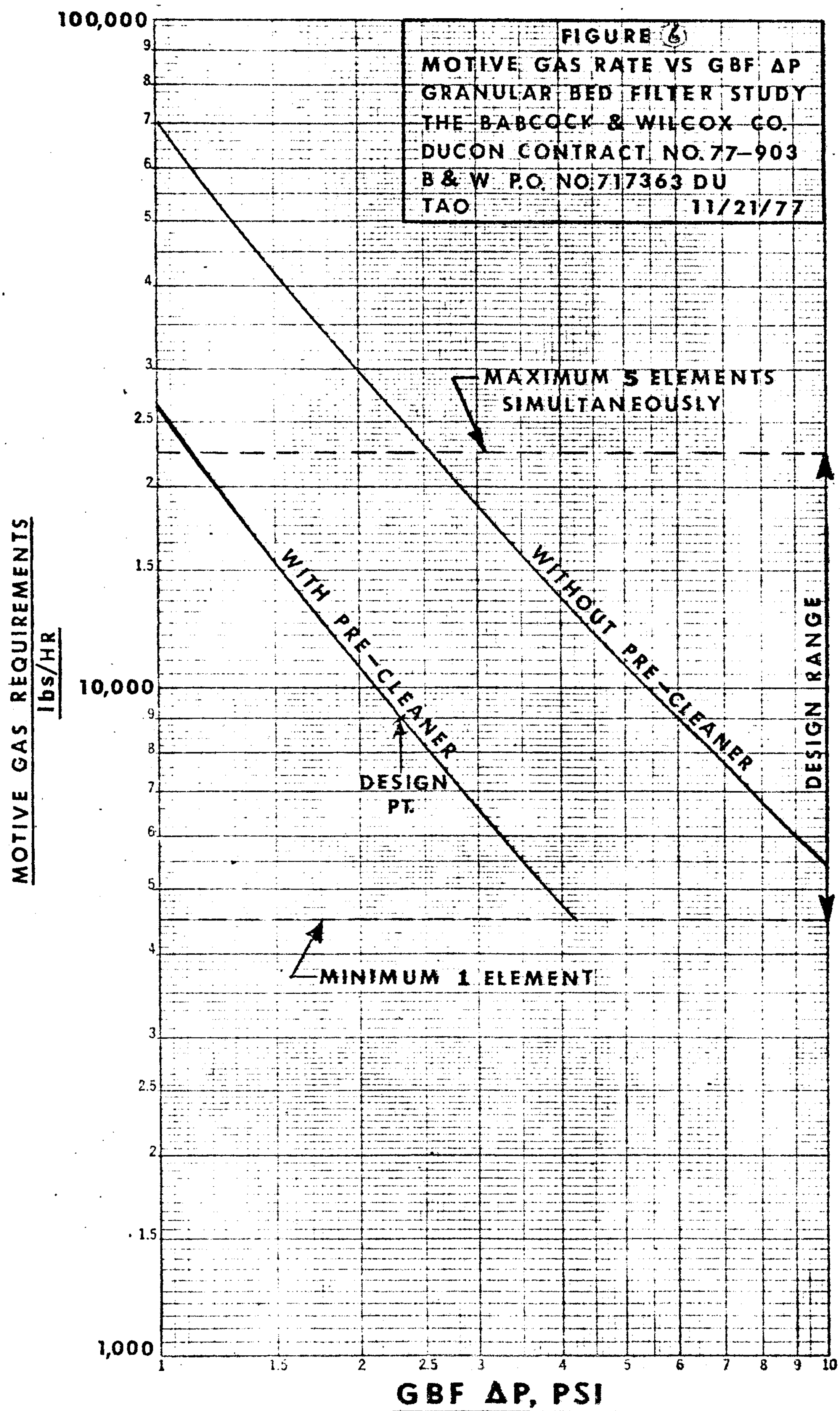


FIGURE 6
MOTIVE GAS RATE VS GBF ΔP
GRANULAR BED FILTER STUDY
THE BABCOCK & WILCOX CO.
DUCON CONTRACT NO. 77-903
B & W P.O. NO 717363 DU
TAO 11/21/77
MAXIMUM 5 ELEMENTS SIMULTANEOUSLY
WITH PRE-CLEANER
WITHOUT PRE-CLEANER
DESIGN PT.
DESIGN RANGE
MINIMUM 1 ELEMENT
MOTIVE GAS REQUIREMENTS lbs/HR
100,000
10,000
1,000
GBF ΔP, PSI

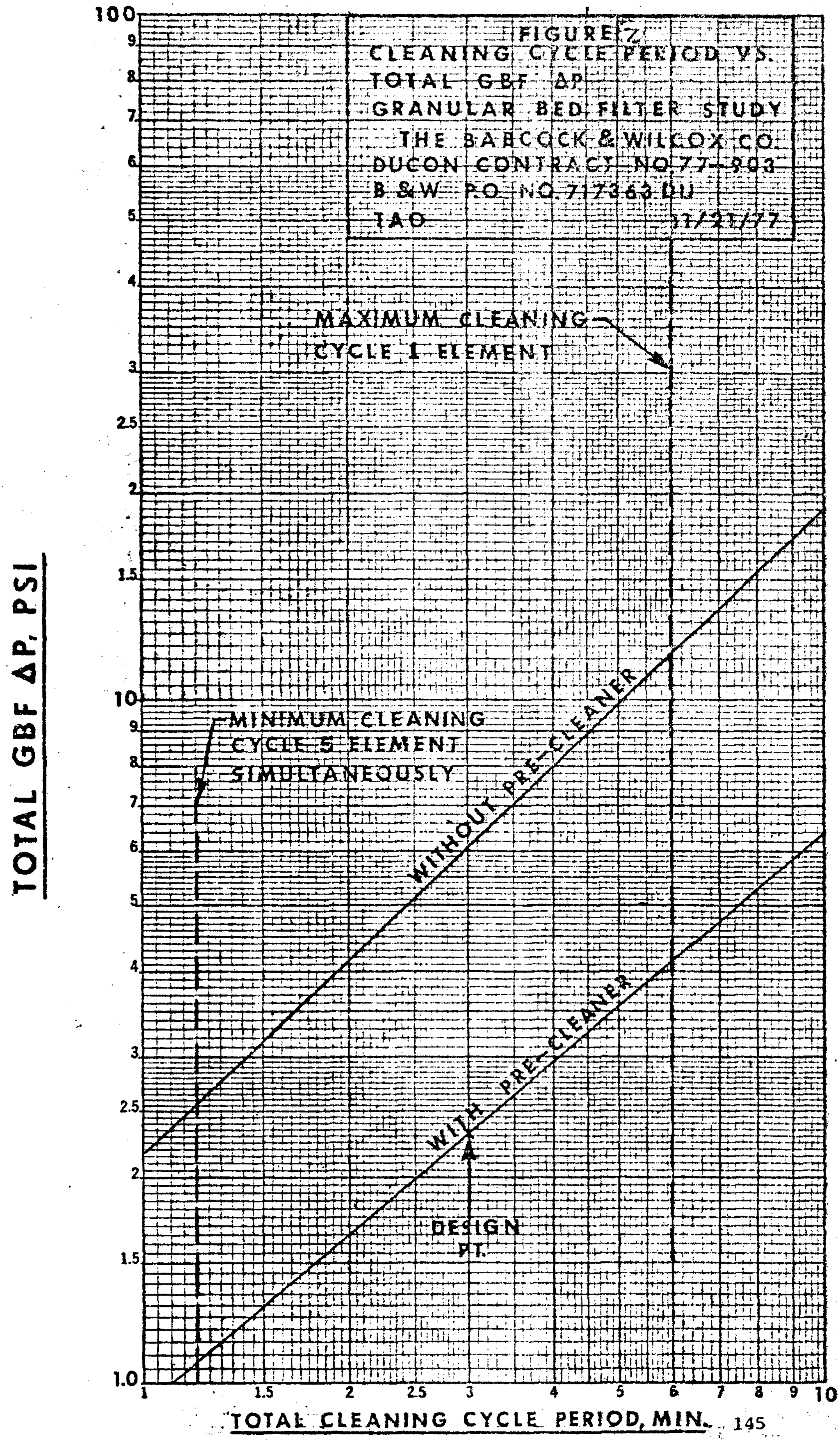


FIGURE 7
CLEANING CYCLE PERIOD VS.
TOTAL GBF ΔP
GRANULAR BED FILTER STUDY
THE BABCOCK & WILCOX CO.
DUCON CONTRACT NO. 77-903
B & W P.O. NO. 717363 DU
TAO 11/21/77
MAXIMUM CLEANING
CYCLE 1 ELEMENT
MINIMUM CLEANING
CYCLE 5 ELEMENT
SIMULTANEOUSLY
WITHOUT PRE-CLEANER
WITH PRE-CLEANER
DESIGN
PT.
TOTAL GBF ΔP, PSI
TOTAL CLEANING CYCLE PERIOD, MIN.
145

The filter development program led to the correlation of ejector performance based on existing information, as well as Ducon tests and Ducon designs. The Ducon Company is therefore well versed in the design requirements specification, sizing, and fabrication of gas - gas ejectors, particularly for Granular Bed Filter applications.

Blowback Gas:

Pressure: The motive gas is to be cooled and compressed to a motive inlet pressure to the ejectors of 314.7 psia.

Temperature: The temperature of the motive gas was in this instance taken as 400°F. This temperature could be altered depending upon compatibility with the selected compressor. Obviously, such change in motive gas temperature and/or pressure would be reflected in minor alterations in the ejector nozzle sizing. In this instance, all designs were based on the assumption of 400°F and 300 psig motive gas conditions.

Flow Rate: For design purposes the motive blowback gas rate was taken as 9,072 lb/hr. However, as indicated by Figures 6 and 7, this could be varied and thereby affect the overall operating pressure drop, as well as the compressor size. Using recycle gas with a cooler and compressor optimizations and/or costs have not been considered in this task.

Solids in Cleaned Gas:

Based upon current laboratory analysis, the filtration efficiency (without blowback) of the Ducon Granular Bed Filter is essentially 100% on all particle sizes, even as low as 0.1 micron. It is anticipated that for this application an overall collection efficiency of 97.5% is achievable, considering the anticipated frequency of blowback as stated above.

Life:

There is no reason to anticipate any curtailed life of the filter elements in relation to their specific design. The filter medium may be selected to withstand particular temperature requirements so that this and the structural steel have operating lives such as those normally accepted for conventional materials of construction. If dust explosions are feared, then rupture discs may be installed as protection against vessel life. The cost of replacement of components is derivable from the cost breakdown.

Replacement Time:

In the event of either an unanticipated weld failure on an element, or by unanticipated breakdown of filter medium, or a desire to alter particle size, the replacement times are estimated below:

Vessel Cooldown	Estimated (by Babcock and Wilcox)
Entry	3 hours
Element Disassembly	3 hours/element
Element Recharge	2 hours/element
Element Reassembly	2 hours/element
Vessel Closure	4 hours

8.2 Detailed Description of Integrated Sorbent Regeneration Processes (Cases I, II, and III)

8.2.1 Description of Cases I and II

8.2.1.1 Regeneration

Spent additive, at approximately 1500°-1600°F, and containing $CaSO_4$, CaO, MgO and ash from the coal and additive used, is discharged from all combustors and cyclones into a common stainless steel air slide line.

Preheated air at 600°F is used to fluidize the spent additive so that it will flow to a rotary vibrating screen which separates any agglomerated particles over 1/8-inch size from the main stream. The fluidizing air, used in the pneumatic transfer and containing fly ash and other very fine particles, is vented back to the Atmospheric Fluidized Bed Cyclones and Electrostatic Precipitators for particulates separation prior to venting to the plant stack.

Oversized material, discarded from the rotary vibrating screen, is first passed through a material separator and then fed to a roto-fin cooler where the material is cooled to 100°F. The cooled material can be either discarded as such or transferred to an ash slurry pond for eventual disposal. Tower water is used for cooling.

For design purposes, 10% of the regenerated additive is taken as losses due to all causes, such as attrition, decrepitation, screening of oversizes, etc.

The sized material from the rotary vibrating screen is now passed over a magnetic separator to remove any magnetic particles in the ash.

From the magnetic separator the material is pneumatically conveyed to a classified material storage via a cyclone separator. The hot conveying air leaving the cyclone is routed to the main power complex electrostatic precipitators for dust separation. A two-hour classified material storage capacity is provided.

The classified spent additive, now at approximately 1300°F, is fed to the spent additive regenerator by a vibrating feeder. The material enters the regenerator at the top of the bed.

Each regenerator has an 11'6" I.D. and an overall height of 41'0". It is operated as a fluidized bed reactor and is designed for the following assumed operating conditions:

Temperature	- 2000°F
Pressure	- 1 - 2 Atm.
Regeneration Efficiency	- 65%
Solids Residence Time	- 5 to 7 Minutes
Fluidizing Velocity	- 4 to 7 feet per second
Regenerated Sorbent Reactivity	72%

The regeneration plant is sized to process the spent sorbents from a 600 MWe PFB-AFB combined cycle power plant that has the following characteristics:

Coal consumption	- 5000 tons/day
Sulfur content in coal	- 4.5%
Ca/S Mole ratio for PFB	- 1.5
Ca/S Mole ratio for AFB	- 4.0
Dolomite Consumption*	- 975 gons/day
Limestone Consumption*	- 1450 tons/day
Total Sorbent Consumption*	- 2425 tons/day

* These consumptions are based on a "once-through" sorbent system.

In order to maintain SO_2 capture in the combustors at the same level as in a corresponding once-through system, circulation rate of regenerated additive must be approximately 5182 tons per day, based on the assumption tabulated above. It is estimated that ten percent of the circulation rate is lost due to screening, attrition, decrepitation, elutriation, etc. Therefore, a minimum of 518 tons per day of makeup sorbent is required, which is 20% of the once-through requirement.

The amount of pulverized coal to be burned in the regenerator is 171 tons per day. This amount is required to preheat the classified spent additive to 2000°F, preheat the combustion air (15% excess) also to 2000°F, and provide the necessary endothermic heat required to convert the $CaSO_4$ to CaO.

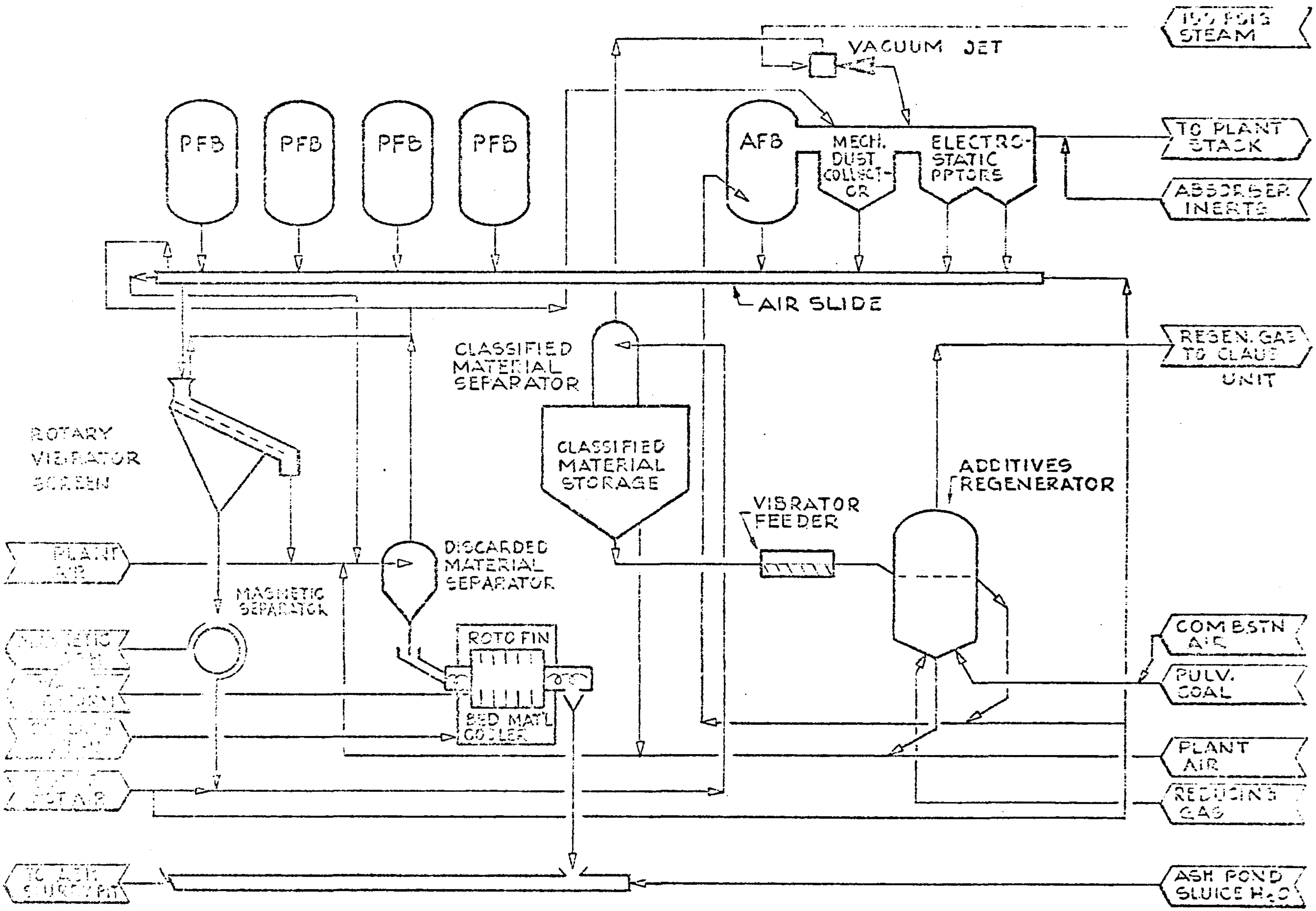
The required one mole of reducing gas for each mole of $CaSO_4$ is supplied at 2000°F and 1-2 Atm. pressure by a Koppers-Totzek type gasifier. This gas enters the regenerator below the grid plate to provide the necessary fluidization medium.

The expanded height of the fluid-bed in the regenerator is approximately 10 feet, and the transport disengaging height is 24 feet.

The regenerated additive containing approximately 40% CaO will leave the bed at 2000°F and be pneumatically transferred to the AFB for reuse. Makeup dolomite is fed only to the PFB's. Figure F-1 depicts the additive regeneration flow diagram for Cases I and II.

The gases leaving the regenerator are also at 2000°F and 1-2 Atm pressure and will have an SO_2 mole concentration of approximately 5.3% for Case I and 7.6% for Case II. The difference is due to the use of air in Case I instead of oxygen.

Due to the high temperature of regeneration, the regenerator itself, the reducing gas line entering the product gas leaving, and the regenerated additive lines leaving the regenerator all are refractory-lined.



CASES I AND II ADDITIVE REGENERATION

FIGURE F1

8.2.1.2 Reducing Gas Generation

Current literature indicates the feasibility of generating the reducing gas required for additive regeneration within the regenerator itself. However, for reasons explained previously in Section 5.5, a Koppers-Totzek (K.T.) Coal Gasifier Package Unit has been provided. It is an entirely self-contained package which uses dried pulverized coal, steam, and oxygen as raw materials, in addition to small amounts of nitrogen.

For Case I, where H_2S is purchased, air is used with sufficient coal to produce a reducing gas mixture containing approximately 36% carbon monoxide (CO) and hydrogen (H_2). The remaining gases are principally nitrogen. Air is used in this case instead of oxygen for economical reasons and to provide a sufficient quantity of fluidizing gases for a reasonably proportioned regenerator. Preliminary calculations indicate that a significant yearly savings in operating costs is effected by using air in lieu of oxygen.

The gas mixture leaving the K-T unit is fed to the bottom of the regenerator at approximately 2000°F and 1-2 Atm pressure.

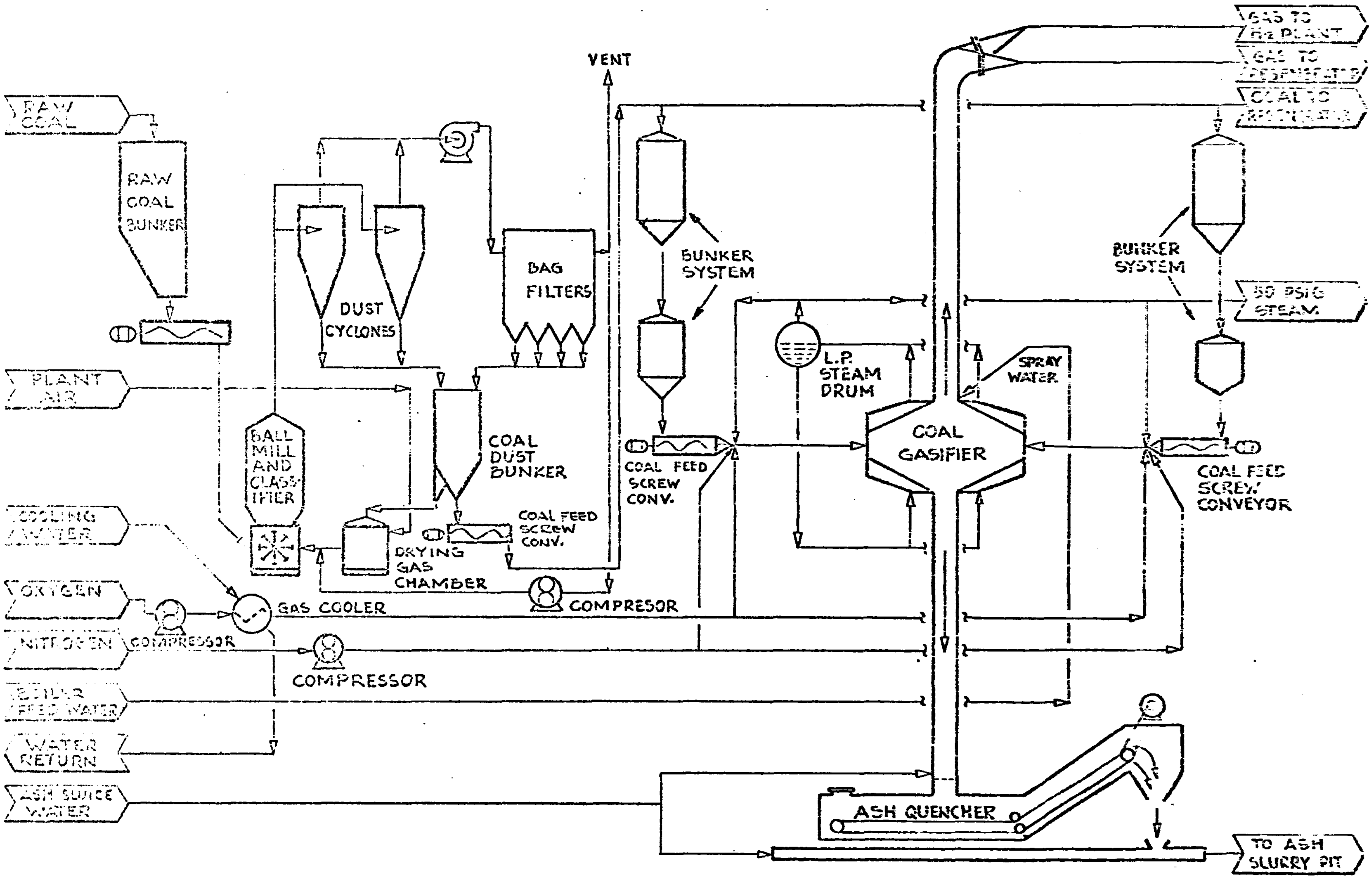
For Case II, where H_2S is manufactured, oxygen for coal combustion and nitrogen for transporting pulverized coal pneumatically within the unit is purchased from "Over-the-Fence". The reducing gas produced contains approximately 84% reducing gas components and only 1% nitrogen. This gas mixture leaves the gasifier at 2500°F and 1-2 Atm pressure.

Since Case II requires more process equipment than Case I, the capital costs have a greater impact on overall annual costs in Case II. Therefore, pure oxygen (O_2) is used for the gasification and combustion of coal to minimize the volume of the product gas and the size of the gasifier and the subsequent equipment (i.e., regenerator, H_2 plant, and H_2S plant). Use of O_2 also ensures higher carbon conversion and reduced retention time.

The equipment in the package K-T plant includes a complete coal preparation section, a gasification and heat recovery section, and a slag removal section, as shown in Figure F-2.

The coal entering the battery limits area is fed to the pulverization system where it is reduced to a size of 70-90% minus 200 mesh and simultaneously dried to a moisture level of approximately 2%. Windswept roller mills are used in a closed system for pulverization. Combustion gas to be used for drying is tempered to 800-900°F to keep the coal particulate temperature at 180°F to keep the coal particulate temperature at 180°F, thereby preventing devolatilization.

The pulverized coal is then transported pneumatically with nitrogen to service bins located above the gasifier. From each service bin, coal is fed to a feedbin. All vent lines from the various bins lead to bag filters to prevent dust emissions. The system of control used ensures a continuous coal feed at a uniform density to the screw



KOPPERS-TOTZEK COAL GASIFIER PACKAGE UNIT

FIGURE: F2

feeders. The function of the feeders is to feed coal to the mixing head at a uniform and controlled rate over a designed range.

At the mixing head, a mixture of steam and oxygen entrains the pulverized coal leaving the metering head and projects the coal particles (through transverse tubes) at velocities above the speed of flame propagation. This is critical to prevent flashback.

The gasifier itself is a steam-jacketed refractory lined carbon steel shell and shaped like two spheroidal cones with the burners spaced 180^{o} apart. The annulus between inner and outer shell is water cooled and connected to a steam separation drum. The low-pressure steam generated in the jacket is used as process steam which enters the gasifier through the mixing heads.

The carbon in the pulverized coal, upon entering the gasifier with the entrained oxygen-steam stream, is exothermically oxidized, thereby producing a high temperature flame zone in the region of 3500^{o}F. The endothermic reactions between carbon and steam substantially reduce the flame temperature to about 2700^{o}F while continuing in the process of oxidizing carbon and producing additional hydrogen.

Ash in the fuel is liquefied in the high temperature flame zone. For most coals, about 50% of the ash flows down the gasifier walls as a molten slag and then is solidified by contact with water in a quench tank situated beneath the gasifier. The granulated ash, somewhat below ¼-inch size, is removed from the quench tank by means of a scraper-conveyor assembly and conveyed to a storage bin for eventual disposal.

The remaining ash leaves the gasifier as a fine fly ash entrained in the exit gas. Since the gasifier temperature is in the region of 2700^{o}F, a problem may be encountered in that the molten ash can cool to solidification and adhere to the gasifier inner walls and heat exchanger surfaces. In Case II, water sprays are utilized to solidify the molten particulates in the high temperature gas prior to entering a waste heat boiler to avoid possible pluggages of the boiler tubes.

The size of the K-T package unit for Case I is approximately 20% the size of the K-T unit for Case II.

8.2.1.3 Sulfur Recovery from Regeneration Off-Gas

A standard, commercially proven, Claus Sulfur Recovery Process Plant has been selected to extract the sulfur values from the additive regenerator off-gas.

The Claus process is commercially used, not as a gas-purification process, but primarily to recover sulfur from acid gas streams containing hydrogen sulfide. The sulfur recovered is of extremely good quality, and this is a source of a valuable basic chemical.

The gases leaving the regenerator are at $2000^{o}F$ and 1-2 Atm pressure, and they contain approximately 10 to 15 weight percent SO_2. These gases must be cooled to approximately $700^{o}F$ before passing through a standard Claus sulfur recovery unit. A waste heat boiler package, including particulate collecting cyclones, is used to cool the gases and generate 400 psig ($1000^{o}F$) superheated steam for use elsewhere in the process.

The basic form of the Claus process used involves mixing two moles of H_2S with one mole of SO_2 at $700^{o}F$ and passing the mixture through several catalytic stages, with condensation of the sulfur vapors formed after each stage. The catalyst used in this process is activated alumina (Al_2O_3) usually shaped into pellets or balls to minimize excessive pressure drop.

The reaction gases, after passing through each converter stage, pass through individual sulfur condensers to drop the temperature from approximately 850^{o} to $325^{o}F$. Low pressure steam is generated in the sulfur condenser in order to cool the reaction gases to obtain maximum sulfur condensation. The reaction gases leaving the first sulfur condenser must be reheated to maintain the temperature of the reaction above the sulfur dew point as it passes through the second converter, since any condensation of sulfur would lead to rapid catalyst deactivation.

For our purposes, two catalytic converters, each with sulfur condensers, are provided to remove all SO_2 in the gases leaving the regenerator. However, the tail gas will contain approximately 2.5 weight percent of H_2S, which is removed in the subsequent Tail-Gas Cleanup System. The Claus unit utilizes 87% of the H_2S supplied to it.

The most important variable in the operation of the Claus sulfur recovery plant is maintaining the ratio of hydrogen sulfide to sulfur dioxide in the reaction gases entering the catalytic converters. Close control of this ratio is necessary.

Deactivation of the catalyst beds by entrained sulfur may also be a problem. Sulfur mist separators after each sulfur condenser have been included to minimize this problem.

The exit gases from the additive regenerator are at 1-2 Atm pressure and contain 10-15% SO_2. The gas must be cooled from $2000^{o}F$ to approximately $700^{o}F$ before passing through the Claus sulfur recovery unit. This is accomplished by passing the gases first through a waste heat boiler which generates 400 psig superheated steam and then through a heat exchanger.

The waste heat boiler package includes particulate collecting cyclones. The separated particulates are recycled back to the spent additive air slide.

Approximately 25,000 pounds per hour of superheated steam are generated here, of which half is used to vaporize and superheat to 700°F approximately 19 tons per hour of purchased liquid H_2S required for Case I.

The gases leaving the waste heat boiler are essentially free of particulates and at the predetermined temperature of about 1300°F and 1-2 Atm pressure. This gas is then used to reheat the vent gas from the Claus unit first sulfur condenser together with the recycled gas from the Claus tail-gas stripper.

The cooled gases leaving the vent gas reheater at 700°F are mixed with the stoichiometric quantity of H_2S; i.e., two moles of H_2S for every mole of SO_2 in the cooled gas. The mixture is compressed to 3 Atm pressure by a multi-stage compressor with interstage cooling, so that the gas temperature at the compressor discharge essentially remains at 700°F.

The gases leaving the compressor pass through the first of two Claus sulfur converters, where approximately 70% of the SO_2 in the incoming gas will be converted to elemental sulfur.

The first converter is packed with activated Al_2O_3 catalyst and is designed to operate at a gas loading of 1800 lb/hr/sq ft. The exit gases leaving this converter are at 800°-900°F and 2-3 Atm pressure. They next pass through the first sulfur condenser. By heat exchange with boiler grade feedwater, the gases are cooled to 325°F to allow the sulfur vapors to be condensed to elemental liquid sulfur. Recovered liquid sulfur is pumped to storage (4-day capacity) for shipment or disposal, and also for use as feed stock for H_2S generation as required for Case II. Approximately 52,000 lb/hr of 50 psig steam is generated in the first sulfur condenser.

As previously mentioned, vent gases leave the first sulfur condenser at 325°F and 2 Atm pressure and pass through a sulfur mist separator so as to minimize catalyst deactivation from this source. The gas then combines with recycled Claus stripper tail-gas for reheating to 700°F prior to passing through a second Claus sulfur converter.

This second converter is also packed with activated Al_2O_3 catalyst and is designed to operate at a gas loading of 1800 lb/hr/sq ft. In this converter the remaining SO_2 in the gas stream is converted to elemental sulfur. The exit gases leaving the second converter are at 750°-850°F and at a pressure of 2 Atm. These gases now pass through a second sulfur condenser and, by heat exchange with boiler grade feedwater, are again condensed to liquid sulfur and transferred to storage. Approximately 35,000 lb/hr of 50 psig steam are generated in the second sulfur condenser.

The tail gases from the second sulfur condenser also pass through a sulfur mist separator to avoid carrying over any sulfur mist to the Claus tail-gas cleanup plant.

For the purposes of this study, it has been assumed that a third Claus converter is not required to meet EPA requirements, since the following Claus tail-gas cleanup plant is able to remove the remaining 2.5% H_2S from the gas stream and recycle it to the Claus converters.

The flow diagram for the Claus sulfur recovery plant is shown in Figure F-3.

8.2.1.4 Claus Tail-Gas Cleanup

With the Claus sulfur recovery plant, a standard Claus tail-gas cleanup plant is included.

It should be noted that the Claus tail-gas clean-up plant would only be required if the recycle of tail-gas to the fluid bed combustors proved technically or economically impractical. While this is considered unlikely, the clean-up system is included here as a conservative measure. The cost of this plant does not significantly affect the final conclusions.

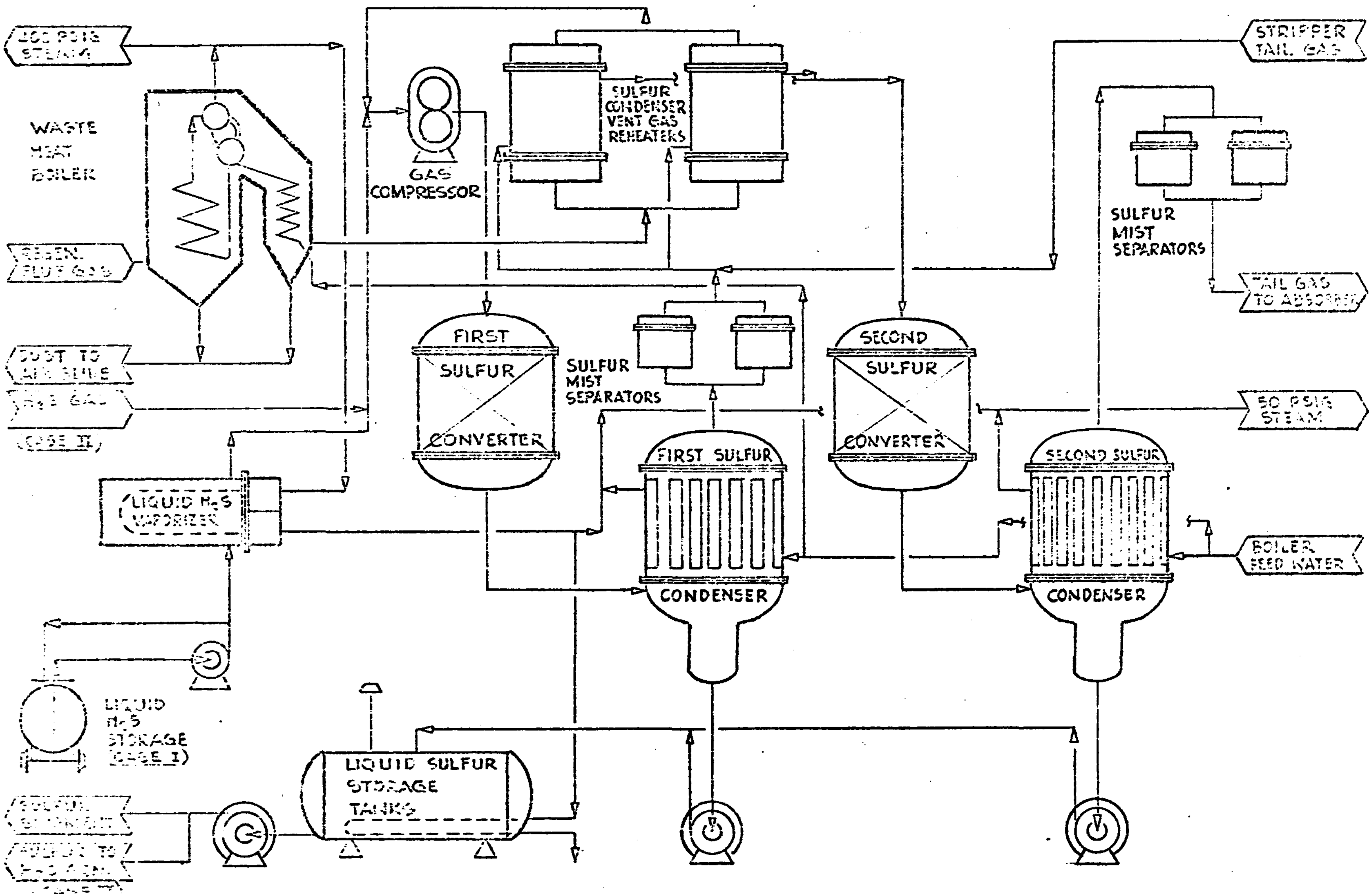
Vent gases leaving the second sulfur condenser at 325°F and 2 Atm pressure contain approximately 2.5 weight percent of H_2S, 22% CO_2, 13% H_2O, with the remainder being essentially N_2. The H_2S must be separated and returned to the Claus sulfur recovery unit.

Normally, monoethanolamine (MEA) solution would be used to absorb both the H_2S and CO_2. But, due to MEA's comparatively greater chemical losses from irreversible side reaction products, its higher vaporization losses (particularly at low pressure), and its greater corrosiveness relative to other amine solutions, a 10% solution of diethanolamine (DEA) in water was selected for use. The absorber operates at a pressure of 7 atmospheres and a temperature of 90°F. In general, DEA solutions are less corrosive than MEA solutions, because the acid gases (H_2S and CO_2) are stripped more easily, and less vigorous reboiling is required. In addition, the decomposition products from side reactions are essentially noncorrosive.

In actual operation, the gases leaving the second sulfur condenser mist separator are first compressed to seven atmospheres by a multi-stage compressor with interstage cooling. After compression, the gas is cooled to 90°F by heat exchange with cooling tower water, and finally passes through a moisture mist separator prior to introduction to the bottom of the H_2S absorber.

The gas flows countercurrent to the 10% DEA solution which absorbs approximately 97% of both the H_2S and CO_2 in the incoming gas.

For design purposes, the absorber packing was assumed to be intalox saddles for high liquid holdup and low pressure drop. The design is based on a gas loading of 1500 lb/hr/ft^2 and a liquid loading of 1000 lb/hr/ft^2.



CLAUS SULFUR RECOVERY PLANT

FIGURE F3

The gases leaving the top of the absorber contain approximately 97 weight % N_2, 0.01% H_2S, with the remainder being H_2O and CO_2. These gases are vented to the power plant stack.

The rich amine solution leaving the bottom of the absorber is essentially 90°F and is preheated to 195°F by heat exchange with hot lean-amine solution discharged from the bottom of the H_2S stripper at 223°F. The preheated rich-amine solution is fed to the top of a packed stripping column operated at approximately 2 psig pressure. Lean amine solution from the heat exchanger is recycled to the top of the absorber at about 88°F.

The H_2S stripper is also packed with intalox saddles for high liquid holdup and low pressure drop. Here, the rich-amine solution at 195°F is stripped of its H_2S and CO_2 content by heated vapors at 250°F passing upward through the intalox saddles.

The hot gases, stripped from the amine solution, pass through a water-cooled condenser and are cooled to 100°F. The non-condensable gases, comprised of H_2S and CO_2, are recycled to the Claus vent gas reheater in the sulfur recovery plant, while the condensed vapors are returned to the top of the stripper as reflux.

Heat for the stripper is supplied by a steam heated reboiler, using 50 psig steam.

The flow diagram for the Claus tail-gas cleanup plant is shown in Figure F-4.

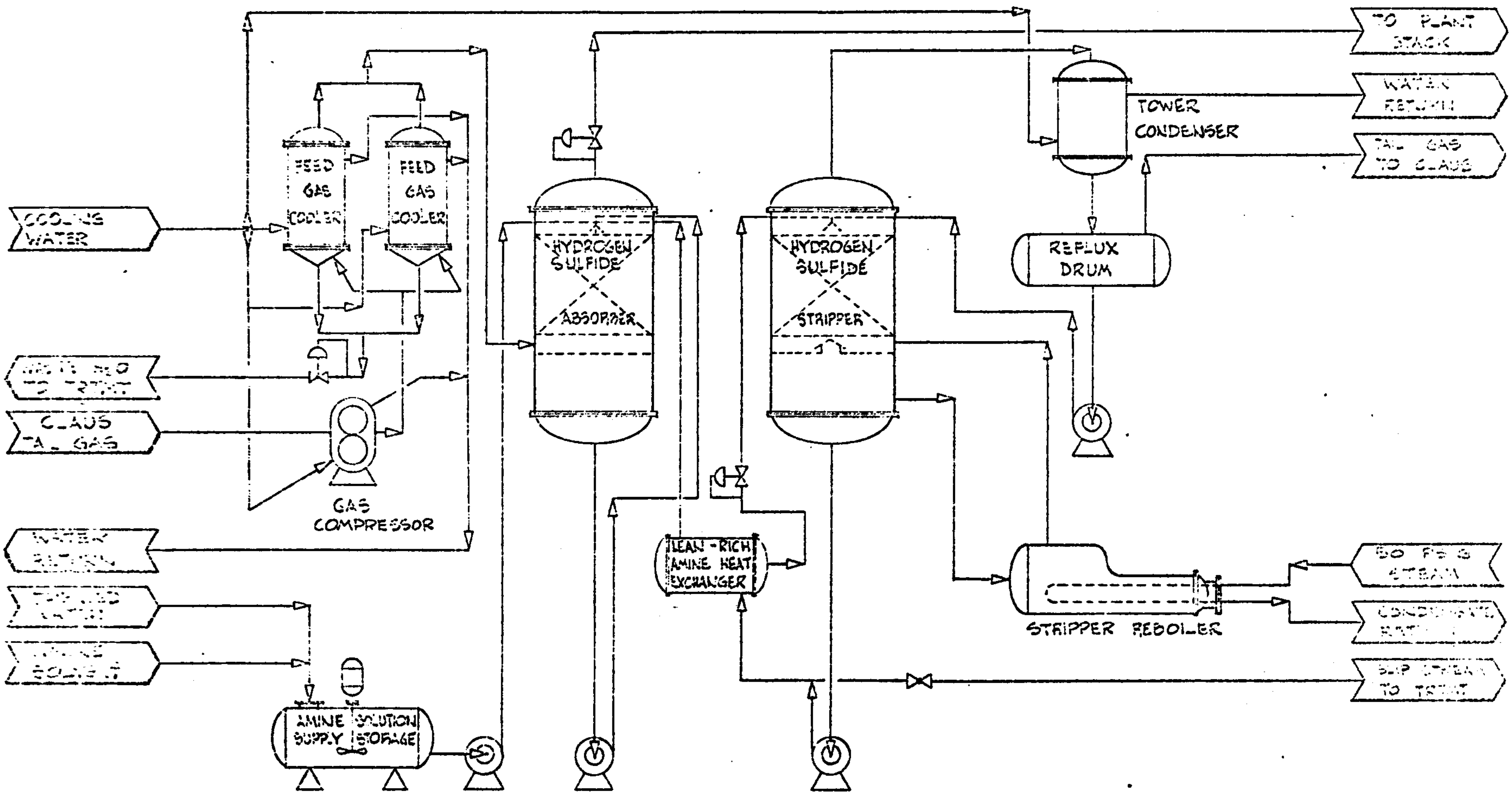
8.2.1.5 Hydrogen Manufacturing Plant

Hydrogen, to be used for the manufacture of hydrogen sulfide, is made by catalytically reacting a water-gas mixture with steam at approximately 900°F. The carbon dioxide impurity is removed by scrubbing the gas with monoethanolamine. For design purposes, water-gas shift conversion is taken at 95% minimum and hydrogen product purity at 98%.

The reducing gas mixture is obtained from an expanded Koppers-Totzek coal gasifier package unit at 2500°F at 1-2 Atm pressure, and with the following approximate composition by weight: 73% CO, 3.5% H_2, 12.5% CO_2, 8.9% H_2O, 1.5% N_2, and the remainder miscellaneous sulfur compounds.

The hot gases at 2500°F from the K-T coal gasifier are first passed through a waste heater boiler package that includes appropriate dust cyclones to remove the major portion of any coal ash particulates contained in the gas. Through heat exchange with boiler grade feedwater, the gases are cooled, thereby generating approximately 28,000 pounds per hour of 400 psig superheated steam.

The cooled gases first pass through a surge gas holder before entering a saturator where they contact hot (190°F) water sprays and are



CLAUS TAIL-GAS CLEANUP PLANT

FIGURE F4

heated to 160°F. The excess water scrubs some of the sulfur compounds from the gas for subsequent waste treatment.

One volume of water-saturated gas is mixed with three volumes of low pressure steam and preheated to 480°F through heat exchange with superheated steam generated by the waste heat boiler.

The preheated gases are passed through a two-stage catalytic converter, where the carbon monoxide (CO) reacts with water vapor to form hydrogen (H_2) and carbon dioxide (CO_2). The first stage contains a high temperature chromium promoted Fe_2O_3 catalyst, and the second stage a low temperature copper-zinc catalyst. This first stage operates at approximately 900°F and the second stage at 500°F, with heat exchangers between stages. The first stage requires 600 pounds of Fe_2O_3 catalyst per 1000 ft^3 of original gas for an 85% conversion. The second stage will require 100 pounds of copper-zinc catalyst per 1000 ft^3 of original gas to complete the conversion of CO.

Two stages are employed because of the exothermic character of the reaction and the decreased conversion at higher temperatures. By using a two-stage procedure, the major part of the conversion takes place with a relatively small amount of catalyst, whereas the balance is brought about at a lower temperature that is conducive to high overall yield. The catalyst has a long life, is sulfur resistant, and converts the small amounts of sulfur compounds into hydrogen sulfide, which is removed with the carbon dioxide.

The gases leave the preheater at 480°F and pass through both heat exchangers before entering the first stage converter at 800°F. Then the gases leave the first converter at 900°F, are cooled by heat exchange to 500°F, and then leave the second stage converter at 700°F prior to final cooling to 400°F. Both heat exchangers here are assumed to be of straight fin-plate construction.

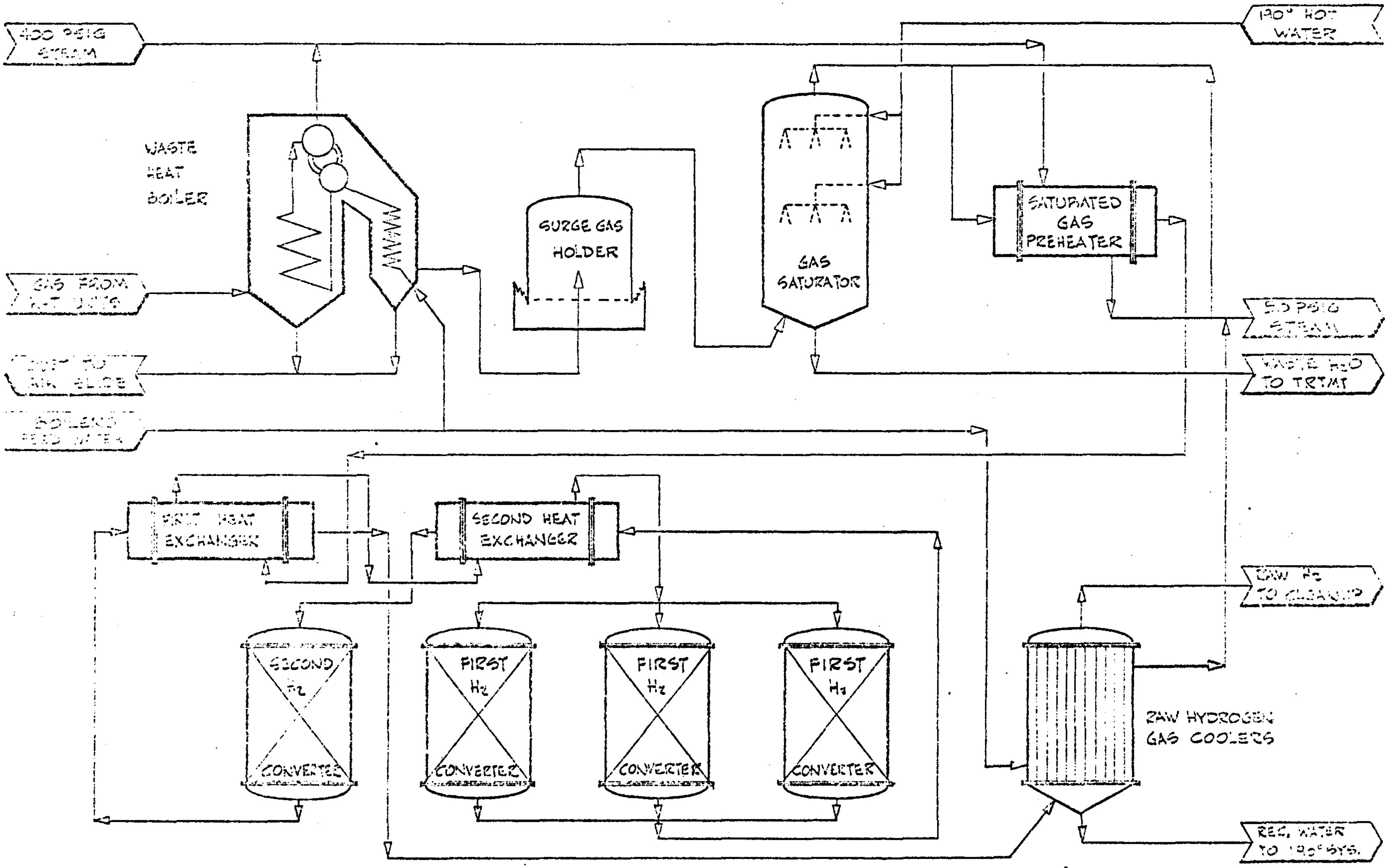
Before purification of the raw hydrogen gas is undertaken, the gas is further cooled to 95°F. In cooling this gas to 95°F, nearly all of the moisture vapors in the gas are condensed and separated for reuse as process water. In the process of cooling by exchanging heat with boiler grade feedwater, approximately 108,000 lb/hr of low pressure steam is produced.

The flow diagram for the raw hydrogen plant is shown in Figure F-5.

8.2.1.6 Hydrogen Purification Plant

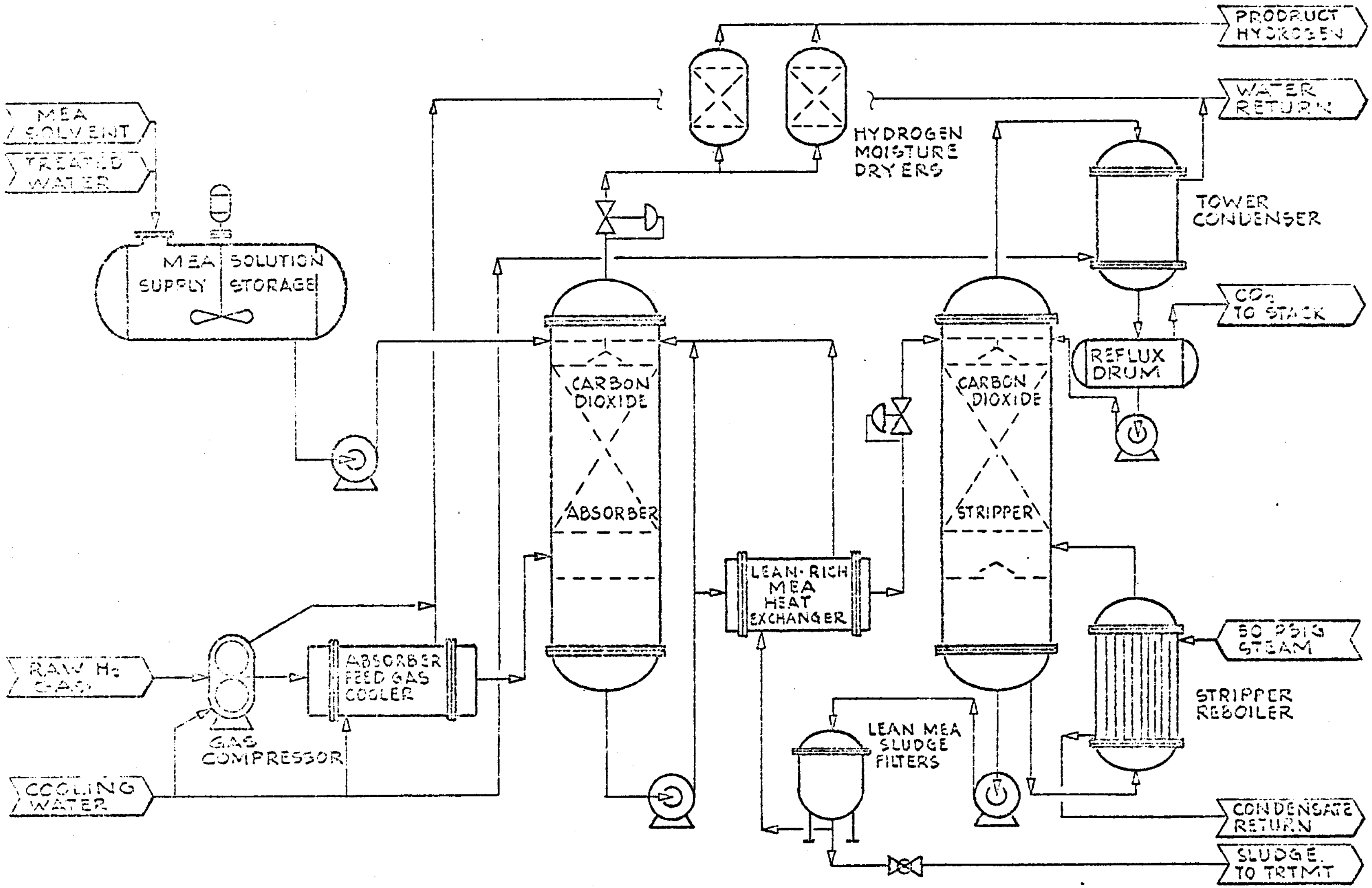
The gases leaving the H_2 manufacturing plant contain hydrogen, carbon, and small amounts of water and nitrogen. A 30% monoethanolamine solution is used to absorb the CO_2 in the gas.

As shown on Figure F-6, the gases leaving the cooler at 95°F are compressed to 17 Atm and cooled again to 95°F. This is done to reduce monoethanolamine losses and to improve absorption efficiency.



RAW HYDROGEN MANUFACTURE PLANT

FIGURE F5



HYDROGEN PURIFICATION

FIGURE F6

The absorber is designed as a packed column using intalox saddles, and it operates with a gas loading of 1500 lb/hr/ft^2 and a liquid loading of 1000 lb/hr/ft^2.

The rich monoethanolamine solution leaving the bottom of the absorber is at 110°F and contains essentially all of the CO_2 and any small amounts of sulfur compounds that may have been part of the raw hydrogen gas.

The unabsorbed gases leaving the top of the absorber are first passed through a pressure reducing valve, and then through alternately operated silica gel dryers to remove the last traces of moisture in the product hydrogen. The composition of this gas is now 99% H_2 and 1% N_2, and it is sent to the hydrogen sulfide generator.

The rich monoethanolamine solution leaving the bottom of the absorber is preheated to 200°F by heat exchange with lean monoethanolamine at 250°F which comes from the bottom of the CO_2 stripper via sludge filters.

This rich monoethanolamine solution at 200°F, after passing through a pressure reducing valve, enters the top of a CO_2 stripping tower. This tower is packed with intalox saddles for high liquid holdup and low pressure drop without flooding in operation. The solution flows countercurrent to steam generated by vaporizing water from some of the solution in a reboiler at the bottom of the tower. The lean or regenerated monoethanolamine, leaving the bottom of the tower at 250°F, is first passed through sludge filters to remove small amounts of byproducts from any sulfur compounds picked up in the absorber and then cooled by heat exchange with the 110°F rich monoethanolamine leaving the bottom of the CO_2 absorber.

Operating conditions for the stripper assume a stripping factor of 0.95 for the CO_2, so that the number of transfer units are 36 in number. This assumes a 99% removal of CO_2 in the stripper.

Gases composed principally of CO_2 and H_2O leave the top of the stripper and are cooled by cooling tower water to 100°F. The condensed water returns to the stripper as reflux, and the separated CO_2 gas (approximately 98% pure) is vented to the power complex stack.

Since some dilution of the regenerated monoethanolamine solution takes place during the hydrogen purification step, some fresh make-up solution must be periodically added to the system from storage.

The material of construction is 304 stainless steel throughout.

8.2.1.7 Hydrogen Sulfide Manufacture

Hydrogen sulfide (H_2S) gas required for the Claus sulfur recovery plant is generated by catalytically reacting hydrogen gas with elemental sulfur vapors. The reaction is exothermic, so that considerable heat is evolved, with the H_2S leaving the generator at approximately 1260°F and slightly above atmospheric pressure.

Dry hydrogen at 100°F is preheated to 875°F by heat exchange with product H_2S at 1260°F. Liquid sulfur pumped from storage is vaporized and superheated to 875°F by 400 psig 1000°F steam in coils. The sulfur vapors and hydrogen gas both at 875°F now pass upward through an alumina (Al_2O_3) catalyst bed and leave the top of the generator at approximately 1260°F.

For design purposes, 98% conversion has been assumed with a space velocity of 1800 volumes of total gases per volume of catalyst per hour.

The H_2S product gas leaving the top of the generator first passes through a sulfur mist separator, then through the incoming hydrogen gas preheater. The cooled H_2S gas leaves the heat exchanger at approximately 725°F and 1-2 Atm pressure for use in the Claus sulfur recovery plant.

The material of construction is high-chrome steel throughout.

The flow diagram for the hydrogen sulfide plant is shown in Figure F-7.

8.2.2 Description of Case III

8.2.2.1 Regeneration

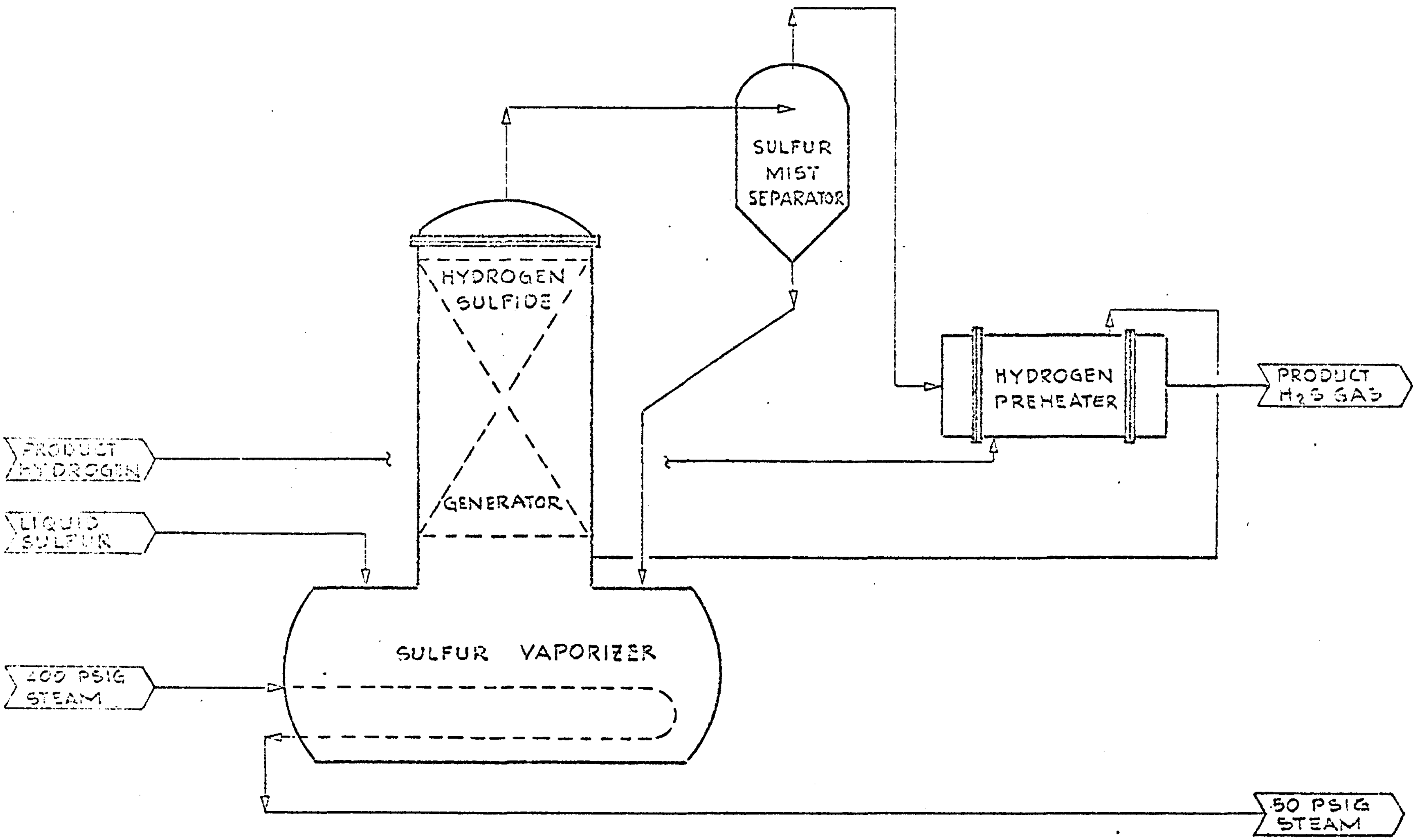
For Case III, the regeneration process in which SO_2 is separated from the spent additive is the same as described for Cases I and II, except that all reducing gas is produced via coal gasification within the additive regenerator vessel itself, rather than in a K-T gasifier. The process flow diagram for Case III has been provided by Argonne National Laboratories for use in this study. As indicated by Table D-2, the conditions are similar to those assumed in the previous cases.

The flow diagram for the spent additive regeneration process is shown in Figure F-8.

The gas leaves the regenerator at 2000°F and 1-2 Atm pressure. Its composition on a weight basis is: 17.0% SO_2, 1.6% CO, 25.8% CO_2, 4.7% H_2O and 50.8% N_2. This gas first passes through a dust cyclone for removal of particulate matter, and then through a heat exchanger to preheat incoming air to the regenerator. Any collected particulates, together with regenerated additive, are recycled back to the atmospheric fluidized bed combustor.

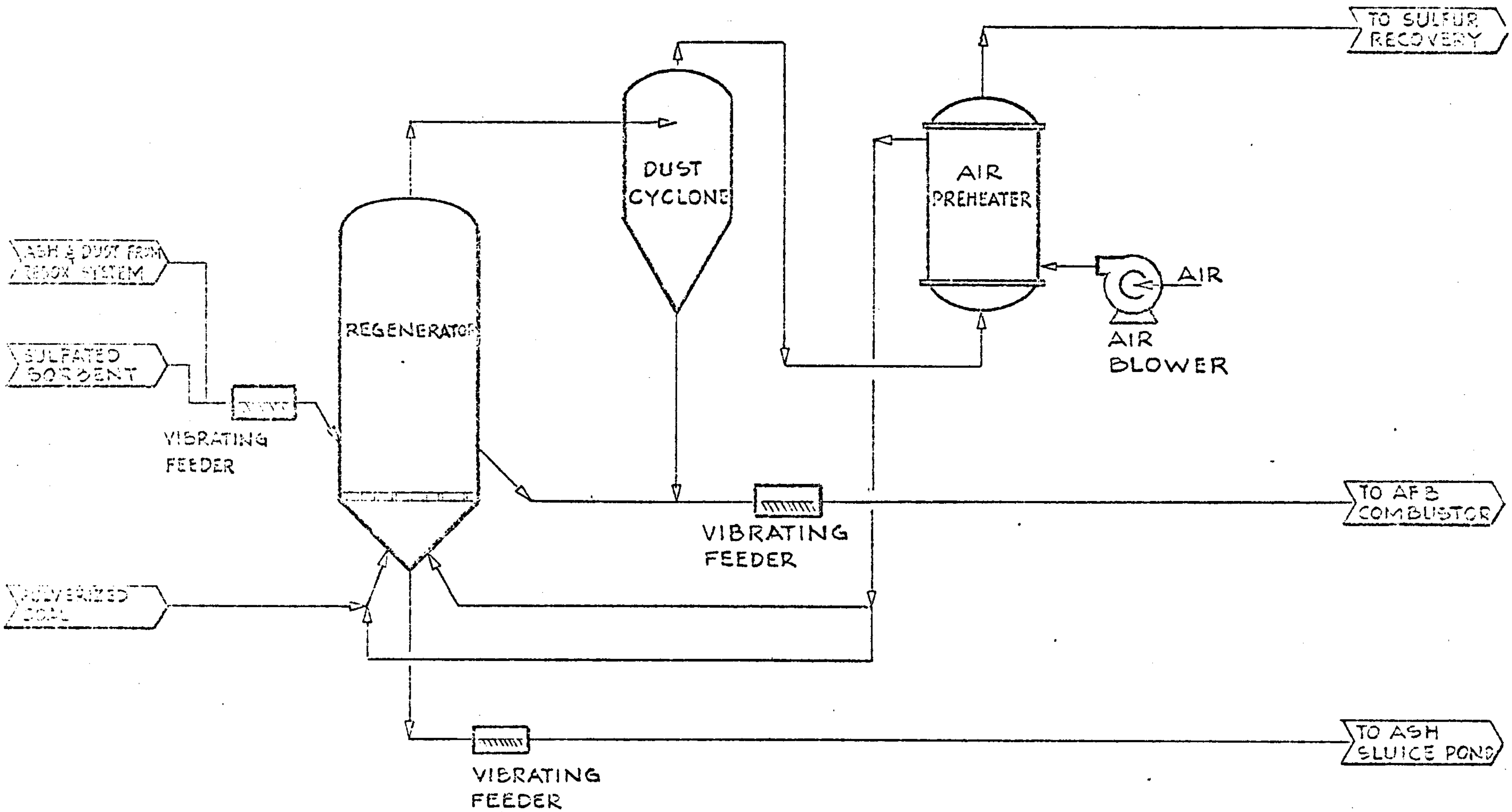
8.2.2.2 RESOX Sulfur Recovery Process

The RESOX sulfur recovery process uses rice size (5/16" X 3/16") crushed coal, preferably anthracite, as a reducing agent to produce gaseous elemental sulfur from the SO_2 component in the gas. No other catalyst is required. After leaving the reactor vessel, the gaseous elemental sulfur is later condensed from the gas stream in a sulfur condenser.



HYDROGEN SULFIDE MANUFACTURE

FIGURE F7



CASE III ADDITIVE REGENERATION

FIGURE F8

Basically, the sulfur dioxide rich gas enters the bottom of the moving bed reactor, together with steam and preheated air, and flows countercurrently with a downward moving bed of crushed anthracite coal. The carbon in the coal reacts with the SO_2 and steam. The two reactions interact synergistically so that both are promoted, and practical rates of SO_2 reduction are obtained at relatively low temperatures. The products of the two reactions are primarily gaseous elemental sulfur, CO_2 and H_2O, with smaller amounts of H_2S, COS, and CS_2 (Refs. 9, 10).

The above is a simplified explanation of reactions taking place, although actually these reactions are quite complex and varied. Foster Wheeler has found (Refs. 9, 10) that when the ratio of H_2O to SO_2 is increased, the percentage of SO_2 conversion is increased. The percentage of SO_2 conversion increases, and the selectivity of the reaction toward H_2S (rather than S) increases with rising reaction temperatures and water concentrations. Between 1220° and 1400°F and with an H_2O to SO_2 ratio of approximately 4, nearly 100% conversion of SO_2 to H_2S has been obtained. When 100% of the SO_2 is converted at moderately lower temperatures, 90% is converted to elemental sulfur, with the remainder going to H_2S, COS, CS_2, etc. The maximum contact time for the SO_2-containing gas with the carbon is fixed at 6 seconds for the tests discussed here. The optimum operating temperature using anthracite coal has been found to be 1200-1475°F.

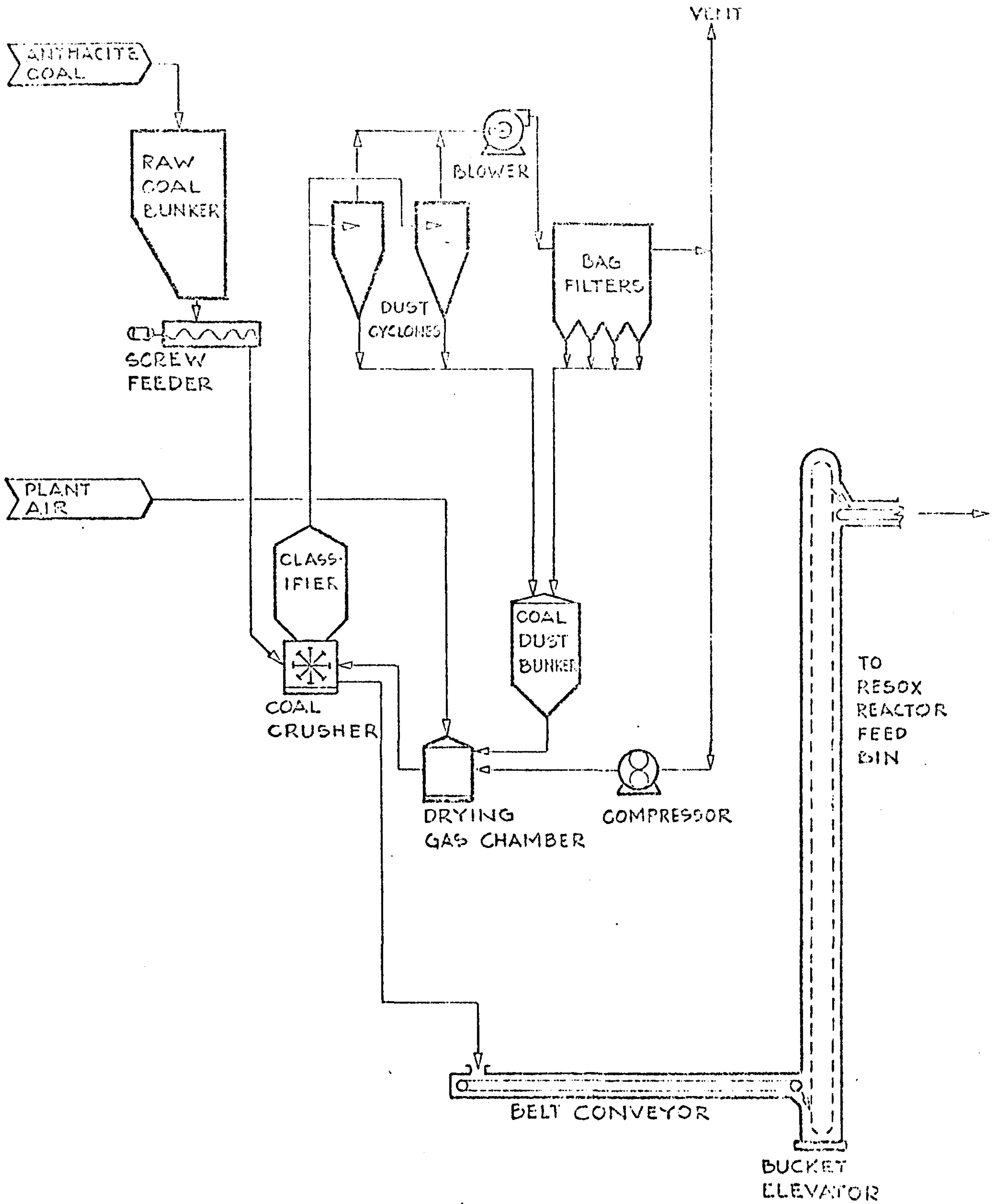
Anthracite coal, for the RESOX sulfur recovery plant, is first crushed to rice size and dried in a coal preparation plant prior to use in the RESOX reactors as shown in Figure F-9.

The coal is then transferred by belt conveyors and bucket elevators to a coal feed bin located over each RESOX reactor. The coal is fed by gravity to each reactor, moving downward slowly and countercurrent to the SO_2 rich gas stream. The coal feed is controlled at a fixed ratio to the amount of SO_2 entering the reactor. The bottom ash (now called "Recoal") is withdrawn at the rate required to maintain a constant bed volume in the reactor. Each reactor is provided with feeders to transfer the ash, which contains a large amount of unused carbon, to a surge bin and thence by conveyors back to the additive regenerator.

The SO_2 rich gas, at a temperature of approximately 1600°F and 1-2 Atm pressure, leaves the air preheater and enters the bottom cone of each reactor (See Figure F-10) through multiple ports around the periphery of the cone, together with the necessary preheated air and steam.

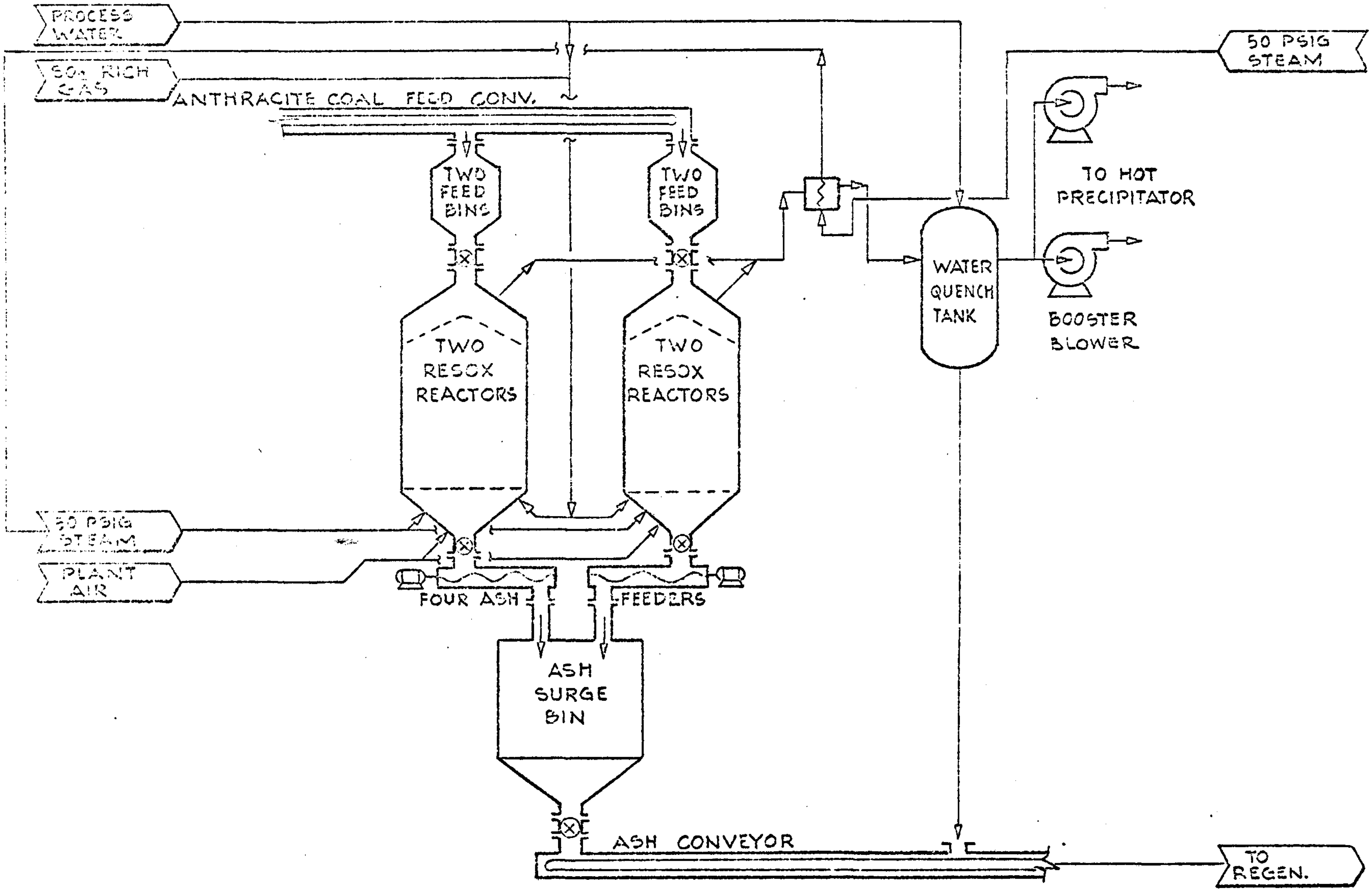
In the RESOX reactors approximately 75% of the SO_2 content in the feed gas is converted to elemental sulfur vapors. The off-gas is at approximately 1200°F and at a pressure just over 1 Atm. On a volume basis, this exit gas contains approximately 4.4% gaseous sulfur vapors, 17.9% CO_2, 30.1% H_2O, 47.1% N_2 and the remainder H_2S, COS, CS_2, etc.

In order to recover the elemental sulfur in as pure a state as possible and to prevent pluggage of the sulfur condenser, the exit gases



RESOX SYSTEM COAL PREPARATION PLANT

FIGURE F 9



RESOX REACTION SYSTEM

FIGURE F10

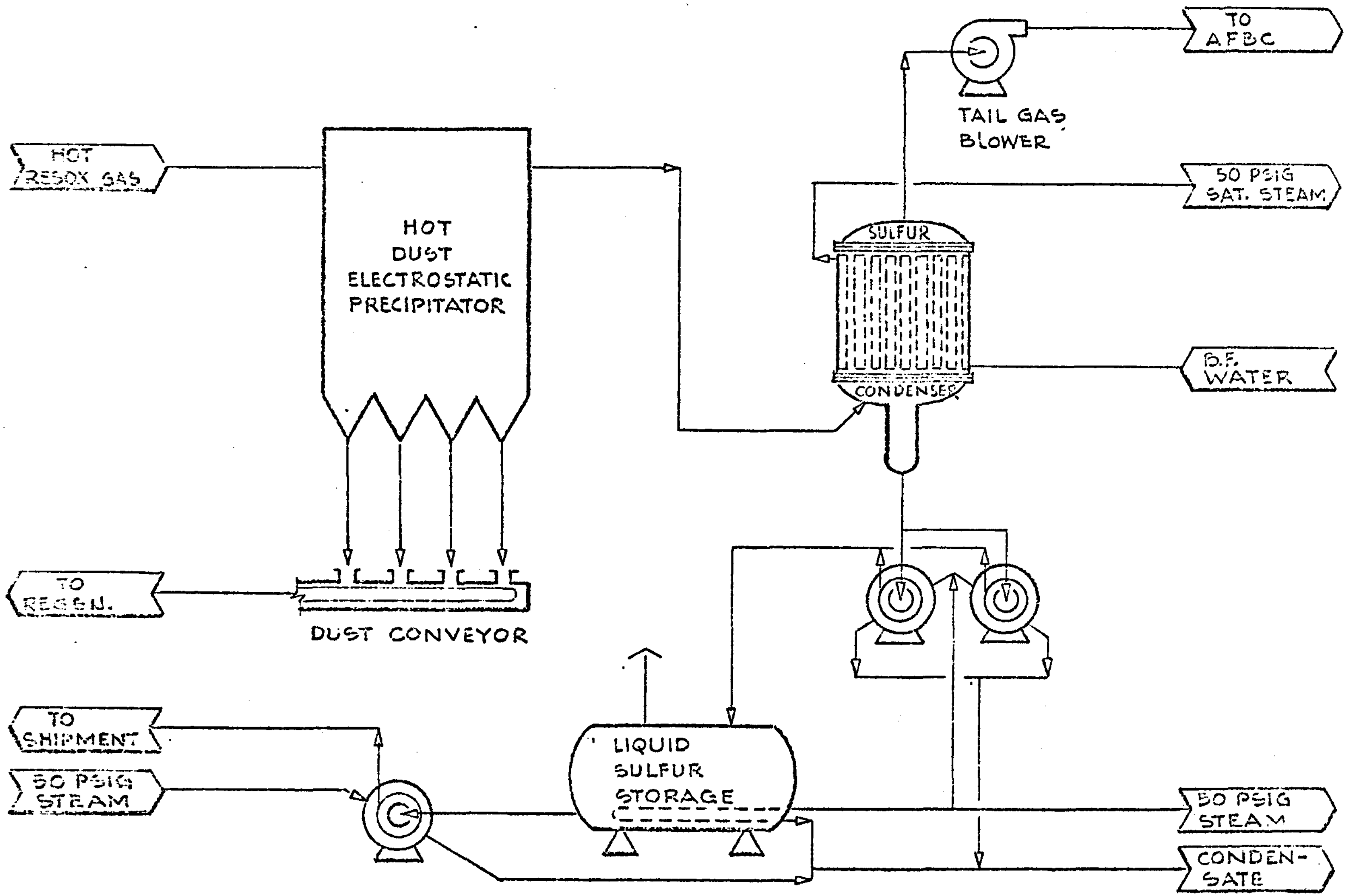
are passed through an electrostatic precipitator to remove all particulate matter. These gases must first be cooled to 700°F, the maximum operating temperature of the electrostatic precipitator. For this purpose, a water quench tank is used wherein a controlled amount of process water is directly sprayed into the incoming gas.

Booster fans then transfer the gases through a hot dust electrostatic precipitator (See Figure F-11) where essentially all the particulates are removed and belt conveyed back to the additive regenerator. The efficiency of the precipitator is 95%, and the gases leaving it contain less than 1 grain per cubic foot.

From the electrostatic precipitator the gases then pass through a sulfur condenser where, through heat exchange with incoming boiler grade feedwater, they are cooled from 700°F to 300°F so as to condense the sulfur vapors to liquid sulfur. The collected liquid sulfur is transferred to storage for sale or disposal. Liquid sulfur storage capacity has been designed for 4 days at full load.

Approximately 40,000 lb/hr of 50 psig steam is generated during sulfur condensation.

Tail gases leaving the sulfur condenser at 300°F are then recycled back to the atmospheric fluidized bed combustor by a booster fan. These gases contain approximately 1% sulfur compounds, 26% CO_2 and 29% H_2O, with the remainder being nitrogen.



RESOX SYSTEM SULFUR SEPERATION

FIGURE F11

FE-2371-36

EVALUATION OF A PRESSURIZED-FLUIDIZED BED COMBUSTION (PFBC) COMBINED CYCLE POWER PLANT CONCEPTUAL DESIGN

Final Report

Reliability and Maintainability Evaluation

with Advanced Technology Assessment

Subtask 1.4

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Date Published - June, 1978

PREPARED FOR THE UNITED STATES

DEPARTMENT OF ENERGY

Under Contract No. EX-76-C-01-2371

ABSTRACT

In June 1976, the U. S. Department of Energy (DOE) awarded a contract to an industry team consisting of Burns and Roe Industrial Services Corp. (PRISC), United Technologies Corp. (UTC), and the Babcock & Wilcox Company (B & W) for an "Evaluation of a Pressurized, Fluidized Bed Combustion (PFBC) Combined Cycle Power Plant Design."

The results of this program indicate that pressurized fluidized-bed combustion systems, operating in a combined-cycle power plant, offer great potential for producing electrical energy from high sulfur coal within environmental constraints and at a cost less than conventional power plants utilizing low sulfor coal or flue gas desulfurization (FGD) equipment.

As a result of various trade-off studies, a 600 MW combined cycle arrangement, incorporating a PFB combustor and supplementary firing of the gas turbine exhaust in an atmospheric fluidized bed (AFB) steam generator, (i.e., a CCFBC plant) has been selected for detailed evaluation.

The overall program consists of the following subtasks:

1.1 Commercial Plant Requirements Definition

1.2 Commercial Plant Design Definition

1.3 System Anslysis and Trade-Off Studies

1.4 Reliability and Maintainability Evaluation With Advanced Technology Assessment

1.5 Environmental Analysis

1.6 Economic Analysis

1.7 Evaluation of Alternate Plant Approaches

1.8 PFB/Gas Turbine/Waste Heat Boiler Cycle Study

1.9 PFB/Gas Turbine/Power Turbine Reheat Cycle Study

This interim report discusses the results of studies performed under Subtask 1.4.

REPORT ON SUBTASK 1.4

RELIABILITY AND MAINTAINABILITY EVALUATION

WITH ADVANCED TECHNOLOGY ASSESSMENT

EVALUATION OF A PRESSURIZED-FLUIDIZED BED COMBUSTION (PFBC) COMBINED CYCLE POWER PLANT

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1.0 SUMMARY

An analysis is presented of powerplant reliability based on the reliabilities of mature plant components. Pressurized Fluidized Bed Combustor (PFBC) and Atmospheric Fluid Bed Combustor (AFBC) availabilities, including coal feeding, cyclones and associated auxiliaries, are set at 88% and 90% respectively. These components are similar in function to fossil fired boilers which have achieved these values in service. Steam turbine and generator reliabilities of 98% and 99% were used which are consistent with current units based on EEI data. Comparable data for base load gas turbines are not available from EEI data and a parametric study was made of the effect of increasing gas turbine reliability from 90% to 99%. The level that should be achieved by mature base load gas turbines is approximately 95%.

The powerplant was calculated to have a plant factor of 75.0 to 80.6% and an Equivalent Forced Outage Rate (EFOR) of 20.4 to 14.5% when the gas turbine reliability was varied from 90 to 99%. At 95% gas turbine reliability, the plant factor was 78.2%.

Development of advanced technology in the areas of pressurizing and distributing the coal, cleaning particulates from the combustion gases and minimizing the effects of corrosion, erosion and deposition in the combustor and turbine is needed to achieve the high reliability necessary for commercial powerplants.

2.0 CONCLUSIONS AND RECOMMENDATIONS

Using mature plant reliability factors consistent with the PFBC/GT/AFPB (90%, 95%, 88% respectively), the plant factor for the commercial powerplant was determined to be 78.2% and the plant factor is the KWHr output divided by the base load rating and period hours.

Varying the gas turbine reliability from 90% to 99% significantly increases powerplant reliability. The plant factor increases from 75.0% to 80.6% and the equivalent forced outage rate is reduced from 20.4% to 14.5%. Comparable changes would occur if the other major plant components were varied over the same range.

The optimum schedule for planned maintenance from a powerplant reliability viewpoint is to do all planned maintenance during the same three-week period. This is due to the dominant effect of the coal/AFBC planned maintenance of three weeks. If some of the possibly uneconomic part load operating modes are eliminated, the plant factor would be calculated to be 1 to 2% lower.

3.0 MAINTAINABILITY EVALUATION WITH ADVANCED TECHNOLOGY ASSESSMENT

3.1 INTRODUCTION

In this section, the specific effects of incorporating maintainability features into the design of the gas turbine, PFBC and balance of plant are discussed. In this program it has been assumed that for a commercial powerplant to be viable the advanced technology for PFB's, gas turbines and advanced balance of plant equipment will all have been developed to have complete specifications, performance and cost from the equipment manufactures. It is then possible to evaluate advanced systems in the light of today's experience on similar equipment.

The approach to minimize maintainance downtime and cost is 1) to use the best available practice for currently available equipment, 2) to recognize that new equipment such as that introduced to meet environmental needs must be integrated into the overall maintenance concept, 3) to incorporate maintainability features into the new major equipment, ie., the gas turbine, PFB, cleanup systems, etc., early in the design process.

3.2 GAS TURBINE MAINTAINABILITY

The Subtask 1.2 selected cycle imposes modest demands on the gas turbine as far as cycle parameters are concerned. The major parameters are a 10:1 overall pressure ratio and a turbine inlet temperature of 1600°F. These cycle conditions are typical of industrial gas turbines that have been in operation for 10 or more years and aircraft engines that were introduced more than 25 years ago.

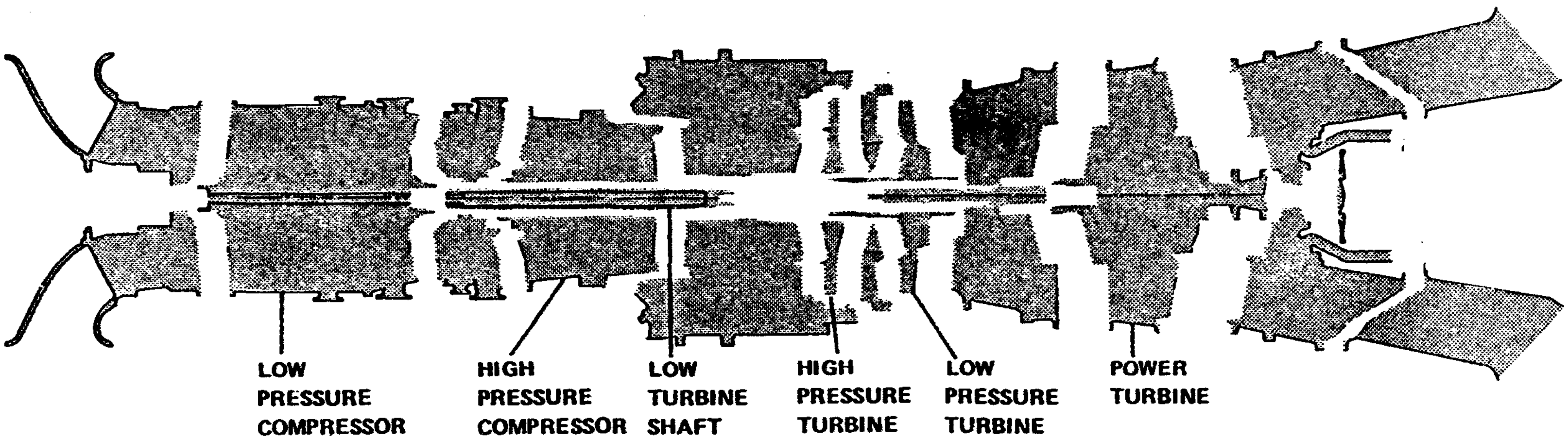
The advanced technology demand placed on the turbine, therefore, is limited to the corrosion, erosion and deposition in the turbine. A substantial amount of R&D must be done before the best economic compromise between cost of the hot gas cleanup system and turbine maintenance cost is determined. While the cycle demands can be met with existing gas turbines it is expected that advanced turbines will be designed with mechanical and aerodynamic features which will minimize corrosion, erosion and deposition problems.

It can be shown that the best gas turbine economics for the commercial powerplant will occur when the turbine blading is refurbished more often than compressor blading or structural sections of the engine. It can also be shown that wear or damage often affects one component without influencing other components. Modular construction of gas turbines for maintainability has, therefore, become a design philosophy at UTC. All recent engine designs, including the FT50 industrial engine as well as the later models of the FT4, use this principle. Figure 1 shows a typica modular construction as applied to the FT50 along with estimated time to remove and replace. These times, which represent maintenance hours to recover from a forced outage and return to service, are all less than two days. The outage times to replace aircraft derivative gas generators are slightly less than FT50 major module replacement, while present industrial gas turbines other than the FT50 can take six weeks or more to return to service.

The specifics of applying the modular maintenance approach for the PFB gas turbine will be determined in a future design/development program. The ducting to and from the PFB must be considered and will affect the location of joints, direction of removal and capacity of maintenance equipment.

MODULAR CONSTRUCTION

FIGURE 1



MODULE	WEIGHT (LB)	TIME TO REPLACE (HOURS)
LOW COMPRESSOR	31,600	16
HIGH COMPRESSOR	19,200	34
HIGH TURBINE	4,000	26
LOW TURBINE	5,000	22
POWER TURBINE	35,000	34

ALL DIMENSIONS ARE UNDER 11 FEET

ALL WEIGHTS ARE UNDER 43,000 LBS

Pratt & Whitney Aircraft

DIVISION OF UNITED AIRCRAFT CORPORATION

J4655-25
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3.3 COAL HANDLING, PFBC, CLEANUP SYSTEM AND AFBC MAINTAINABILITY

All of this equipment will be of advanced design. Numerous technical advances will take place during the R&D programs which will be conducted during the next few years. Technically, however, present day equipment is performing the functions that must be accomplished in the advanced equipment. For example, in pressurized pumping and distribution of coal, B&W has equipment operating at moderate pressure levels and is experienced in manufacturing boilers of all sizes. Cyclones are available commercially which operate at high temperatures. Experience with present equipment will be used to meet specific future performance levels and maintainability objectives with updating as required.

In the design of this equipment, it can be anticipated that critical components that are subject to wear or failure will be identified. Access to these components will be part of the design criteria as well as providing removal space, laydown areas and maintenance equipment such as fixtures and cranes.

Tube maintenance in the PFB will follow boiler practice where extra tubes are included in the design phase so that a number of tubes can be plugged or removed when leaks are detected. Access to the tube bundles for inspection and replacement is provided for in the original design phase. In the B&W PFBC design positive air pressure blows the bed material outward from the tube permitting operation with some leakage.

Each of the cleanup systems investigated (Aerodyne, Tan - Jet, and Ducon) included an assessment of expected maintenance requirements (See Task 5 report). This assessment favored the Aerodyne type of equipment with the Ducon system requiring the most maintenance.

3.4 BALANCE OF PLANT MAINTAINABILITY

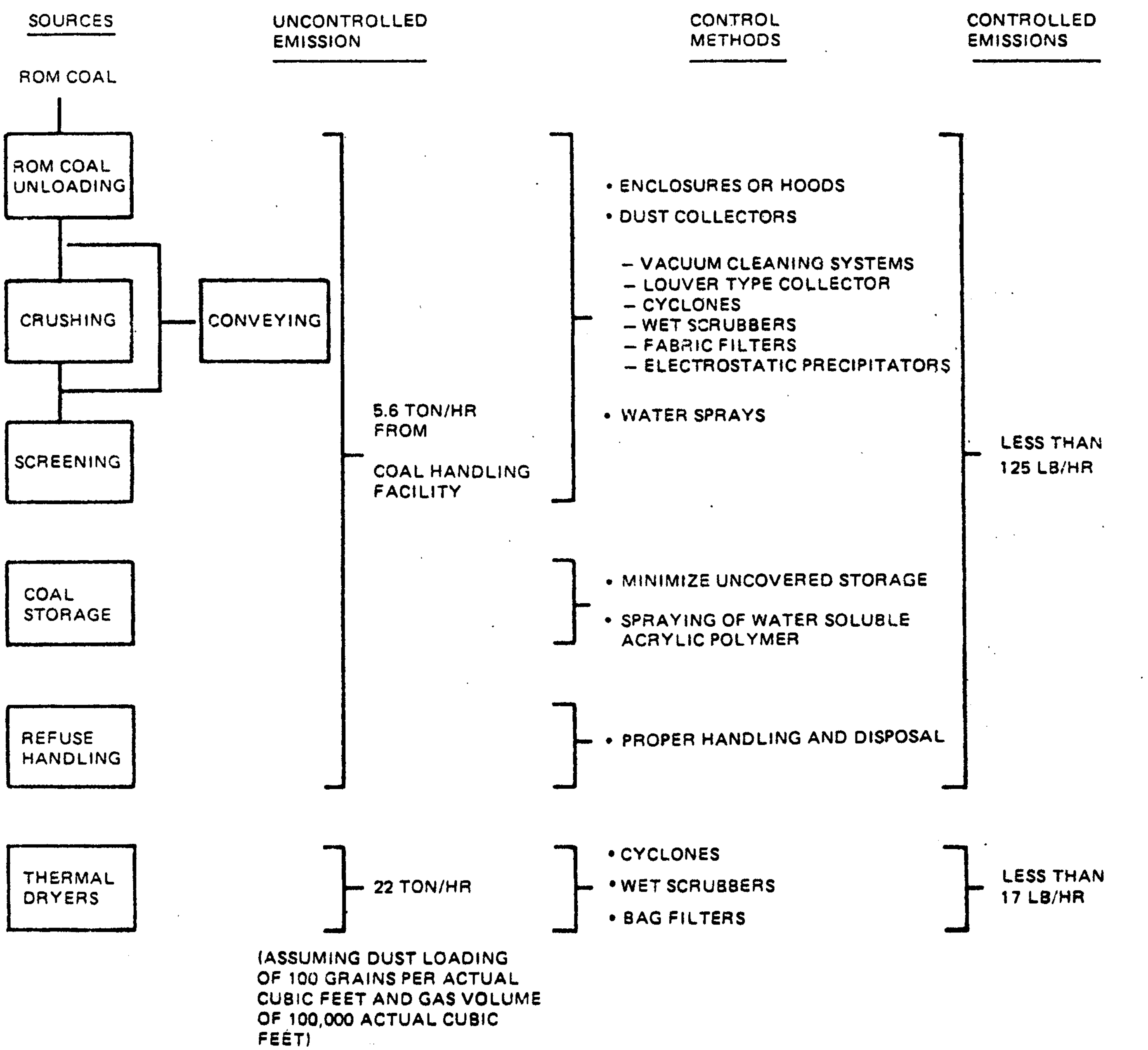
The maintainability of the plant is determined to a large measure by the experience of the architect and engineering firm responsible for the design. Equipment specifications from manufacturers and drawings showing access points for performing maintenance functions are used in the plant layout to ensure that maintenance and repairs and replacement of components can be accomplished.

The addition of environmental equipment increases the inherent need for maintenance. A typical chart showing the control methods required due to coal preparation and handling is shown in Figure 2. It can be seen that enclosures, dust collectors and other equipment must be used to control particulate emissions. Even the passive equipment affects maintenance because it often has to be removed before the active motors, drives, etc. can be reached for repair.

To maintain overall plant reliability it is necessary to reduce the average maintenance time per component. It may also prove beneficial to provide more redundancy in certain systems to reduce forced outages so that as much maintenance as possible can be accomplished in the scheduled maintenance periods.

FIGURE 2

PARTICULATE EMISSIONS FROM COAL PREPARATION AND HANDLING



4.0 RELIABILITY DEFINITIONS AND DATA

4.1 INTRODUCTION

This section presents the definitions used in Task 4 for the reliability evaluation and discusses the data used for these calculations.

The overall goals for powerplant reliability must be expressed in specific terms in order for design requirements to be quantified and results measured. Reliability, availability and durability must be predictable using equipment and component data. The predicted quantities must be combined to evaluate overall reliability of the powerplant. The following sections discuss methods of prediction and how these predictions relate to overall powerplant goals.

4.2 RELIABILITY MEASUREMENT

Reliability is a broadly used term and reliability definitions are varied to suit particular applications. Electric utility generating reliability measurements focus on forced outage rate, availability, and starting reliability. These results are used to determine long term generation capacity needs, short term spinning reserve requirements, and ultimately--the cost of electirity. Forced outage rate and availability are closely related to Mean Time Between Failure (MTBF) and Mean Time to Repair (MTTR) which are widely used by equipment manufacturers. Definitions of these reliability terms and the relationships between them are given in the following paragraphs and tables.

Electric utility reliability terms are defined [1] [2] [3] from a series of basic powerplant operating parameters which are summarized in Table 1. Period hours (PH), service hours (SH), forced outage hours (FOH), reserve shutdown hours (RSH), available hours (AH) and the number of forced outages (N) are reported to EEI for each powerplant. EEI publishes these data in publicly available summary reports and proprietary reports to each participating utility on its plants and each manufacturer on its equipment. [1] [2] [3]

The electric utility terms defined from the operating terms in Table 1 are summarized in Table 2, together with the definitions of two reliability terms generally used by equipment manufacturers. Availability, reliability, forced outage rate, starting reliability, mean time to repair (MTTR) and mean time between failure (MT BF) are all easily calculated from the basic terms in Table 1 and data on starting successes and failures. These terms are used in statistical reliability analyses applicable to the powerplant.

4.3 AVAILABILITY

Availability is defined as the amount of time a powerplant is available for power production, divided by the period hours. Unavailable hours the sum of forced outage hours (FOH) and scheduled outage hours (SOH). The availability is a widely used measure of reliability and generally has a real meaning for base load units. For peaking units, however, information reported to EEI may not represent the actual equipment availability. Users may elect to postpone maintenance or repair actions, until the equipment is expected to be needed. The causes for these "discretionary outages" should be isolated where calculating availability. As a result of the EEI reporting method both pumped storage hydroelectric and gas turbine powerplants are assumed to have availabilities of 90 percent and above by utility planners [4] in contrast to the EEI data for FOR of 15-30 percent [1] [2] [3]. In addition, the repair of a small peaking unit may be postponed while a base load unit is undergoing maintenance adding to the discretionary outage rate. Since availability includes forced and scheduled outages and the time a plant is on reserve shutdown

TABLE 1

EEI DEFINITIONS OF BASIC POWER PLANT OPERATIONS TERMS

POWER PLANT	DEFINITION
Period Hours (PH)	The Clock Hours in the Period under Consideration. (Generally 8760 Hours/ Year.) PH = SH + FOH + SOH + RSH.
Service Hours (SH)	The Total Number of Hours the Unit was Actually Operated with Breakers Closed to Station Bus.
Forced Outage Hours (FOH)	The Time in Hours During Which a Unit or Major Equipment is Unavailable due to a Forced Outage. (A Forced Outage is the Occurrence of a Component Failure or other condition which requires that the unit be removed from service immediately or up to and including the very next weekend).
Scheduled Outage Hours (SOH)	The time in hours a unit or major equipment is unavailable due to a scheduled outage. (A scheduled outage is the occurrence of a component failure or other condition which requires that the unit be removed from service but not before the very next weekend. This includes scheduled maintenance).
Reserve Shutdown Hours (RSH)	Reserve shutdown duration in hours. (Reserve shutdown is the removal of a unit from service for economy or similar reasons. This status continues as long as the unit is out but available for operation.
Available Hours (AH)	The time in hours during which a unit or major equipment is available. AH = SH + RSH.
Number of Forced Outages (N)	The number of forced outages during the period hours under consideration.

TABLE 2 SUMMARY OF ELECTRIC UTILITY GENERATING RELIABILITY TERMS AND THEIR INTERRELATIONSHIPS

RELIABILITY TERM	DEFINITION
AVAILABILITY	$A = \frac{PH-(FOH + SOH)}{PH} \times 100 = \frac{SH + RSH}{PH} \times 100 = \frac{AH}{PH} \times 100$ $= \frac{PH-N\ (MTTR)-SOH}{PH} \times 100 = \frac{N \times MTBF + RSH}{PH} X \times 100$
RELIABILITY	$R = \frac{PH-SOH-FOH}{PH - SOH} \times 100 = \frac{PH-SOH-N\ (MTTR)}{PH-SOH} \times 100$
FORCED OUTAGE RATE	$FOR = \frac{FOH}{SH + FOH} \times 100 = \frac{MTTR}{MTBF + MTTR} \times 100$
STARTING RELIABILITY	$SR = \frac{\text{SUCCESSFUL STARTS}}{\text{ATTEMPTED STARTS}}$
MEAN TIME TO REPAIR	$MTTR = \frac{FOH}{N}$ WHERE N = NO. OF OUTAGES
MEAN TIME BEFORE FAILURE	$MTBF = \frac{SH}{N}$

(refer to Table 2), it gives a measure of the overall time a plant could be run at base load. The value is on the optimistic side since, if the plant were run through all the reserve shutdown hours, a forced outage could occur.

The availability of a new powerplant is classically viewed as a "bathtub function" (Figure 3). The plant is seen as going through a debugging or learning stage (Region 1 of Figure 3). during which its availability increases. It then bottoms out to a steady, normal operating or useful life with a constant availability, Region 2. The plant then enters a "wear out" time when its availability decreases, represented by Region 3. "Wear out" refers to the condition where a part has reached its ultimate durability life and either fails by stress rupture or permanent set as a result of excess creep, or, in the cases of material erosion/corrosion, loses an excessive quantity of surface material to the extent that replacement is essential. The reliability measures defined in the preceding section are for the normal operating life of the plant. If the learning and wear out regions could be well defined, the electric utility system reliability could take these into account. Typically , however, these periods are quite irregular and decend (and ascend) in an irregular staircase fashion rather than the smooth curve shown.

In any case, Availability and Reliability can be related to MTTR and MTBF as shown in Table 2 and the Table 2 definitions were used in making the calculations reported in Sections 6.0 and 7.0

4.4 STARTING RELIABILITY

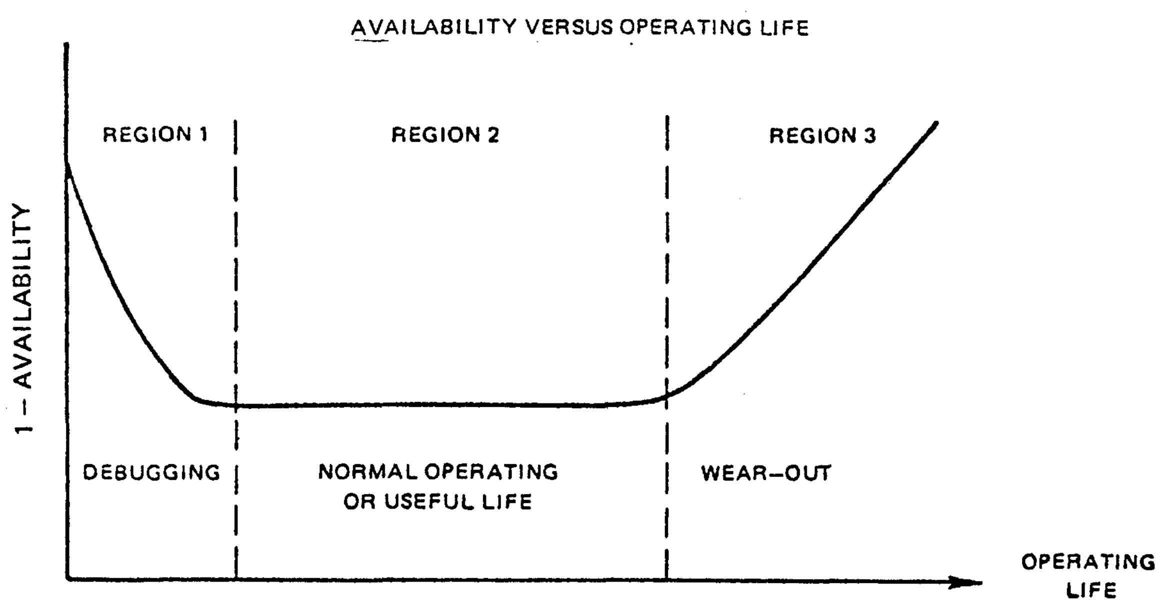
The starting reliability of large fossil units is currently reported to be 82.8 percent. While this rate should be improved, the long run times between starts means that availability is relatively unaffected by the starting reliability. The starting reliability of jet engine powerplants is currently 85.3 percent but the order of magnitude greater number of starts required of peaking units reflects the very different demands of peaking as opposed to base load units (1) and availability is greatly reduced for peaking units if starting reliability is poor.

4.5 DURABILITY QUANTIFICATION

Durability is an engineering calculation of the normal replacement interval or design life of a part. The equipment manufacturer may design longer life components and in this way have a direct influence on reliability. Durability differs from reliability but contributes to it. Durability of single part or component does not necessarily imply reliability because failure of other components may But, nondurability of failure of any part may reduce reliability. If the durability of specific components can be pinpointed as causing unreliability, then improving the durability of or introducing redundancy in these components can improve the reliability of the plant. The durability of a part can be referred to the same "Bathtub curve" (Figure 3) as for the entire plant. The horizontal axis is time and the vertical axis is the failure rate ($\frac{1}{MTBF}$) of the part. The durability of the part is the time from the beginning of the bathtub base to the point where the failure rate increases to (about) 5 percent above the base rate. This situation occurs when parts begin to wear out, that is, reach their durability life limits.

Reliability design analysis for each part will determine which failure mode is most critical. The mode may be low cycle fatigue, high temperature erosion, start cycles, coating failure, and the like. In some cases more than one failure mode may be damaging. It must be noted that recent advances in fracture mechanics and low cycle fatigue prediction techniques have been made and prediction techniques in this heretofore low confidence area, can now be made more accurately. From these analyses the durable life of each part can be predicted in terms of hours of service, given the definition of the usage cycle. The effect of other usage cycles can also be ac-

"BATHTUB" CURVE OF AVAILABLE VERSUS OPERATING LIFE



curately predicted, but in spite of all the best design and reliability engineering effort, some infant mortality problems may be encountered during early prototype testing. The reliability numbers generated in this Task, however, are for a developed commercial powerplant.

4.6 POWERPLANT RELIABILITY ANALYSIS

Electric utilities use the reliability terms given above to determine long term generation capacity needs and short term spinning reserve requirements. Long term generation planning considers individual plant reliabilities and estimated future demand to determine the capacity needed to meet system loss of load probability (LOLP) requirements. LOLP, commonly one day in ten years, -- that is, on at most one day in ten years the systems generating capacity will be insufficient to meet the system's electricity demand, is often written into the utility's requirements for service with the state Public Utility Commission. Spinning reserve requirements are typically determined 24 hours ahead on an hourly update basis by utilities. Markov processes are often used to estimate the probability of carrying the load (5) (6). The necessary parameters for a given powerplant can be calculated easily from the MTTR and MTBF for the powerplant components. Thus, the reliability terms defined above will enable electric utilities to evaluate more reliable powerplants on their systems.

These reliability definitions will likewise serve the manufacturers' purpose by enabling the evaluation of the reliability differences between candidate configurations. MTBF and MTTR are readily obtained from reliability prediction models. This provides for consistent case-to-case comparisons in terms meaningful to the manufacturers and to the electric utilities for calculation of reliability and availability.

Equipment manufacturers generally use MTTR and MTBF as basic reliability measurements. These are easily calculated from the basic utility terms in Table 1 using the definitions in Table 2. The reverse is also true since specifying the service hours required of a plant and the MTBF yields the number of expected forced outages N. N and MTTR give the forced outage hours (FOH) and the remaining terms can be calculated using the formulae in Table 2.

5.0 EQUIPMENT RELIABILITY AND ADVANCED TECHNOLOGY ASSESSMENT

5.1 INTRODUCTION

This section provides the background data from UTC, B&W and BRISC for the reliability data used in generating the overall powerplant reliability. These data are based for the most part on the published EEI information for comparable equipment.

5.2 GAS TURBINE RELIABILTY

UTC has an extensive gas turbine reliability data bank with over 8 million hours of industrial gas turbine experience (and 400 million hours of aircraft gas turbine operation). Most of this data is for peaking gas turbine uses and data on gas turbines in a base load environment are relatively scarce. Peaking use differs greatly from base load use with the gas turbine powerplant being brought to full load in a few minutes and the cycling and thermal shock loads far exceeding those for an equal number of base load hours. In addition engine modifications would be made for base load operation by optimizing various design factors. For these reasons and as a general aid in design, UTC has developed a very detailed gas turbine reliability model which projects the reliability of present and planned gas turbines in specified operating environments. A description of this model is discussed below.

Prediction techniques are based on actual mature engine failure rates to provide the most realistic estimate of advanced gas turbine reliability, since an entirely analytical approach would have greater uncertainty. UTC has developed a Reliability Prediction Model employing actual gas turbine failure rate data as illustrated in Figure 4.

In the model, the complete engine system is divided into major functional sections containing critical components and subcomponents to form a complete and valid Reliability Prediction Model. The levels of the breakdown in these functional sections is detailed enough to enable assessment of engineering features and comprehensive enough to encompass a sufficient number of failure incidents to provide statistical accuracy in the documented base failure rate.

The engine represents a basic mathematical series system; the sum of the individually predicted sectional failure rates $\sum_{i=1}^{n} \lambda_i$ equals the overall predicted average failure rate (λ engine) for the engine as illustrated in Figure 5.

A component-by-component comparison between the base-data engine components and the high reliability engine components is used to derive design and duty cycle difference adjustment factors. These factors, including differences in detail design features, duty cycle, operational environments, materials properties, number and complexity of parts and anticipated maintenance techniques, are applied to the base-data failure rates to adapt them to high reliability engine components. The predicted engine components failure rates are entered into the Reliability Prediction Model. Reliability equations then calculate the predicted overall engine failure rate and MTBF,

$$\lambda_{\text{High Reliability}} = \lambda_{\text{Base Component}} \times K_{\text{Design}} \times K_{\text{Duty Cycle}} \times K_{\text{Diagnostic}}$$

RELIABILITY PREDICTIONS ARE CREDIBLE SINCE THEY ARE BASED UPON ACTUAL DATA

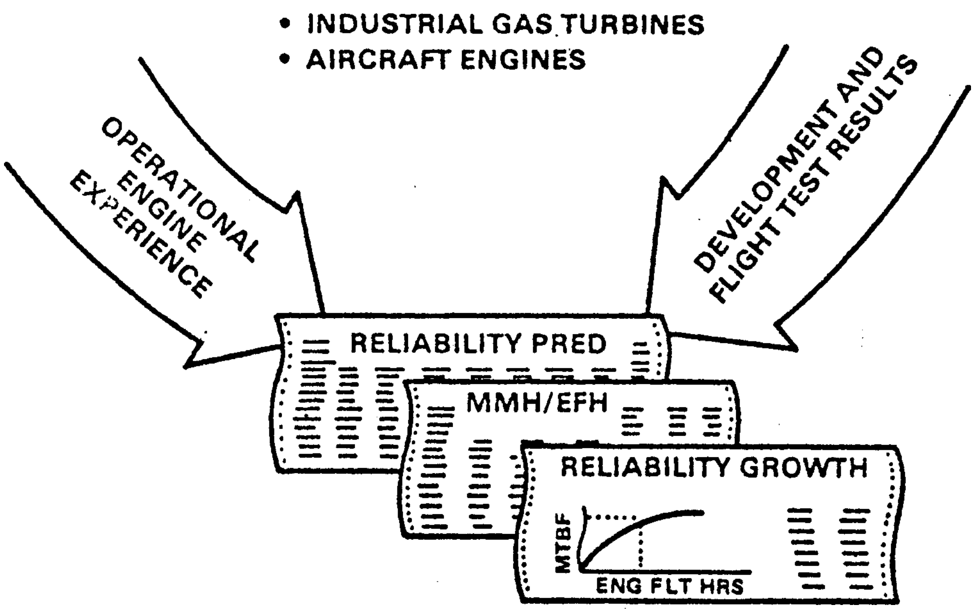
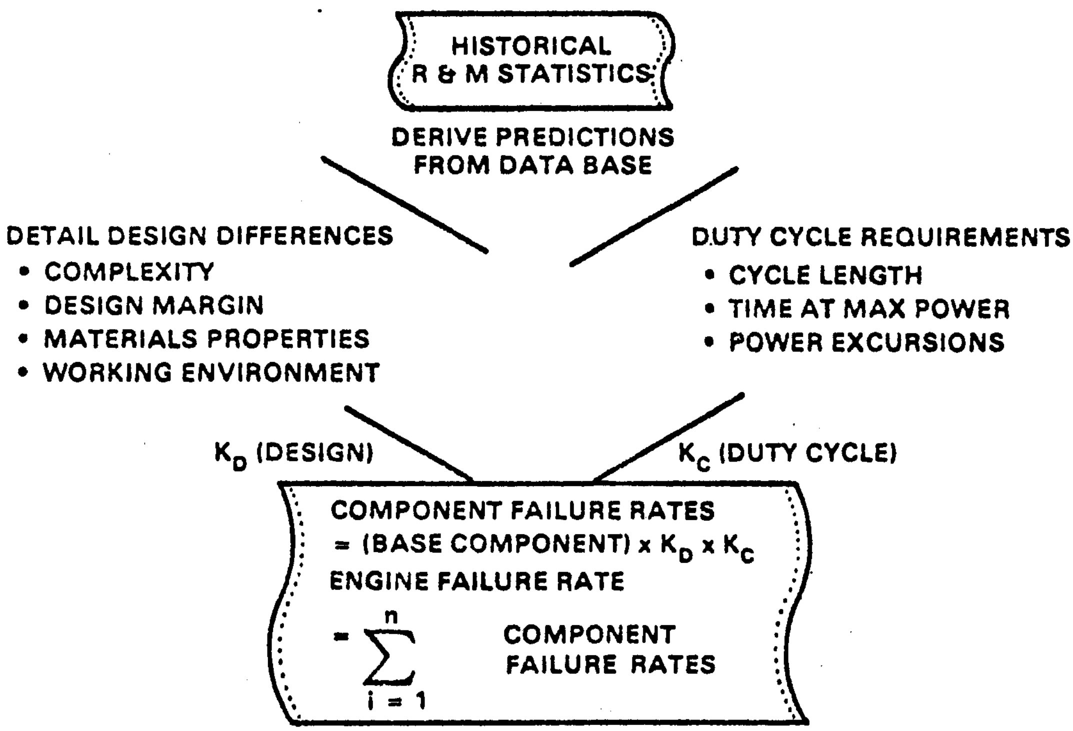


FIG. 5

RELIABILITY PREDICTION ACCOUNTS FOR DESIGN AND DUTY CYCLE DIFFERENCES FROM DATA BASE.



Successful use of the failure rate prediction technique depends upon the ability to divide and subdivide the engine into functional sections. The division must collect components which have bearing upon overall engine failure rate and which are responsive to design and duty cycle difference adjustment factors.

K Design in the foregoing equation refers to design differences such as illustrated in Figure 6 . Further adjustments are required to consider duty cycle (K Duty Cycle); and improved diagnostic techniques for detection of incipient failures (K Diagnostic). The latter factor directly accounts for maintenance as a means for reducing parts failure probabilities.

Analysis of expected plant operation serves to establish values for the duty cycle difference adjustment factor (K Duty Cycle). This factor takes into account: (1) the percent of total operating time that the engine is at maximum power; (2) the number of starts per operating hour; (3) the quantity of cycles (power excursions) per operating hour. Only one of these duty cycle adjustment factors is applied to an individual component failure rate. For example, the operating time factor would be applicable to turbine blades and the cycling factor to the compressor case assembly.

The Reliability Prediction Model has the flexibility to respond to modifications in duty cycle, improvements in diagnostic capabilities, changes in engine configurations, test and field reports, or any other formation that influences the accuracy of the model. In addition, the model becomes an effective means to expose areas expected to have a high rate of failure.

The Reliability Prediction Model is modified and developed as the design progresses to reflect the latest hardware configuration and duty cycle requirements of service engines. Documented failure history will serve to complement the initial base failure rates and refine the failure rate adjustment factors used in the model. The process of updating the model is initiated by applying the pertinent field and/or engineering change data generating a new set of failure predictions and submitting them to the same review procedure as the original predictions. Since data is continuously being collected, this cycling to maintain an updated model is a constant process.

The gas turbine component data could not be used for the PFB commercial powerplant reliability evaluation because: (1) base load gas turbine data are not available; (2) design data for the advanced gas turbine are not complete. The reliability analysis was, therefore, conducted assuming a range of reliability factors.

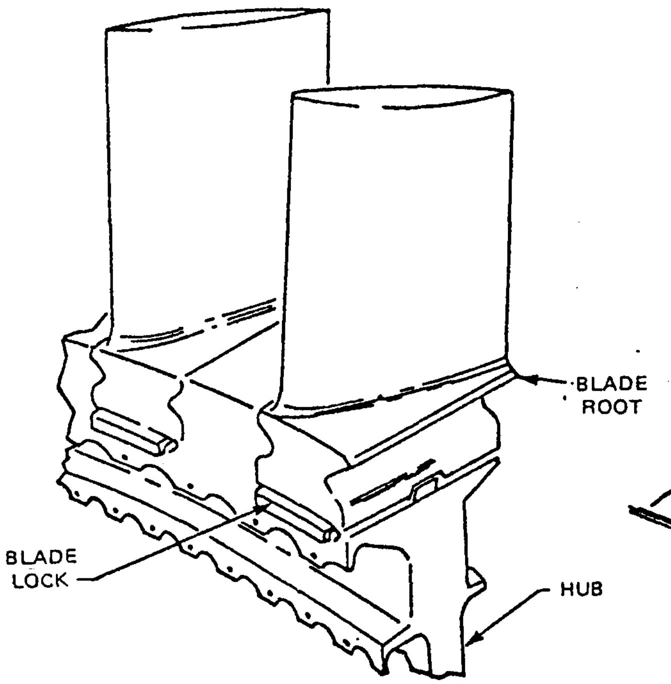
5.3 COAL HANDLING, PFBC, CLEANUP SYSTEM AND AFBC RELIABILITY

These components have similar functions to components presently used in fossil fueled powerplants. It is assumed that development of these components will result in reliability factors similar to those presently being achieved with adjustments representing more severe or less severe service due to higher or lower parameters such as pressure, temperature, corrosion and erosion.

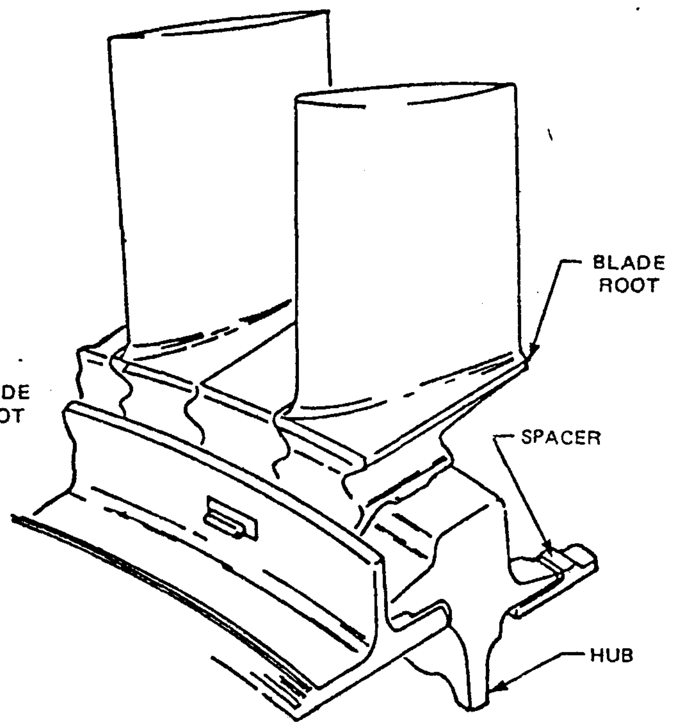
Availibility and reliability of fossil fueled power depends on five major factors:

- Design
- Quality Assurance
- Operation
- Maintenance
- Fuel Characteristics

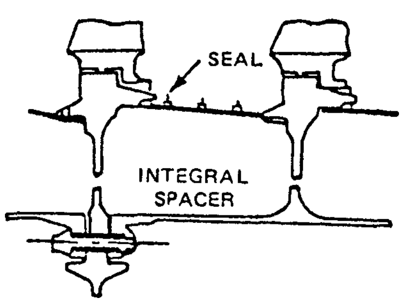
EXAMPLES OF DETAIL DESIGN DIFFERENCES CONTRIBUTING TO FAILURE RATE ADJUSTMENT FACTORS



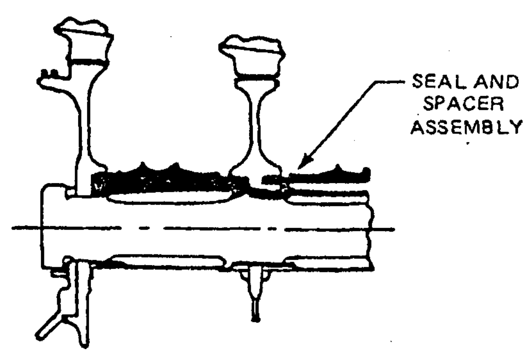
BASE ENGINE HIGH COMPRESSOR BLADE RETENTION METHOD



ALTERNATE BLADE RETENTION METHOD



BASE ENGINE DISK SPACER CONFIGURATION



ALTERNATE DISK SPACE CONFIGURATION

The design and quality assurance aspects are normally within the control of the equipment manufacturer while the operation, maintenance and fuels are usually a shared concern with the equipment user. The manufacturer supplies operating instructions, maintenance guidelines and fuel information applicable to the equipment design, but the final control of these important factors that have a major effect on availability and reliability rests with the user.

The fluid bed systems may be less affected by variation in fuel (including sulfur sorbent) characteristics than conventional boilers, but there are still areas of concern that may affect system availability. For example, changes in fuel moisture content may affect the reliability of the feed systems; changes in abrasive characteristics, the wear life of the fuel preparation and feeding systems; the ash content, the combustion gas solids loading and ultimately the life of components exposed to the gases; and the ash composition, the corrosion characteristics.

Since attention to operation and maintenance varies between users, it might be expected that availablity results would also vary. This is demonstrated by considering data available for conventional boilers.

One comparison that can be made is the availability of twenty pulverized coal fired boilers that Babcock and Wilcox has put into service since 1966. These units, all greater than 600 MW capacity have an availability of 83.9% as compared to the 82.5% of availability reported by the Edison Electric Institute (EEI) for all manufacturers over the same time period. The EEI data is based on 94 units including oil and gas fired boilers with a higher potential availability than pulverized coal fired boilers.

Even more dramatic, however, is the availability of five large pulverized coal fired units that Babcock and Wilcox has been tracking to evaluate availability performance. These units have been in commercial operation since late 1973 and are all supercritical, once through units. They include the two 1100 MW units at Belews Creek Station of Duke Power, the two 1300 MW units at Gavin Station of American Electric Power and the one 1300 MW unit at Amos Station of American Electric Power. The total operating availability of these units has been 89.8% (through 1976) and the 1976 availability was 91.2%.

Also significant is the variation in forced outage rate between these five units and the average data reported by EEI. The 1966 - 1975 EEI availability data indicates an 8% forced outage rate for units in the 600 - 800 MW size and nearly 12% for units larger than 800 MW. By contrast the forced outage rate for these five units varies from less than 2% for the Amos unit to slightly more than 4% for the Gavin unit.

The five units follow the same design philosophy as other recent B&W boilers. It is therefore concluded that much of the availability performance is beyond the control of the manufacturer.

Availability performance of new equipment cannot be predicted but can only be determined by operating history. The manufacturer designs the equipment with attention given to potential problem areas to provide good availability performance. Fluid bed combustion technology is new and there is no data available to predict reliability. Therefore availability values similar to those of the better conventional boilers have been assigned to permit overall system evaluation.

The atmospheric pressure fluid bed boiler has been assigned an availability of 90% while a lower availability of 88% has been assumed for the pressurized fluid bed because of the more complex solids handling systems.

For both systems a scheduled outage of three weeks (505 hours) has been assumed. The forced outage rate for the PFB system than is 6.6% and for the AFB system, 4.5%. Since for conventional boilers the EEI data indicates forced outage hours on the order of 50 for each occurrence, 10 forced outages have been assumed for the PFB system and 8 for the AFB system.

Designing for high reliability is an evolutionary process involving the consistent application of sound design criteria and the continual feedback of operating experiences into the design. It is quite probable that the conceptual design developed within this report will undergo significant modification as more detailed investigations are made in preparation of commercial application and still more modification as operating experience is accumulated.

The current design has paid significant attention to differential thermal expansion as past boiler experience has indicated the importance of this area. Also, to improve reliability redundancy has been provided in the critical lock hopper and valving area of the pressurized solids handling system.

Although general problems can be minimized in the design evolution process, features must also be incorporated to reduce outage time when it does occur. Such features include ready access to areas requiring inspection and maintenance (in the PFB this includes both sides of the distributor plate, the solids feed lines, the various packing arrangements provided to accomodate extreme differential expansion).

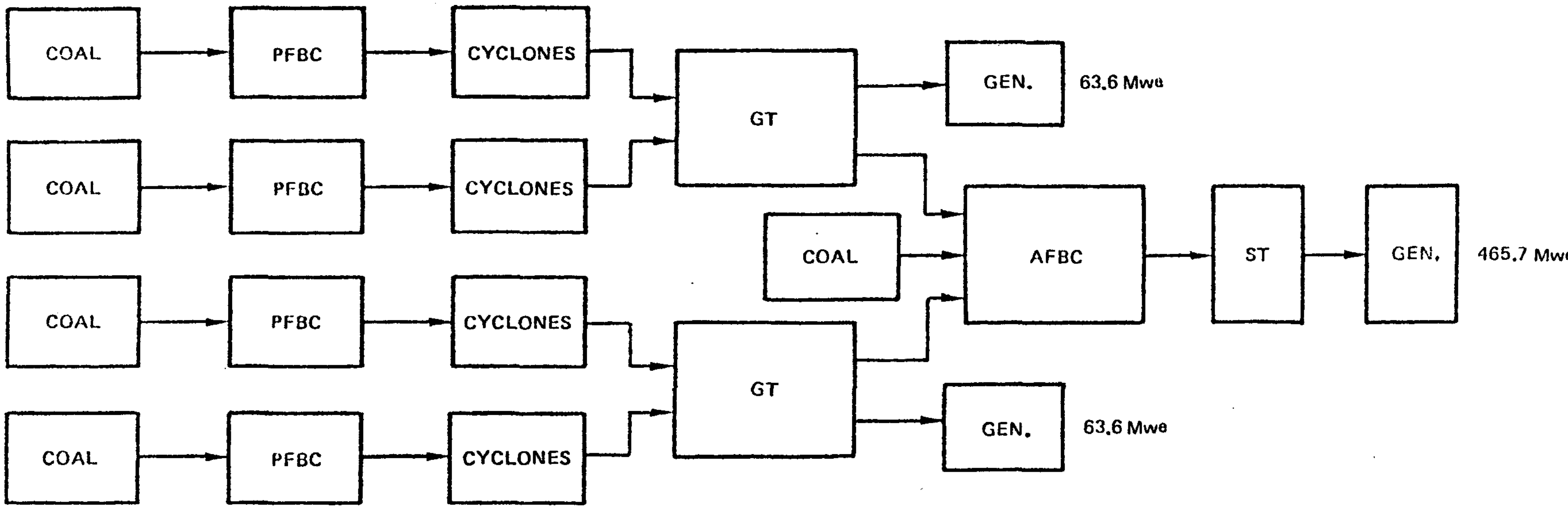
5.4 BALANCE OF PLANT RELIABILITY

The components used in the balance of plant will be advanced relative to today's units, but will be current state of the art at the time. B & R has had extensive experience in powerplant design and the high reliability of balance of plant equipment will be maintained. From an overall powerplant reliability viewpoint the steam turbine/generator reliability is most influential.

Both steam turbines and electric generators have high reliabilities as reflected in EEI data on electric utility experience. These high reliabilities reflect both the intrinsic reliability of the equipment and, particularly for steam turbines, the detailed instructions given plant operators by the manufacturers. Steam turbines must be, and are, operated with great care to avoid excess wear and fatigue from thermal expansions. Gradual warmups and shutdowns and gradual changes in loads (eg., changes limited to 3% of rated power per minute) are part of the manufacturer's operating instructions.

Steam turbine and generator reliabilities were estimated based on a survey of EEI data (1) (2) (3) for various steam powerplants. Assuming base load operation for generators, one outage of 1 day and 1 week of scheduled maintenance per year were assumed. The resulting generator availability is 98% and reliability is 99.7%. For steam turbines, one outage of 1 week and 2 1/2 weeks of scheduled maintenance per year were assumed. The resulting steam turbine availability is 93% and reliability is 98.0%.

BLOCK DIAGRAM OF PFB/AFB BASE DESIGN FOR RELIABILITY CALCULATIONS



GROSS PLANT OUTPUT 592.9 Mwe
NET PLANT OUTPUT 574.2 Mwe

FIG. 7

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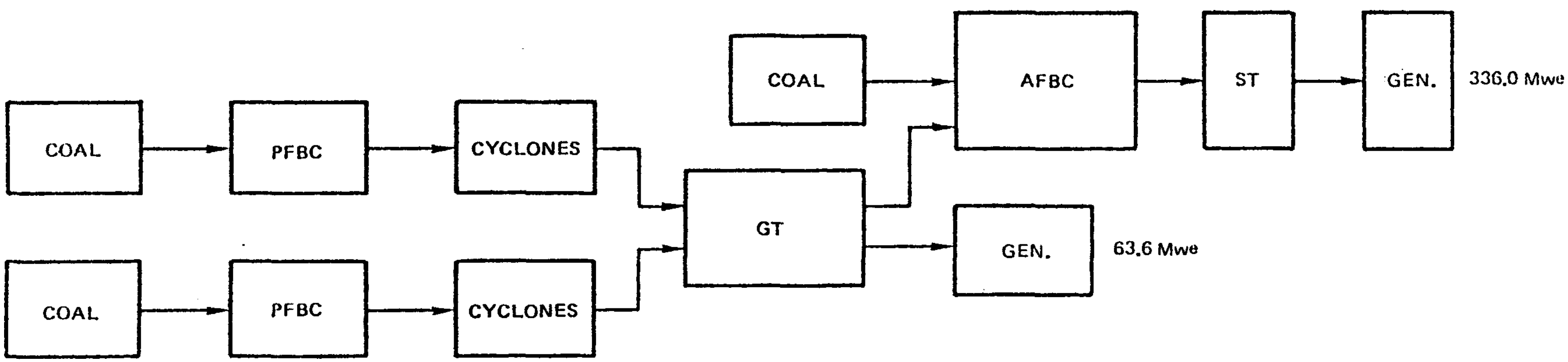
BLOCK DIAGRAM OF PFB/AFB PLANT WITH ONE GT SYSTEM OPERATING



GROSS PLANT OUTPUT 399.6 Mwe
NET PLANT OUTPUT 387.0 Mwe

FIG. 8

6.0 POWERPLANT RELIABILITY MODEL

6.1 INTRODUCTION AND GENERAL DESCRIPTION

This section discusses the UTC Powerplant Reliability Model which was developed specifically to determine overall powerplant reliability given the reliability of the plant's components. The model was originally developed to determine the value from a reliability viewpoint, of adding redundancy to a gasified coal/combined cycle powerplant and the effect on reliability of using a large common steam turbogenerator with several parallel gasifier/gas turbine waste heat boiler trains (7). One result of the study was that the worth of the redundancy only shows up in powerplant part load characteristics. This is consistent with recent results showing that electric utility system loss of load probablilties (LOLP) can be met with a lower capacity margin (less equipment) if the LOLP is calculated using part load reliabilities rather than an equivalent full load reliability [6]. The model is currently being extended to include Markov process representations of the short term (spinning reserve) reliability for gasified coal/combined cycles.

Input data for the powerplant reliability model include the mean time between failures (MTBF) and mean time to repair (MTTR) and planned maintenance (weeks/year) for the major equipment and auxiliaries in a powerplant. For example, since the previous study [7] focused on the effect of parallel trains, the equipment considered was limited to the coal gasifier/cleanup system, gas turbine, waste heat boiler, steam turbine and generators. For that study, these key components delineated the advantages and disadvantages of parallel and spare trains with individual or common steam turbogenerators. For the Task 4 studies the key components were specified as applicable giving more detail to those components which plan a key role. Once the number and kind of components and the way they are linked together (from a reliability point of view) is specified, the MTBF and MTTR and planned maintenance for each component is specified as input data.

Probabilistic combinations of the component reliabilities (calculated from their MTBF and MTTR) are then made by the model giving an overall plant reliability. In addition, the plant availability is then calculated using the planned maintenance times. The basic formulae for these probabilistic combinations are the standard textbook formula for components connected in a series or parallel [5] from a reliability viewpoint. These basic formulae are combined for a powerplant including the parallel trains using common equipment. The resulting formulae become lengthy and graphical output of (e.g.) available power versus percent of time available --i.e., part load characteristics, are derived to simplify the output.

The output of the powerplant reliability model includes the reliability and availability of the overall powerplant as well as a graphical representation of the part load characteristics of the overall powerplant. Detailed data on the probabilities of plant outage under the various combinations of component outages and the overall plant MTBF and MTTR are also printed out.

6.2 APPLICATION TO PFB/AFB SYSTEM

The general powerplant reliability model was applied to the Task 2 PFB/AFB system and is shown in block diagram form in Figure 7. In this system four coal/PFB/cyclones trains feed two gas turbines which drive electric generators to produce 63.6 MWe each. Part of the exhaust gas from the two gas turbines, together with coal feeds a single AFB which, via a steam turbine and generator, produces 465.7 MWe. The gross plant output is 592.9 MWe and the net plant output, after auxiliaries, is 574.2 MWe. To calculate reliability it is assumed that the gas turbine cannot be run with just one PFB operating. However, if either of the Coal/PFB/Cyclones/GT systems is out, then the system can be run as shown schematically in Figure 8. Under this part load condition a total of 399.6 MWe gross (387.0

MWe net) is produced. (More coal is fed into the AFB than in the half load case.)

The power output described above is combined with the reliability data (summarized in Table 3) in the powerplant reliability model to give the overall powerplant reliability. The PFB and AFB reliability data are based on availabilities of 88% and 90%, respectively, three weeks scheduled maintenance per year (scheduled outage hours) and 10 outages per year for the PFB and 8 for the AFB. These reliability data are consistent with what B & W could expect for mature coal fired steam plants. The data include the coal feeding and cyclone cleanup systems and associated auxiliaries. Gas turbine reliability was varied from 90 to 99% to test the effect of improved gas turbine reliability on overall powerplant reliability. Scheduled maintenance for the gas turbines was set at fifteen days per year with 3.5 days MTTR per outage. The electric generator MTTR was set at 25 hours with 1 outage per year and 7 days scheduled maintenance. This results in a high reliability of 99%. The steam turbine reliability was set at 98% with 1 1/2 weeks scheduled maintenance per year and 1 outage per year.

Variations in reliability could have been assigned to each of the major components and similar resultes in plant factor would be shown. The gas turbine was selected to provide an example of the analytical methods used.

TABLE 3

SUMMARY OF POWER PLANT COMPONENT RELIABILITY DATA

Base Load Operation

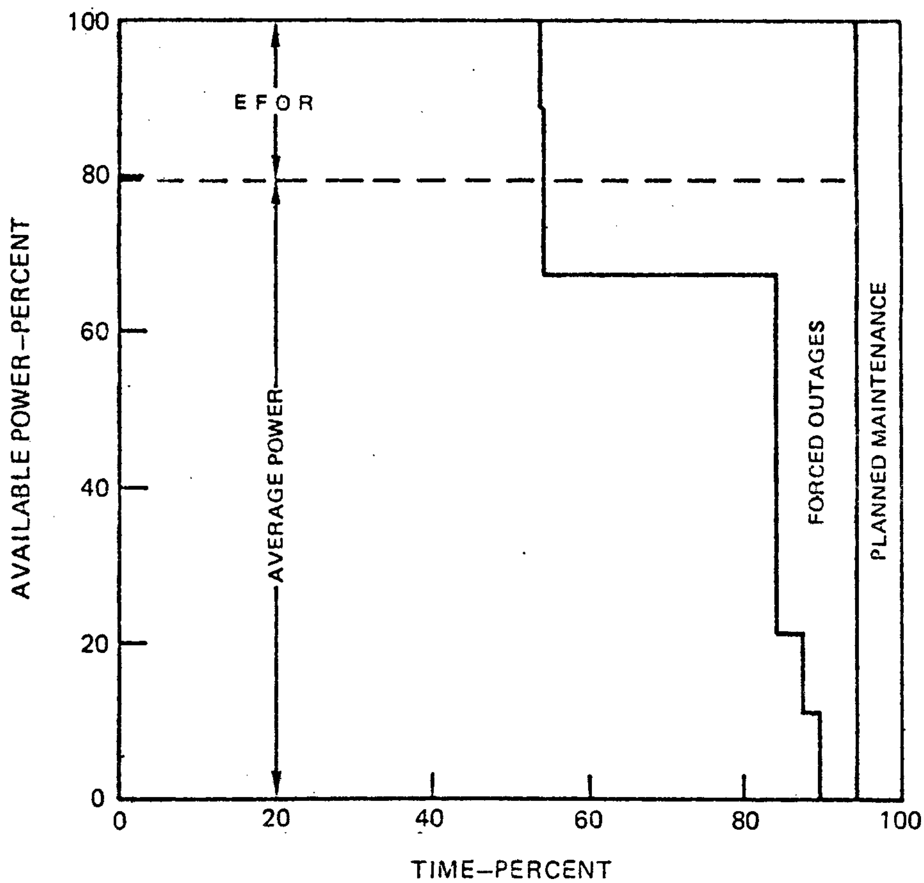
	MTBF	MTTR	N	SOH	R	A
Coal/PFB/Cyclone	770.9	54.6	10	505	93.4	88.0
Coal/AFB	985.5	46.4	8	505	95.5	90.0
Gas Turbine	756	84	10	360	90.0	86.3
	1596	84	5	360	95.0	91.1
	8316	84	1	360	99.0	94.9
Generator	8567	25	1	168	99.7	97.8
Steam Turbine	8170	170	1	420	98.0	93.3

MTBF Mean Time Between Failures
MTTR Mean Time to Repair
N Number of Outages/Year
SOH Scheduled Outage Hours/year
R Reliability (of component)
A Availability (of component)

MTBF, MTTR, SOH in hours
R, A in percent

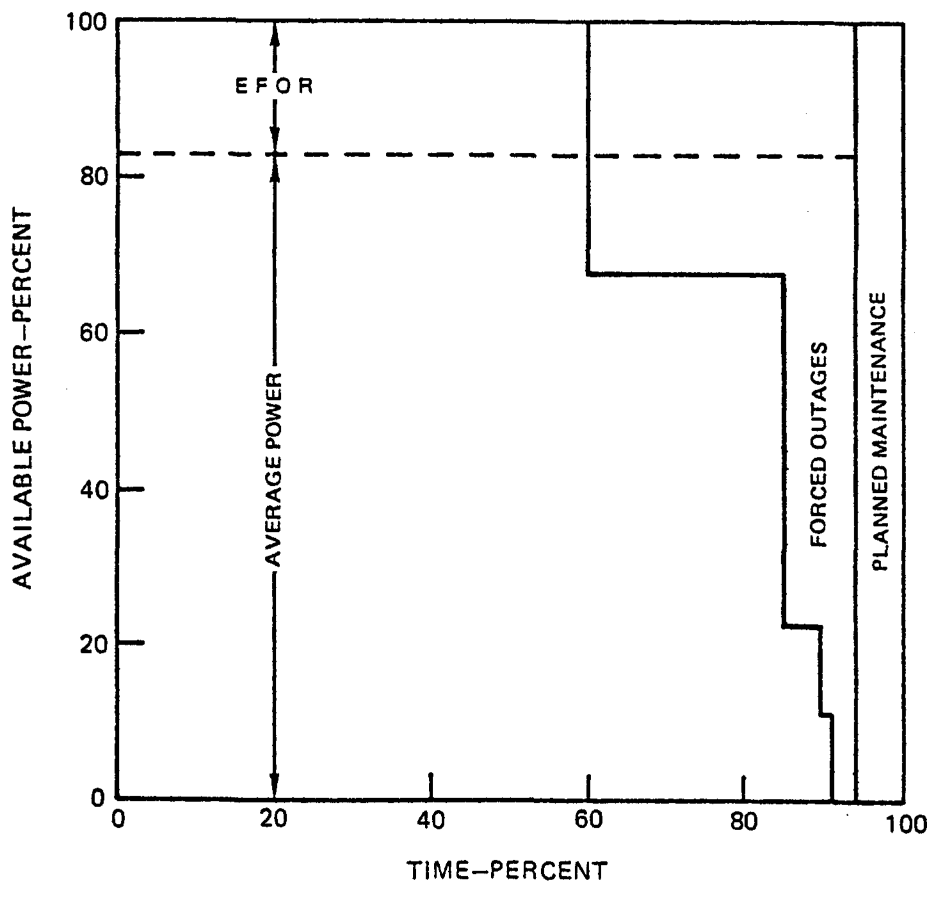
POWER PROFILE FOR PFB/AFB POWER PLANT WITH 90% RELIABILITY GAS TURBINE

R(GT)	90%
E F O R	20.4%
AVE. POWER	79.6%
PF	75.0%



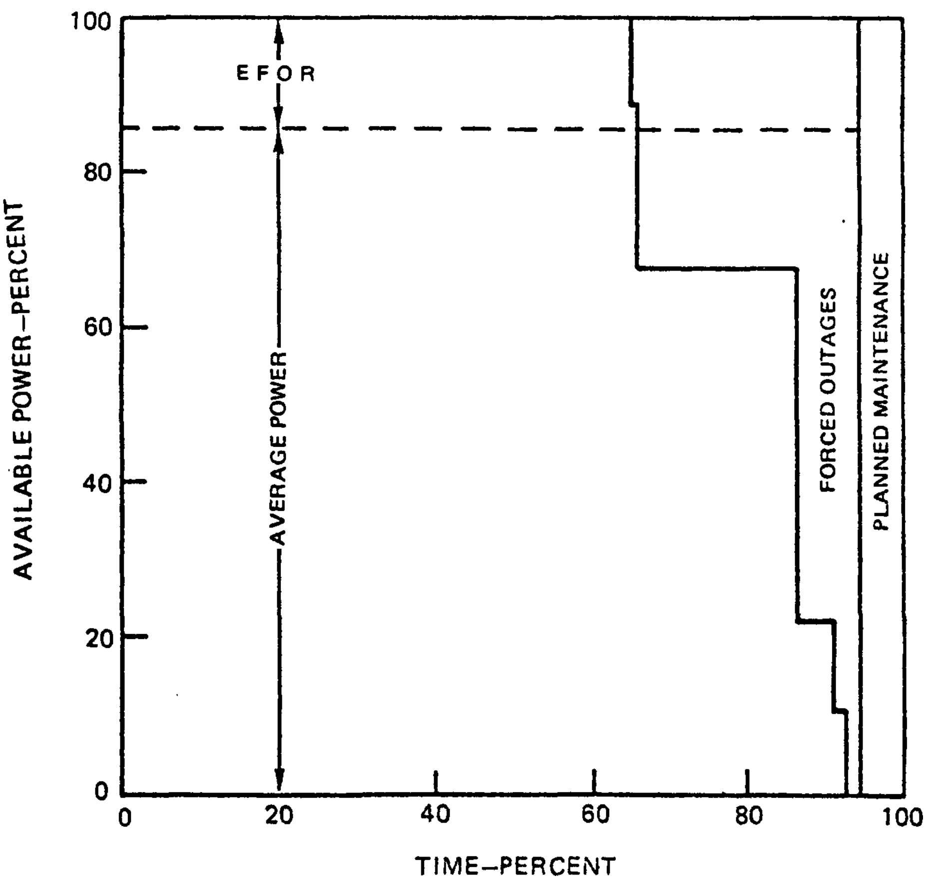
POWER PROFILE FOR PFB/AFB POWER PLANT WITH 95% RELIABILITY GAS TURBINE

R(GT)	95%
E F O R	17.0%
AVE. POWER	83.0%
PF	78.2%

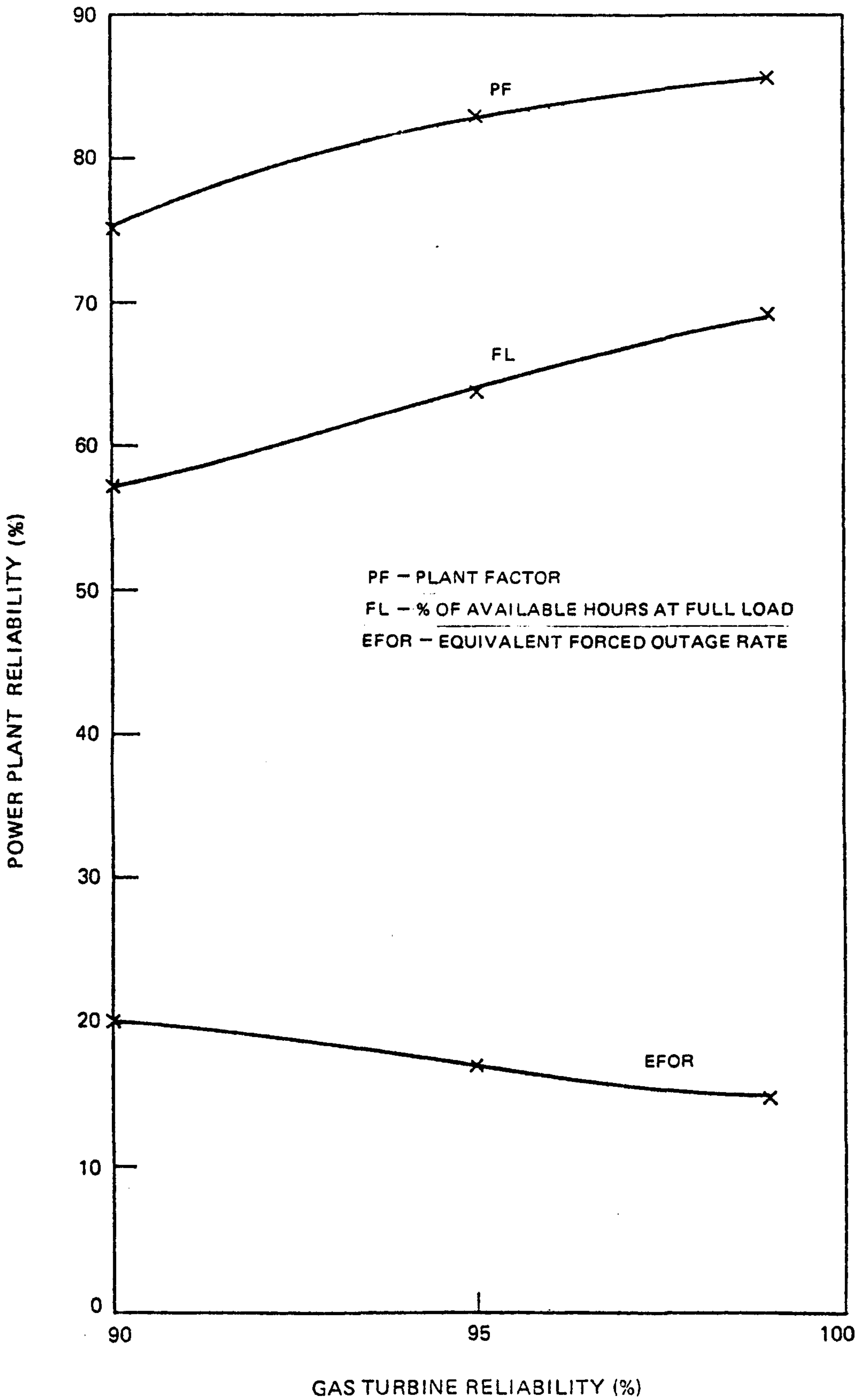


POWER PROFILE FOR PFB/AFB POWER PLANT WITH 99% RELIABILITY GAS TURBINE

R(GT)	99%
E F O R	14.5%
AVE. POWER	85.5%
PF	80.6%



POWER PLANT RELIABILITY VS GAS TURBINE RELIABILITY



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7.0 RESULTS

The effects on powerplant reliability while varying gas turbine reliability from 90 to 99% are discussed below and illustrated in Figures 9-12. Increasing the gas turbine reliability from 90 to 99% caused the powerplant Equivalent Forced Outage Rate (EFOR) to decrease from 20.4% to 14.5% and the Plant Factor (PF = KWHr produced/KW rated X PH) to increase from 75.0% to 80.6%. The time the powerplant runs at full load increases from 53.8% to 65.1% of the year with the increased gas turbine reliability. These results are summairzed in Figure 12.

The detailed power profiles shown in Figures 9-11 summarize the reliability calculations made by the UTC powerplant reliability model. Figure 9 shows the full and part load operating times of a gas turbine with a reliability of 90%. Planned maintenance (scheduled outage hours) of 3 weeks are indicated to the right--during which time the maintenance of all powerplant components would be accomplished concurrently. Since base load operation is assumed, the remaining time is divided between powerplant operation and forced outages. Most of the time the plant is operating with one or both gas turbine generators and the steam turbogenerator producing output. The cases of the AFB/ST/GEN system being forced out and the gas turbine generators alone producing power are represented by the two small steps on the right side of the graph. Elimination of these uneconomic operating modes has little effect on powerplant reliability because of their low probability. The rest of the power profile shows full load operation for 53.8% of the year. The part load operation shown in Figure 9 has one PFB/GT operating for another 29.4% of the year. Based on these two modes, the powerplant has a plant factor of 73.6% and a EFOR of 21.9%--compared to 75.0% and 20.4% considering all possible operating modes.

Figure 10 shows the power profile for powerplants including gas turbines with 95% reliabilities. The powerplant factor is 78.2% and EFOR is 17.0%. Similar to the 90% reliability gas turbine case discussed above, the plant operating modes illustrated in Figures 7 and 8 dominate the powerplant reliability. If these two modes alone are considered the plant factor is 76.6% and EFOR is 18.6%. The powerplant operates at full load 59.9% of the time and at the part load operation shown in Figure 8 for another 24.9% of the time.

Figure 11 shows the power profile for the powerplants including gas turbines with 99% reliabilities. The powerplant factor is 80.6% and EFOR is 14.5%. As in the above cases, only a small change in the powerplant reliability is obtained by dropping part load conditions other than the two considered above. If these two modes alone are considered the plant factor is 79.1% and EFOR is 16.1%. The powerplant operates at full load 65.1% of the time and at the part load operation shown in Figure 8 for another 20.7% of the time.

Noting the parallel trains of Coal/PFB/Cyclone/GT/Gen in Figure 7, a question arises whether performing maintenance on one train while the other is operating would improve plant reliability. However, analysis shows that due to the dominant effect of the Coal/AFB planned maintenance of 3 weeks, the plant factor is in fact lowered by not doing all the planned maintenance at the same time. Thus, the results presented above represent optimum scheduled maintenance from a powerplant reliability viewpoint.

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FE-2371-36

EVALUATION OF A PRESSURIZED-FLUIDIZED BED COMBUSTION (PFBC) COMBINED CYCLE POWER PLANT CONCEPTUAL DESIGN

Final Report

Environmental Analysis

Subtask 1.5

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Date Published - June, 1978

PREPARED FOR THE UNITED STATES

DEPARTMENT OF ENERGY

Under Contract No. EX-76-C-01-2371

ABSTRACT

In June 1976, the U. S. Department of Energy (DOE) awarded a contract to an industry team consisting of Burns and Roe Industrial Services Corp. (BRISC), United Technologies Corp. (UTC), and the Babcock & Wilcox Company (B & W) for an "Evaluation of a Pressurized, Fluidized Bed Combustion (PFBC) Combined Cycle Power Plant Design."

The results of this program indicate that pressurized fluidized-bed combustion systems, operating in a combined-cycle power plant, offer great potential for producing electrical energy from high sulfur coal within environmental constraints and at a cost less than conventional power plants utilizing low sulfur coal or flue gas desulfurization (FGD) equipment.

As a result of various trade-off studies, a 600MW combined cycle arrangement, incorporating a PFB combustor and supplementary firing of the gas turbine exhaust in an atmospheric fluidized bed (AFB) steam generator, (i.e., a CCFBC plant) has been selected for detailed evaluation.

The overall program consists of the following subtasks:

1.1 Commercial Plant Requirements Definition

1.2 Commercial Plant Design Definition

1.3 System Analysis and Trade-Off Studies

1.4 Reliability and Maintainability Evaluation With Advanced Technology Assessment

1.5 Environmental Analysis

1.6 Economic Analysis

1.7 Evaluation of Alternate Plant Approaches

1.8 PFB/Gas Turbine/ Waste Heat Boiler Cycle Study

1.9 PFB/Gas Turbine/ Power Turbine Reheat Cycle Study

This interim report discusses the results of studies performed under Subtask 1.5.

REPORT ON SUBTASK 1.5

ENVIRONMENTAL ANALYSIS

EVALUATION OF A PRESSURIZED-FLUIDIZED BED COMBUSTION (PFBC) COMBINED CYCLE POWER PLANT

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1.0 SUMMARY

Recent studies indicate that combined cycle power plants, using fluidized-bed combustors, offer the potential for the production of electrical power from coal in an environmentally acceptable manner at higher efficiency and at lower lost than conventional coal-fired steam power plants incorporating flue gas desulfurization systems. The team of Burns and Roe Industrial Services Corporation, United Technologies Corporation and the Babcock & Wilcox Company, under contract to the Department of Energy, conducted a conceptual design study of such a combined-cycle power plant.

The major objectives of the study are to prepare a conceptual design and cost estimate for a 600 MWe combined cycle plant using coal fired, air cooled pressurized fluidized-bed combustors and gas turbines, coupled with supplementary firing of the gas turbine exhaust in an atmospheric fluidized-bed steam generator. In addition, areas requiring further technology development were to be identified.

Subtask 1.5 involves an evaluation of the environmental aspects of a nominal 600 MW combined cycle PFBC/AFBC (CCFBC) plant design. To accomplish this evaluation the following areas are considered:

Sources of emissions/effluents

Characteristics of gaseous, liquid, and solid wastes

Capability for compliance with existing and projected environmental regulations

Options for environmental control

Land use

Health and Safety

Much of the material contained in this report was obtained from an earlier study performed at Burns and Roe, Inc. on atmospheric fluidized bed power plants (Ref. 1).

Fluid bed combustion, as currently discussed, involves the combustion of coal in a fluid bed containing a crushed sulfur acceptor such as limestone or dolomite. The temperature of the combustion process is controlled by heat extraction from the bed and/or by controlling the fuel-air ratio in the bed. It is necessary to maintain the bed temperature below the coal ash softening temperature and in a range that provides efficient capture of sulfur dioxide by the sulfur acceptor (1550-1650F). With this approach it is possible to burn high-sulfur coals and still meet Federal Environmental Protection Agency (EPA) limits for SO_2 emissions. Also, due to the low combustion temperature NO_x formation is well within Federal EPA limits. Pressurized fluid bed (PFB) combustion is similar to atmospheric fluid bed (AFB) combustion except that the process takes place under a pressure of several atmospheres such as that which exists at the exhaust of the compressor of a gas turbine unit. PFB combustors, therefore, offer the potential of serving as the energy source for the gas turbine working fluid.

The FBC flue gases, however, are characterized by a high dust loading having a wide particle size distribution which reflects the size characteristics of the combustor feed material. While control of particulate emission from the PFB is a major problem from the gas turbine viewpoint, a variety of control techniques are available for the final particulate collection stage prior to the stack. Therefore, with proper selection of equipment, stack particulate emissions are not expected to pose an environmental concern beyond that which exists for conventional coal fired plants. In the FBC plant design, mechanical collectors and electrostatic precipitators are included for particulate control. It is noteworthy that the fugitive dust emissions from coal, sorbent and waste material handling and storage may be equivalent to 50% of the EPA stack emission limits.

From an environmental viewpoint, one of the primary concerns with fluid bed technology is the disposal and/or utilization of the solid wastes generated.

The solid wastes from the fluid bed system are composed of the spent material from the beds and the particulates captured in the various particulate collection systems. This material is dry, which suggests the idea of disposal by landfill. In addition, the waste material may be used almost immediately where dry CaO and $CaSO_4$ are the desired materials.

Consideration for the disposal and/or utilization of both the PFB solid wastes and the AFB solid waste must consider the potential for the presence of CaO. Calcium sulfate ($CaSO_4$) and calcium carbonate ($CaCO_3$) are chemically stable and are suitable for direct disposal. However, unreacted lime (CaO) will readily hydrate to form calcium hydroxide ($Ca(OH)_2$) on contact with water and will recombine with CO_2 in the presence of moist air. Both these reactions, although exothermic and resulting in a significant heat release, may pose no environmentally adverse effects. However, further efforts are required to confirm this.

Other environmental concerns with FBC solid waste disposal include leaching, dusting, and run-off. Potential problems with FBC leachate include pH, total dissolved solids (TDS), sulfate, and calcium.

Although no actual set of standards exists for leachate, threshold limits will be established by the Federal EPA as soon as the data is compiled. With this set of standards for comparison, the trace elements concentrations of actual landfills may exceed the allowable limits. It is noteworthy that the solubility of trace metal decreased with increasing pH. The high pH of the fluid bed material tends to cause trace metals to precipitate in the limestone mixture. This may prevent most of the trace metals from leaving the landfill in the leachate. Monitoring of a test landfill area would provide the information needed to evaluate this problem.

In addition to landfilling, a number of FBC waste material utilization options are being tested such as, agricultural fertilization, stabilization and neutralization of municipal sewage waste, the production of building blocks and road bases, etc. The principal obstacles to large-scale utilization of FBC solid residue by the above methods appear to be market saturation, availability of competitive raw materials, and transportation costs. Other approaches, such as sorbent regeneration, use of alternate sorbents, or improving the reactivity of calcium based sorbents (i.e. controlled precalcination, catalyst addition, reduced sizing, increased residence time, etc.) may eventually reduce the waste disposal problems significantly. However, much additional development work is required in these areas to demonstrate technical and economic feasibility.

The FBC process may also release a variety of trace constituents in the form of organic compounds, acid and acid anhydrides, halogens, compounds of nitrogen and of sulfur, radioactive isotopes, and trace elements and their compounds, in addition to the previous emissions discussed. Although a substantial portion of the trace constituents present in the combustor feed is retained in the coarse bed material residue or captured in the flyash removed by the particulate collection equipment, trace substances are increasingly of concern environmentally. Some trace substances vaporize and are emitted with the flue gas; while others selectively concentrate on sub-micron size particles, escaping collection in the particulate control equipment. These substances may also be leached from the solid residue, creating potential ground and surface water problems.

The overall water demand, water and waste water treatment, noise emissions, and land requirements for the CCFBC plant are expected to be similar to those of a conventional coal fired plant. Since such items are site specific, determination of plant requirements must await selection of such a site. However, since a portion of the plant's output comes from the gas turbines, cooling tower make-up flows may be less than for a conventional 600 MW steam plant. At this time, liquid wastes have not been identified as a significant environmental concern except indirectly as leachate water from the solid wastes.

Plant cooling systems are conventional and should not pose any unusual problems. The health and safety concerns of the FBC generating plant are comparable, in general, with those found in conventional pulverized coal-fired plants. There are processes included in the CCFBC plant, however, that are not found in conventional power plants. The design of these special processes include safety systems which are expected to meet all environmental safeguards and be acceptable to the regulatory agencies.

Table 1 summarizes the predicted environmental intrusions versus the current standards for the 600 MW combined cycle FBC power generating plant. It is evident that the pressurized and atmospheric fluidized-bed combustion techniques can burn coal in an environmentally acceptable manner, without the use of costly flue gas desulfurization equipment. Additional effort, however, is needed to verify FBC performance and to demonstrate its viability in commercial applications. As of this time, information concerning fluidized-bed combustion technology is based upon analytical models, bench-scale testing, and operation of process development units. There is virtually no direct experience with

TABLE 1

EFFLUENTS* FROM PFB/AFBC COMBINED CYCLE POWER PLANT
(574 MW Net Output)

Solid Waste Disposal Method	Off-Site
Solid Waste Output Rate, tons/hr	68.5
Specific Solid Waste Output, lbs/KW hr	0.24
Emissions, lb/10^6BTU Input*	
SO_2/SO_2 Standard	1.2/1.2 = 100%
NO_x/NO_x Standard	0.4/0.7 = 57%
Particulate/Particulate Standard	0.09/.1 = 90%
Emissions, lb/MW hr Output*	
SO_2 (6000 lbs/hr)	10.75
NO_x (2000 lbs/hr)	3.58
Particulate (460 lbs/hr)	.81
Thermal Discharge x 10^6BTU/hr*	
Cooling Towers	2352
Stack and Miscellaneous	700
Cooling Tower Blowdown Rate, gpm*	1600
Waste Water Rate from W.W. Treatment Plant, gpm	130 (avg)

*At operating conditions and 100% load factor, HHV coal = 12453 BTU/lb as fired.

operation of a large commercial-scale FBC unit for utility application. The 30 MWe FBC demonstration plant at Rivesville, West Virginia, has started up recently. Any evaluation of the environmental impact of utility-size FBC coal-fired power plants will be aided by the design, construction, and operation of larger FBC demonstration plants.

Before commercial application of the CCFBC plant concept can become a reality, further investigation is required in various areas of environmental concern. These areas have been indicated throughout this report and are summarized below.

1. Characterization of emissions from utility-scale PFB and AFB power plants as they vary with unit size and operating parameters.

2. Performance of carbon burn-up cells.

3. Performance of particulate collection equipment (especially the final clean-up device).

4. Minimum required Ca/S mole ratio for acceptable sulfur retention.

5. Optimum coal and sorbent particle sizing.

6. Sorbent characterization (prediction of composition, sulfur reactivity, attrition rate).

7. Calcination phenomena in PFB and AFB combustors.

8. Effect of sorbent enhancement agents (performance, corrosion potential, environmental releases).

9. Feasibility of alternate sorbent materials (performance, cost, regeneration potential).

10. Technical and economic feasibility of sorbent regeneration (including conversion efficiency, sorbent deactivation, off-gas concentration, capital and operating cost).

11. Characterization of FBC solid wastes (quantities, physical and chemical properties, utilization/disposal methods, runoff composition).

12. Effectiveness of equalization basins and clarifiers in removing elements from waste water.

13. Suitability of waste clarifier sludge for landfill or other uses.

14. Suitability of solid waste removed from waste water for land fill or other ultimate disposal method.

15. Effluent limits on combined source discharge when individual source rates are variable and not well known.

16. Trace element emissions and methods for controlling them.

17. CO and hydrocarbon emissions from FBC units and methods for reducing them.

2.0 SITE DESCRIPTION

The plant site description outlined in this report for the 600 MW combined cycle PFBC/AFBC (CCFBC) power generating plant is hypothetical and is not intended to describe a specific plant location. The climatic data and river water conditions are typical of those which can exist and are based on many years of designing and building utility power plants all over the U.S.A.

2.1 TOPOGRAPHY AND GENERAL FEATURES

The site is located on the east bank of the North River at a distance of 25 miles north of Middletown, the nearest large city. The North River flows from north to south and is 2600 feet wide adjacent to the plant site. A flood plain extends from both river banks an average distance of one-half mile, ending with hilltops generally 150 to 250 feet above the river level. Beyond this area, the topography is gently rolling, with no major critical topographical features. The plant site itself extends from river level to elevations of 50 feet above river level. The primary structures and the switchyard will be located on level ground at an elevation 18 feet above the mean river level (Grade Level 0'-0"). This elevation is 10 feet above the 100-year maximum river level, according to U.S. Army Corps of Engineers studies of the area.

2.2 SITE ACCESS

Highway access is provided to the site by a secondary road entering at the northeast corner and connecting to a state highway. This road is in good condition and needs no additional improvements. Therefore, no capital funds are included for roads outside the plant fence. Railroad access is provided by constructing a railroad spur which intersects the B&M Railroad. The length of the required spur from the main line to the plant site is approximately five miles in length. The North River is navigable throughout the year with a 40 foot wide channel, 12 feet deep. Coal and limestone will be normally delivered by barge, with backup by railroad. The Middletown Municipal Airport is located three miles west of the state highway, 15 miles north of Middletown and 10 miles south of the site.

2.3 POPULATION DENSITY AND LAND USE

The hypothetical site is near a large city (Middletown, 250,000 population) but in an area of low population density. Variation in population with distance from the site boundary is:

Miles	Cumulative Population
0.5	0
1.0	310
2.0	1,370
5.0	5,020
10.0	28,600
20.0	133,000
30.0	1,010,000

There are five industrial plants within 15 miles of the plant site. Four are small plants employing fewer than 100 people each. The fifth, near the airport, employs 2,500 people. Densely populated areas are found only in the centers of the small towns so the total land area used for housing is small. The remaining land, including land across the river, is largely used as forest or cultivated crop land.

2.4 PUBLIC UTILITY SERVICES

Utilities available are as follows:

1. Communication lines will be furnished to the project boundaries at no cost.

2. Power and water for construction activities are available at the northwest corner of the site boundary.

2.5 GROUNDWATER AND HYDROLOGY

Groundwater occurs at an elevation of 27'-0", that is, the groundwater is 11 feet below the ground level on which the primary structures and switchyard are located.

The North River provides an adequate source of water makeup for the station. The water is turbid and requires treatment before added to the plant makeup system.

2.6 METEOROLOGY AND CLIMATOLOGY

2.6.1 Prevailing Winds

According to Weather Bureau records at the Middletown Airport, located 10 miles south of the site on a low plateau east of the North River, surface winds are predominantly southwesterly, 4-10 knots during the warm months of the year and westerly, 6-13 knots during the cool months.

There are no large diurnal variations in wind speed or direction. Observations of wind velocities at various altitudes indicate a gradual increase in mean velocity and a gradual veering of the prevailing wind direction from southwest and west near the surface to westerly and northwesterly aloft.

In addition to the above, studies of the area indicated that there is a significant channeling of the winds below the surrounding hills into the north-south orientation of the North River. It is estimated that winds within the river valley blow approximately parallel to the valley orientation in excess of 50 percent of the time.

2.6.2 Atmospheric Diffusion Properties

During the warm months of the year, according to analyses of Weather Bureau records, the atmospheric conditions near the surface are 25 percent unstable (Pasquill A, B and C), 40 percent neutral (Pasquill D) and 35 percent stable (Pasquill E and F). Average wind speeds are approximately six miles per hour during unstable conditions, 10 miles per hour during neutral conditions and four miles per hour during stable conditions.

2.6.3 Severe Meteorological Phenomena

A maximum instantaneous wind velocity of 100 mph has been recorded at the site. During the past 50 years, three storms, all of them in the final dissipation stages, have passed within 50 miles of the site. Some heavy precipitation and winds in excess of 40 miles per hour were recorded, but no significant damage other than to crops resulted.

2.6.4 Ambient Background Concentrations

Background concentrations of SO_2, NO_x, and particulates are typical of a rural area approximately 30 miles from a major industrial metropolitan center.

2.6.5 Climate

The average maximum temperature is 75°F., while the average minimum temperature is 39°F. The mean annual temperature is 57°F.

2.7 GEOLOGY AND SEISMOLOGY

2.7.1 Soil Profiles and Load Bearing Characteristics

Soil profiles for the site show alluvial soil and rock fill to a depth of eight feet; Brassfield limestone to a depth of 30 feet; blue weathered shale and fossiliferous Richmond limestone to a depth of 50 feet; and bedrock over a depth of 50 feet. Allowable soil bearing if 6,000 psf and rock bearing characteristics are 18,000 psf and 15,000 psf for Brassfield and Richmond strata, respectively. No underground cavities exist in the limestone.

2.7.2 Seismology

The site is Zone 1, as designated by the Uniform Building Code, based on the observation of three earthquakes of seismic intensities 4-6 on the Modified Mercalli scale during the period 1870-1958, causing minor damage to towns in the surrounding area.

2.8 ENVIRONMENTAL DESIGN CRITERIA

The plant has been designed to meet present federal environmental regulations as well as those regulations most commonly applied by state and local jurisdictions as presented in the following sections.

2.8.1 Air Pollution Design Criteria

2.8.1.1 Standards of Performance for New Stationary Sources (NSPS)

This standard, published in the Federal Register on December 23, 1971, presents the following air quality emissions standards for coal fired steam generators:

Air Contaminent	Max. Emissions
1. Particulate Matter	0.1 lbs/10^6 Btu Heat Input
2. SO_2	1.2 lbs/10^6 Btu Heat Input
3. NO_x	0.7 lbs/10^6 Btu Heat Input

This standard is applicable for each generating unit of more than 250 x 10^6 Btu/hr heat input. Standards are for maximum 2 hour average emission. The Clean Air Act Amendment of 1977 requires the EPA to revise these new source performance standards by August, 1978.

2.8.1.2 Federal Primary and Secondary Ambient Air Quality Standards (AAQS)

These standards, published in the Federal Register on April 20, 1971, are presented in Table 2.

Federal secondary SO_2 standards for annual arithmetic mean and 24 hour average have been revoked (Federal Register, dated September 14, 1973).

2.8.1.3 Standards for Prevention of Significant Air Quality Deterioration (PSD)

The Federal EPA (FEPA) has also issued standards, to prevent significant air quality deterioration. these regulations, published in the Federal Register of December 5, 1974 and amended in the Clean Air Act of 1977, established three "classes" permitting different allowable incremental increases in total suspended particulates and sulfur dioxides. The EPA air quality deterioration concentration increments from station emissions above regional baseline air quality concentrations for sulfur dioxide and suspended particulates for Class I, Class II and Class III areas, are presented below.

EPA AIR QUALITY DETERIORATION
CONCENTRATION INCREMENTS

	Concentration Increment, g/m^3		
Pollutant	Class I Area	Class II Area	Class III Area
1. Particulate Matter			
Annual Geometric Mean	5	19	37
24 Hour Maximum	10	37	75
2. SO_2			
Annual Arithmetic Mean	2	20	48
24 Hour Maximum	5	91	182
3 Hour Maximum	25	512	700

All areas of the country (with the exception of International Parks, National Wilderness Areas, National Memorial Parks, and National Parks which are designated as "Class I" and may not be redesignated)

TABLE 2

PRIMARY AND SECONDARY AMBIENT AIR QUALITY STANDARDS (a)

(ug/m^3)

	Primary Standards			Secondary Standards		
	SO_2	NO_x	Suspended Particulates	SO_2	NO_x	Suspended Particulates
3 hour average concentration shall not exceed (b)				1300		
24 hour average concentration shall not exceed (b)	365		260	(e)		150
Annual Average concentration shall not exceed (c)	80	100	75(d)	(e)	100	60(d)

a) Federal Register, April 30, 1971.
b) The given average concentration shall not be exceeded more than once a year.
c) This refers to the annual average of the 24 hour average samples or the annual mean when continuous monitoring techniques are utilized.
d) Geometric mean for any 12 consecutive month period.
e) The federal secondary SO_2 standards for annual arithmetic mean and 24 hour average have been revoked (Federal Register, September 14, 1973).

were initially designated "Class II" with provisions for allowing each state to reclassify any area to accommodate the social, economic and environmental needs and desires of the public.

Part C - Prevention of Significant Deterioration of Air Quality - of the 1977 Clean Air Act Amendment also includes the requirement that the facility be subject to the best available control technology for each pollutant. The act defines "best available control technology" as a limitation based on a degree of reduction which the permitting authority determines is achievable for such facility.

2.8.1.4 Emission Limits Used for Design

The plant described herein has been designed to meet the present environmental standards described in Section 2.8 and to meet the anticipated future federal environmental regulations.

Compliance with the AAQS and PSD standards (See 2.8.2 and 2.8.3) depends on the existing concentration of each pollutant in the air and on the classification of plant location under the regulations. It has been assumed that the plant location is in a class II area and sufficiently far away from a class I area to avoid possible impact on the class I area. The total allowable increments in the concentration of pollutants in the air since January 6, 1975 are specified by standards for Prevention of Significant Air Quality Deterioration (See 2.8.3.1 above). Hence, the compliance of the plant with respect to these standards may also be dependent on the time of construction of the plant.

No attempt has been made to evaluate the site specific requirements promulgated in the AAQS and PSD standards because the location of plant and the start date of construction are hypothetical. Only, the requirements of the NSPS (See 2.8.1) have been considered.

It is anticipated that, in the near future, the EPA NSPS limits on the emissions from coal fired stations will be changed to the following:

Air Pollutant	Max. Emissions
1. Particulate Matter	0.03 lb/10^6 Btu Input
2. SO_2	-Max. Emission =1.2 lb/10^6 Btu Input -Minimum Sulfur removal of 90% is required unless emissions are below 0.2 lb/10^6 Btu Input
3. NO_x	0.6 lb/10^6 Btu Input

The impact of these anticipated EPA limits on the plant design and operation, are also discussed in subsequent portions of this report.

2.8.2 Liquid Effluent Design Criteria

The Federal Water Pollution Control Act Amendments of 1972 (Public Law 92-500) sets a national goal for elimination of discharge of pollutants into navigable waters by 1985 and requires that "best available technology economically achievable" be used by 1983. Effluent limitations were published in 1974 by the Environmental Protection Agency (EPA). Of particular concern to steam plants generating electricity are the Effluent Guidelines and Standards for Steam Electric Power Generating Point Source Category, 40 CFR 423 (39 Federal Register 36186, October 8, 1974, and 40 FR 7095, February 19, 1975). Table 3 summarizes the limits on chemical - pollutant discharges for new sources.

Under Section 307 of the Water Pollution Control Act, the EPA is required to publish effluent standards for toxic pollutants. The adequacy of EPA's compliance with this section has been challenged in court, resulting in a consent decree setting forth a timetable for promulgation of effluent standards for an agreed list of priority pollutants. Such standards are expected to be finalized by March 1979.

Other court action has set aside limitations on rainfall runoff from construction sites and material storage piles and the standards have been remanded to EPA for possible revision along with the non-discharge limitation on flyash transport water. Furthermore, Section 304 of the Act requires the guidelines to be revised as appropriate at yearly intervals. Therefore, the 1983 limitations are subject to change.

2.8.3 Thermal Discharge Design Criteria

FEPA Guidelines state that the discharge of heat from the main condensers will not be allowed except from cold side blowdown from a recirculatory cooling system. Although exceptions are possible (Section 316a of the Act), this plant conceptual design assumes that none are taken.

TABLE 3

DISCHARGE LIMITS FOR NEW STEAM ELECTRIC POWER GENERATING PLANTS

Discharge, Pollutant	Limits, mg/l: Max. for Any One Day	Limits, mg/l: Average Only for 30 Consecutive Days	Remarks
All Discharges			
pH	6.0 to 9.0	-	Except once-through cooling
Polychlorinated Biphenols (PCB)	Zero	-	
Low-volume Discharge			
TSS	100	30	
Oil and Grease	20	15	
Bottom Ash Transport Water			
TSS	100	100	Allowable Discharge
Oil and Grease	20	15	$= \frac{\text{Flow x Conc.}}{20}$
Flyash Transport Water			
TSS	Zero		
Oil and Grease	Zero		
Metal-cleaning Wastes			
TSS	100	30	
Oil and Grease	20	15	
Total Copper	1	1	
Total Iron	1	1	
Boiler Blowdown			
TSS	100	30	
Oil and Grease	20	15	
Total Copper	1	1	
Total Iron	1	1	

TABLE 3 (Continued)

DISCHARGE LIMITS FOR NEW STEAM ELECTRIC POWER GENERATING PLANTS
DISCHARGE, POLLUTANT LIMITS, mg/l

	Max. Conc.	Ave. Conc.
Cooling Tower Blowdown		
Zinc	No Detectable Discharge	
Chromium	No Detectable Discharge	
Phosphorus	No Detectable Discharge	
Other Corrosion Inhibitors	No Detectable Discharge	
Free Available Chlorine[4]	0.5	0.2
Once-through Cooling		
Free Available Chlorine[4]	0.5	0.2
Area Runoff[5]		
TSS	50	
pH	6.0 to 9.0	

Notes

1. Except where specified otherwise, allowable discharge = Flow x Concentration Limit.

2. Where waste streams from various sources are combined for treatment or discharge, quantity of each pollutant attributable to each waste source shall not exceed the specified limitation for that source.

3. All sources must meet state water quality standards by 1977.

4. Neither free available chlorine nor total residual chlorine may be discharged from any unit more than 2 hours in one day, and not more than one unit of any plant may discharge free available or total residual chlorine at same time unless utility can demonstrate that the unit in a particular location cannot operate below this level of concentration.

5. Applies to all area runoff from power plant site that may reach a navigable waterway. If necessary to meet limits, a runoff storage facility no larger than that to hold a 10 year 24 hour rainfall event is required. Overflow from such facility may be discharged without treatment.

3.0 GENERAL CCFBC PLANT CHARACTERISTICS

3.1 PLANT DESCRIPTION AND PROCESS LAYOUT

The CCFBC plant is located on a 340 acre site. The major structures for this facility include the PFBC's, AFBC, gas turbine-generator enclosures, steam turbine-generator building, service building, precipitators and cooling towers. Other dedicated areas of the plant include the transformer yard, electrical switchyard, live and dead coal, dolomite and limestone storage piles, waste water treatment area and holding ponds.

A detailed description of this facility is presented in reference 2. Figure 1 is a process flow diagram which shows the plant configuration schematically.

The major plant equipment consists of four PFB units, two gas-turbine generating units, one AFB steam generator and one steam turbine/generator. The steam produced in the AFB is utilized in the steam turbine/generator units. Coal is burned in the PFB combustors and in the AFB steam generator.

Each gas turbine compressor discharges compressed air to two PFB combustors. In turn, the two PFB units supply a mixture of hot gases (i.e. air and combustion products) back to the gas turbine unit. Both gas turbine units exhaust to the AFB where the oxygen rich exhaust gas is fired with coal to produce high temperature steam. Two Aerodyne two-stage high efficiency cyclone units are installed at the outlet of each PFB combustor primarily to reduce particulate loading to the gas turbine. A multiclone cyclone arrangement is provided at the outlets of both the Main Atmospheric Fluid Bed (MFB) unit and the Carbon Burnup Bed (CBB) unit. In addition, an electrostatic precipitator provides a final stage of cleanup prior to the stack.

The selected design coal is a high sulfur, Illinois Basin bituminous coal having the analyses and properties shown in Table 4. The sorbent materials used for the conceptual plant design are dolomite for the PFB combustors and limestone for the AFB steam generators. The analyses of both sorbents are presented on Table 5.

3.2 POTENTIAL SOURCES OF EMISSIONS FROM FLUIDIZED-BED PROCESSES

The station discharges include solid, liquid, gaseous and thermal emissions that are associated with the following:

(a) The combustion process itself (SO_2, NO_x, particulates, trace elements)

(b) Storage and handling of coal, sorbents and solid wastes.

(c) Solid Waste Disposal

(d) Water Treatment System

(e) Heat Rejection System

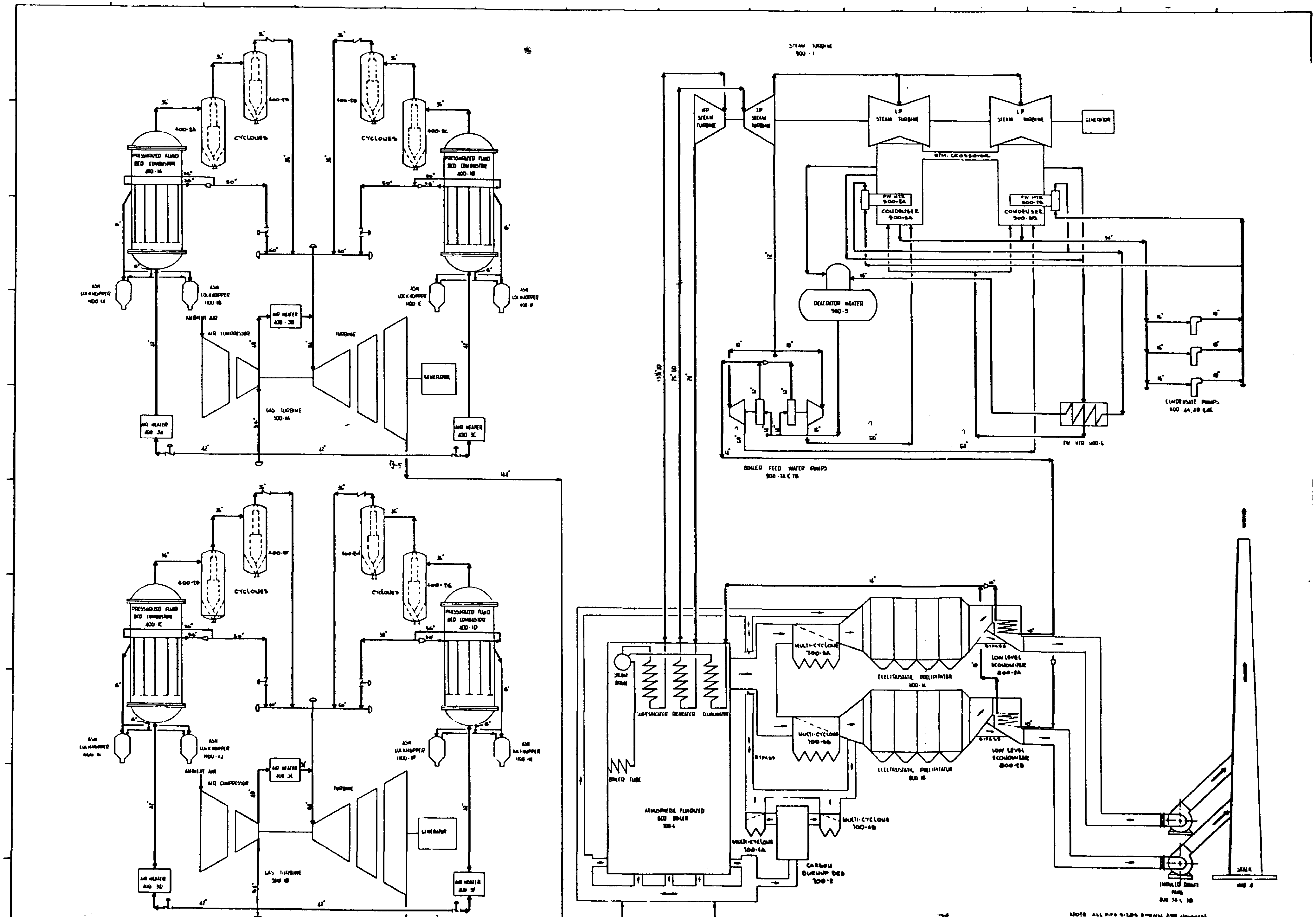

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FIGURE 1

PFBC COMBINED CYCLE POWER PLANT
PROCESS FLOW DIAGRAM

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TABLE 4

FUEL ANALYSIS
AVERAGE OF 82 COALS FROM ILLINOIS BASIN

	% By Wt.-Dry	% By Wt.-As Rec'd	% By Wt.-As Fired
PROXIMATE ANALYSIS:			
-Volatiles	39.78%	35.79%	38.86%
-Fixed Carbon	48.95%	44.05%	47.81%
-Ash	11.27%	10.14%	11.01%
-Moisture	-	10.02%	2.32%
	100.00%	100.00%	100.00%
HHV	12,749 Btu/#	11,472 Btu/#	12,453 BTU/#
ULTIMATE ANALYSIS:			
-C	70.69%	63.61%	69.05%
-S	3.51%	3.16%	3.43%
-H	4.98%	4.48%	4.86%
-N	1.35%	1.21%	1.32%
-O	8.19%	7.37%	8.00%
-Ash	11.28%	10.15%	11.02%
-H_2O	-	10.02%	2.32%
	100.00%	100.00%	100.00%
RANGE OF VARIATION			
-C	62.49% - 79.94%	55.83% - 71.42%	61.04% - 78.09%
-S	1.12% - 5.59%	1.00% - 4.99%	1.09% - 5.46%
-H	4.19% - 5.76%	3.74% - 5.15%	4.09% - 5.63%
-N	0.95% - 1.84%	0.85% - 1.64%	0.93% - 1.80%
-O	4.15% - 14.36%	3.71% - 12.83%	4.05% - 14.03%
-Ash	4.60% - 16.00%	4.11% - 14.30%	4.49% - 15.63%
-H_2O	-	1.60% - 18.20%	0.20% - 2.50%

TABLE 5

SORBENT ANALYSES

DOLOMITE ANALYSIS (For PFBC)

	% By Wgt. (Dry)
$CaCo_3$	53.9%
$MgCO_3$	41.4%
SiO_2	3.1%
Al_2O_3	0.5%
Fe_2O_3	0.8%
Na_2CO_3, K_2CO_3, etc.	0.3%
	100.0%
Moisture	1.0%

LIMESTONE ANALYSIS (For AFBC)

	% By Wgt. (Dry)
$CaCo_3$	97.0%
$MgCO_3$	1.2%
SiO_2	1.1%
Al_2O_3	0.3%
Fe_2O_3	0.2%
Na_2O_3, K_2CO_3, etc.	0.2%
	100.0%

3.2.1. Sources of Emissions from the Combustion Process

The plant stack discharges gaseous products of combustion from the PFB and AFB units, as well as the fines that escape capture by the particulate collection system. Surfur dioxide is generated during the combustion process as a result of the oxidation of the sulfur in the coal. Oxides of nitrogen (NO_x) are generated as a result of high temperatures and excess air in the furnace. There are two nitrogen sources; the bound nitrogen contained in the coal and the free nitrogen in the combustion air. Carbon monoxide and unburned hydrocarbons may be present in the stack gas as a result of incomplete combustion. Particulate loading in the flue gas will depend upon the coal and sorbent feed sizes and operating conditions such as superficial bed velocity, bed temperature and pressure.

Trace elements emitted from coal fired power plants are receiving increased attention as potentially dangerous air pollutants. Typical concentration of elements in representative coals and sorbents are presented in Table 6. The process of coal combustion releases trace elements to the environment as vapors and in association with particulate emissions. Because vaporized trace elements are in the gas phase, they are not captured by the particulate collection devices. A substantial fraction of trace elements present in the coal is retained in the fly ash. There are indications that certain trace elements can concentrate in selected size ranges of particulates. For some elements, such as lead and cadmium, these sizes tend to be less than a few microns in diameter. Such small particles are of environmental concern because they are difficult to remove from the flue gas and, once emitted, they can be readily embedded in the lung.

3.2.2. Sources of Emissions from Solid Material Storage and Transport Systems

The preparation of coal in a power plant involves many steps prior to the actual combustion of the coal. From unloading the railcar, to crushing, to transport to the boiler (combustor), there exists opportunities for fugitive dust or other particulate emissions (which could be a significant fraction of this total power plant particulate emissions. Technology exists to mitigate these emissions if their sources can be identified.

Fluid bed combustors do not require special coal treatment; however, the need for a sulfur acceptor in the bed adds a second potential source of fugitive dust.

The power plant has four sources of solid wastes: (1) spent bed material from the PFB; (2) spent bed material from the AFB; (3) dust captured in the AFB and PFB cyclones; and, (4) dust captured in the electrostatic precipitator. Provision has been made to handle these solid wastes in an environmentally safe manner. A positive-pressure pneumatic transfer system will move the wastes to storage silos. However, when the stored wastes are transferred to railcars or trucks for ultimate disposal in a landfill, there will be some emissions in the form of fugitive dust. Through the use of dust collectors and careful handling, the emissions can be controlled.

TABLE 6

REPRESENTATIVE CONCENTRATION OF ELEMENTS IN COAL AND SORBENTS (ug/g).
(Ref. 3)

Element	Coal	Limestone	Element	Coal	Limestone
Aluminum	12900.00	3998.49	Manganese	25.00	500.00
Antimony	1.30	2.70	Mercury	1.20	0.08
Arsenic	5.00	2.00	Molybdenum	7.50	37.00
Barium	130.00	100.00	Neodymium	6.40	0.00
Beryllium	1.60	0.80	Nickel	21.10	75.00
Bismuth	0.10	0.00	Nitrogen	13800.00	0.00
Boron	102.00	18.00	Phosphorus	71.10	187.00
Bromine	15.00	5.00	Potassium	2300.00	1600.55
Cadmium	2.50	1.00	Praseodymium	76.00	0.00
Calcium	6780.00	374053.62	Rhodium	290.00	0.00
Carbon	730000.00	113711.95	Rubidium	14.00	0.00
Cerium	11.00	0.00	Scandium	3.20	0.00
Cesium	1.00	0.00	Selenium	2.10	0.32
Chlorine	1400.00	55.00	Silicon	30300.00	12713.88
Chromium	20.00	11.00	Silver	0.20	0.00
Cobalt	9.60	100.00	Sodium	1800.00	699.57
Copper	15.00	47.00	Strontium	200.00	490.00
Dysprosium	1.00	0.00	Sulfur	43000.00	0.00
Europium	0.20	0.00	Tantalum	0.40	0.00
Fluorine	61.00	230.00	Tellurium	0.30	0.00
Gallium	3.10	0.00	Terbium	0.23	0.00
Germanium	6.60	0.30	Thorium	2.00	0.00
Hafnium	0.97	0.00	Tin	4.80	40.00
Indium	0.04	0.00	Titanium	700.00	399.27
Iodine	2.78	0.00	Tungsten	1.90	0.00
Iron	21300.00	4295.92	Uranium	1.60	0.00
Lanthanum	6.90	1.60	Vanadium	32.70	16.80
Lead	34.80	30.00	Ytterbium	0.55	0.00
Lithium	25.00	0.00	Zinc	272.30	30.00
Magnesium	500.00	3269.60	Zirconium	180.00	25.50

Because of the storage silo concept for the solid wastes, the only source of water pollution from the solids storage and handling systems would be due to coal and sorbent pile runoff. However, since storage capacity is limited to eight days at full load operation, it is possible that some on-site disposal of spent sorbents and ash may be required during unforeseen events such as transport strikes.

Coal pile runoff is due to the drainage of rain water from the coal storage pile. If allowed to drain into the local water ways, this runoff will be a major source of pollution. There are generally two types of runoff depending upon coal type. One type is neutral or slightly alkaline and contains ferrous ions. This type of runoff originates from alkaline coals with small pyritic content. From coals having large pyritic content, a second type of highly acidic runoff occurs. This runoff contains large amounts of dissolved iron and aluminum. The pyrite is oxidized by the atmospheric oxygen and hydrolyzed to form ferrous sulfate ($FeSO_4$) and sulfuric acid (H_2SO_4). In addition to the solid and liquid emissions, gaseous emissions of hydrocarbons and carbon monoxide may result from spontaneous combustion of reactive coals in the storage piles.

3.2.3 Sources of Emissions from Solid Waste Disposal

From an environmental viewpoint, one of the primary concerns with fluid bed technology is the disposal and/or utilization of the solid wastes generated.

The solid wastes from the fluid bed system are composed of the spent material from the beds and the particulates captured in the particulate collection system. One major problem is the large quantity of material that requires disposal.

Other environmental concerns with FBC solid waste disposal include leaching, dusting, and runoff. Potential problems with FBC leachate include pH, total dissolved solids (TDS), sulfate, and calcium.

3.2.4 Sources of Effluents from Plant Water Treatment Systems

Water usage for any given CCFBC plant design, like that of a conventional pulverized coal plant, is site specific and a direct function of plant generating capacity. Dictated by either thermodynamic principles or water chemistry, the usage rate for any plant process can generally be expressed in terms of percentage of steam generation or water circulating rates.

The cooling tower makeup water is drawn from the North River and accounts for approximately 98 percent of the total plant water requirements. Most of this water requirement is due to evaporative losses which are controlled by the amount of heat rejected from the power generation cycle. The remainder of the requirement is due to cooling tower blowdown.

The AFBC boiler feedwater makeup is the service with the second largest water demand. The rate of water makeup for the subcritical boiler unit is essentially due to steam cycle losses and the boiler blowdown requirements, which, in effect, is controlled by drum water concentrations of dissolved and suspended solids. The degree of solids concentration, dependent on water quality entering the AFBC and steaming rate, will determine the blowdown rate. Make-up boiler feedwater is taken from the city water system.

Waste water is produced as a result of: (1) discharges resulting from use of water in plant systems, and (2) rain water runoff from various plant areas. Facilities are provided to collect and treat this waste water prior to release into the North River. All effluent from the CCFBC plant is in accordance with federal liquid waste discharge limits.

Cooling tower blowdown is one of the major liquid plant waste streams. The blowdown from the cooling tower is metered and continuously monitored for residual chlorine before being returned to the river.

The treatment and demineralization of make-up boiler feedwater generates process waste water. This water, along with the waste water produced in condensate polishing, boiler blowdown, equipment drains, floor drains, oil spills, coal and sorbent pile runoff, etc. (See Table 19) is collected by various piping systems and flows to the central waste water treatment plant for processing prior to discharge into the North River.

3.2.5 Sources of Thermal Discharges

Thermal discharge from the plant occur primarily at the cooling towers and at the stack. In addition, other miscellaneous discharges will occur due to coal drying, solid waste cooling, radiation of plant equipment, water cooling of hot gas valves in the PFB system, etc.

4.0 CHARACTERIZATION, QUANTIFICATION, AND CONTROL OF COMMERCIAL PLANT DISCHARGES

In this section the emissions from the various sources outlined in Section 3.2 are characterized and quantified. In addition, the methods available to control each discharge are discussed.

4.1 COMBUSTION PROCESS EMISSIONS

4.1.1 SO_2 Emissions

It is conservative to assume that all of the sulfur contained in the coal fuel is converted to gaseous SO_2. What distinguishes the fluid bed combustor from a conventional pulverized coal-fired boiler is its ability to accomplish sulfur removal by reacting the SO_2 with a sorbent material in the fluid bed.

The type of sulfur sorbent used and its feed rate are set by the current bed operating conditions and the removal requirements. In order to achieve the current EPA limit of 1.2 lb $SO_2/10^6$ BTU input, approximately 80% of the sulfur in the proposed coal (3.43% sulfur 12453 BTU/lbm HHV) must be removed.

Data from both BCURA and ANL indicate that limestone is not an efficient sulfur sorbent at the operating temperature and pressure considered for the PFB combustors. In both cases, a Ca/S mole ratio of approximately 2.6 was required to achieve 80% retention (Ref. 4, Figure 4.2 and Ref. 5, Figure 7). This occurred despite the fact that the pressure was low enough (5 atm.) for the limestone to calcine in the BCURA work and high enough (8 atm) to prevent calcining in the ANL work.

Dolomite, on the other hand, is a more efficient sulfur sorbent in a PFB combustor. ANL data (Ref. 5, Figures 2 & 7) indicate that a Ca/S ratio of 1 would be required to achieve 80% sulfur retention at the selected bed operating conditions. Therefore, dolomite is used as the sulfur sorbent in the PFB combustor for the commercial plant.

Since the rapid calcination of dolomite under AFB operating conditions causes increased losses of the sorbent from the bed by elutriation, limestone is used as the sulfur capture additive in the AFB steam generator. The limestone feed rate to the AFB was calculated from a relationship between calcium-sulfur ratio, sulfur capture, stone sizing, gas residence time and relative stone reactivity. This relationship was developed by Babcock and Wilcox for another contract. It seems to fit the data reported by PER and ANL as well as data obtained by B & W under an EPRI contract (3'x3' AFB located in B & W's Alliance Research Center) with reasonable accuracy. Using this relationship and the selected AFB bed operating conditions indicates that a calcium-sulfur ratio of 2.5 is required to obtain the SO_2 capture needed to lower the SO_2 emissions from coal combustion in the AFB system to 1.2 lb/10^6 Btu input.

As indicated in the foregoing discussion, the SO_2 emissions from the commercial plant are controlled primarily by variation of the dolomite and limestone feed rates to the PFB and AFB respectively. In addition, variations in gas residence time (i.e., bed height and/or superficial velocity) and sorbent feed size could provide some limited measure of control subject to certain other operational constraints. By using these control methods, the SO_2 emissions from the commercial plant will be maintained at 1.2 lbs $SO_2/10^6$ BTU input.

Concerning the impact of the anticipated changes in EPA limits, it is projected that in order to achieve 90% SO_2 capture the Ca/S feed ratio for the PFB portion of this system must increase from the present 1.0 to 1.5 and that the feed ratio for the AFB portion of the system must increase from the present 2.5 to 3.4. Associated with these increased calcium feed rates will be an increased coal flow to account for the increased heat to the calcining reactions and the increased heat loss in the sensible heat of the spent stone. The coal flow to the PFB must increase by 0.46% and the coal flow to the AFB must increase by 0.68%. Other than the increased solids handling requirements, which remain within the design margins presently incorporated in the systems, the impact of these changes on system design and operation is negligible.

Alternate or supplementary approaches to meeting the new SO_2 rules would involve an increase in the gas residence time within the fluid beds. This would be more effective in the AFB than in the PFB since the residence time in the latter is already relatively high (approximately 7 seconds). An increase in AFB residence time would involve an increase in bed plan area and/or deeper beds. Both approaches would result in a small change in plant capital cost. In addition, deeper beds would cause a higher back pressure on the gas turbine, thereby decreasing its output. In either case, the effect of the changes on the cost of electricity would be relatively small.

4.1.2 Oxides of Nitrogen Emissions

Oxides of nitrogen are generated as a by-product of any combustion process occurring at elevated temperatures in the presence of excess air. The FBC units operate at relatively low combustion temperature (1550-1650°F.), resulting in a lower production of nitrogen oxides than that of conventional coal fired boilers.

The data reported by three investigators (Refs. 4,5 & 6) indicate that NO_x emissions in the range of 0.2 $lb/10^6$ Btu may be expected from the PFB combustor. There appears to be a reasonably good agreement among these three investigators.

As reported in a survey of data from many investigators (Ref. 7) there is a wide range of NO_x emissions measured in AFB combustors. The preponderance of the data is in the range of .3 $lb/10^6$ Btu to .7 $lb/10^6$ Btu and 70% of the reported points lie in the range of .3 $lb/10^6$ to .55 $lb/10^6$ Btu. Little data is available for the NO_x emissions from the CBB but reference 7 sites data to indicate that the emissions are somewhat higher than for the MFB. Because of the wide range of reported data, the NO_x contribution from the AFB is assumed to be the average of the data reported in reference 7, or 0.5 $lb/10^6$ Btu.

Since the coal input is split between the PFB system and the AFB system in the ratio of 0.35/0.65, the projected NO_x emission for the plant is 0.4 $lb/10^6$ Btu. No thermal deceomposition of the PFB system NO_x within the AFB combustor was assumed in this prediction.

Based on the foregoing analyses, it is expected that the commercial plant would meet the anticipated EPA NO_x emission limits of 0.6 lb/10^6 BTU without any further modification in the operating conditions or plant design.

4.1.3 Carbon Monoxide and Hydrocarbon Emissions

In addition to SO_2 and NO_x emissions, carbon monoxide and hydrocarbon emissions must also be considered.

Currently there are no regulations concerning CO emissions from boilers. The CO emissions from a conventional boiler are typically less than 100 ppm. The CO emission levels from the pressurized fluid bed combustion process are projected to be slightly higher. Exxon, for instance, has reported[8] measured CO levels of 150-250 ppm with excess air ranging from 15 to 60%.

The measured CO levels in atmospheric fluid bed combustors are, however, reported as an order of magnitude higher than for the pressurized fluid bed combustor. Battelle[9] for instance, has reported CO emissions in excess of 2000 ppm with a bed temperature of 1650F. The CBB operates at 2000F which should result in more complete combustion and thus decreased CO formation. Tests by Pope, Evans and Robbins indicate that low CO emissions can be expected from the CBB.

The cause of the high CO emissions from the MFB of the atmospheric pressure fluid bed combustion system has not been fully identified. During 1978 tests will be conducted at the EPRI sponsored 6' X 6' AFB test facility located at B&W's Alliance Research Center. These tests are intended to better identify the CO formation and develop means of minimizing the CO emissions from the AFB process.

The PFB/AFB combined cycle studied in Subtask 1.2 should have lower CO emissions than a cycle utiliziang only an AFB combustion system. However, even the combined cycle could be expected to have CO emissions an order of magnitude higher than for conventional power plants. Future developments in the AFB process may point the way to reducing these emissions.

Very little data is available relative to hydrocarbon emissions from either a PFB or AFB process. As reported in Reference 8 the hydrocarbon emissions from one set of AFB experiments was about 100 ppm at approximately 20% excess air. It may be hypothesized that the hydrocarbon emissions are due to the same factors causing the CO emissions. The hydrocarbon emissions from the PFB process would then be expected to be considerably lower than for the AFB process.

In any case the hydrocarbon emissions are greater than those of a conventional boiler which emits negligible hydrocarbons (less than 5 ppm for C_1 through C_6 compounds). While the hydrocarbon emissions of the fluid bed process are greater than desired it is too early to draw any conclusions with respect to environmental impact.

4.1.4 Particulate Emissions

The commercial plant has been designed to meet the current EPA requirements for emission of particulates from the stack. At the present time, empirical information regarding the particle size distribution of the solids elutriated from a PFB combustor is very sparse. Consequently, assumptions were made in order to establish the size distribution of the particulates entering the gas cleanup equipment.

4.1.4.1 PFB Particulate Emissions and Control

The size distribution of the sulfur sorbent elutriated from the bed was based on the size distribution of the stone fed to the bed. To account for abrasion and thermal decrepitation in the bed, a 20% reduction in this size distribution was assumed; the resulting size distribution is shown on Figure 2. Terminal settling velocity analysis indicated that particles less than 300 micron size would be carried out of the PFB combustor. This results in an elutriation rate of 35% for the spent sorbent. It should be noted that based on these assumptions, less than 1/2% of the elutriated spent sorbent will have a size of less than 10 microns.

The size distribution of the coal ash was assumed to be the same as the fly ash size distribution leaving a pulverized coal or stoker fired boiler (Figure 2). This assumption is equivalent to essentially all the coal ash being elutriated from the bed with nearly 40% being less than 10 micron size.

In designing the PFB particulate removal system two conditions were specified. The expected operating conditions are based on a Ca/S molar feed ratio of 1.0. The design conditions for material handling equipment are based on a Ca/S ratio of 3.0. These conditions are shown in Tables 7 and 8 with the corresponding size distribution being shown in Figure 2. Because of the assumptions for the size distribution of the spent sorbent, the dust in the less than 10 micron size range is essentially all coal ash and therefore the dust loading in this size range is the same for both conditions. The higher Ca/S ratio then primarily influences the design of the initial separator stages and the design of the ash let down system.

Since the primary function of the PFB particulate removal equipment is to protect the gas turbine rather than the environment, the permissible level leaving the equipment is based on gas turbine tolerance levels rather than environmental considerations.

Because of a lack of actual operating experience with PFB exhaust gases in a gas turbine, the allowable level of particulate concentration in the gas entering the turbine has not been demonstrated. On the basis of limited data (Ref. 10), an estimate of allowable gas turbine particulate loading has been made and is shown below:

particle diameter, d (microns)	max. particulate concentration (grains/SCF)
$d \leq 2.0$	no limit
$2.0 \leq d \leq 10.0$	0.0100
$d > 10.0$	0.0000

In addition to the relatively stringent collection efficiency requirements, the equipment used in this application must also operate at 1600°F. and at a pressure of 10 atm.

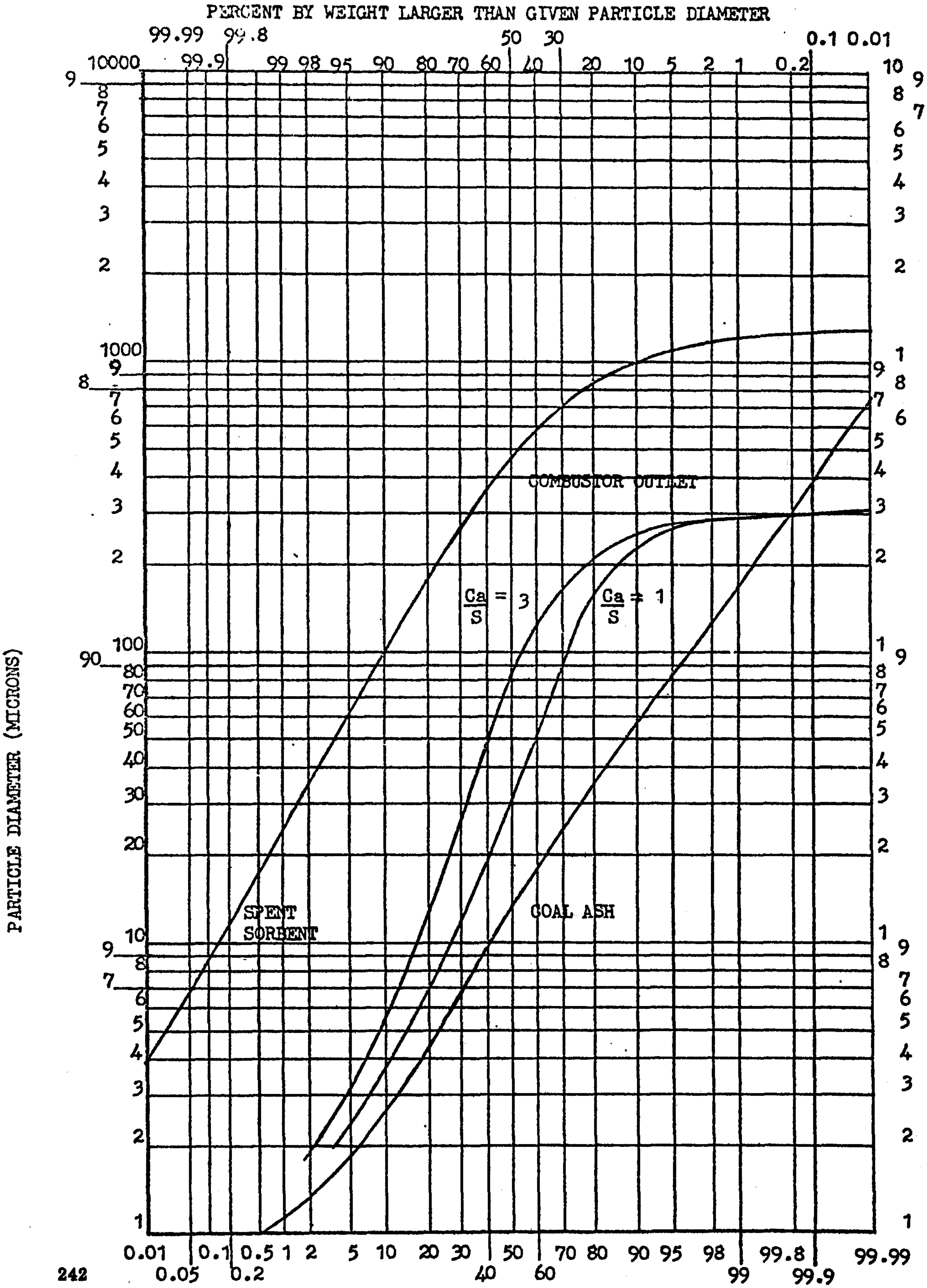


Figure 2. PARTICLE SIZE DISTRIBUTION ENTERING THE GAS CLEANUP EQUIPMENT

TABLE 7

PFB PARTICULATE COLLECTION SYSTEM
OPERATING CONDITIONS FOR Ca/S=1.0

Collector Inlet Gas Analysis

Component	lbm/hr	moles/hr
O	11344.	354.35
N	282976.	10101.
Ar	5055.5	126.7
SO	475.5	7.45
CO	89678.	2038.
H O	16002.	888.
Total	405531.	13515.5

Collector Inlet Gas Molecular Weight	30.005 lbm/mole
Collector Inlet Gas Temperature	1650 °F
Collector Inlet Gas Pressure	136.0 psia
Collector Inlet Gas Density	.1803 lbm/ft
Collector Inlet Dust Flow	6474 lbm/hr
Collector Inlet Dust Concentration	.01596 lbm dust/lbm wet gas

Collector Inlet Particle Size Distribution

particle diameter (microns)	% by weight stated particle diameter
100	27.46
80	31.53
60	36.18
40	43.27
20	57.75
10	73.02
8	76.90
6	82.06
4	88.48
2	96.16

Clean Compressed Air Flow	1085508 lbm/hr
Clean Compressed Air Temperature	1578 °F
Clean Compressed Air Pressure	136.0 psia
Clean Compressed Air Density	.1803 lbm/ft

TABLE 8

PFB PARTICULATE COLLECTION SYSTEM
DESIGN CONDITIONS FOR Ca/S = 3.0

Collector Inlet Gas Analysis

Component	lbm/hr	moles/hr
O	11344.	354.35
N	282976.	10101.
Ar	5055.5	126.7
SO	475.5	7.45
CO	92661.	2105.5
H O	16144.	896.1
Total	408656.0	13591.1

Collector Inlet Gas Molecular Weight	30.068 lbm/mole
Collector Inlet Gas Temperature	1650^{o}F
Collector Inlet Gas Pressure	136.0 psia
Collector Inlet Gas Density	.1807 lbm/ft
Collector Inlet Dust Flow	10372.35 lbm/hr
Collector Inlet Dust Concentration	.02538 lbm dust/lbm wet gas

Collector Inlet Particle Size Distribution

particle diameter (microns)	% by weight stated particle diameter
100	43.77
80	49.33
60	54.94
40	61.78
20	72.90
10	83.01
8	85.50
6	88.76
4	92.80
2	97.61

Clean Compressed Air Flow	1085508 lbm/hr
Clean Compressed Air Temperature	1578^{o}F
Clean Compressed Air Pressure	136.0 psia
Clean Compressed Air Density	.1803 lbm/ft

As part of the Trade-Off Studies performed under Subtask 1.3, various particulate control systems, including cyclones and granular bed filters, have been investigated. On the basis of predicted performance, system cost and projected operating reliability, Aerodyne Development Corporation's "SV-FBC" Series Dust Collector is used in the conceptual plant design. The particular "SV-FBC" Dust Collector used is the Model 22000 SV as shown in Figure 3.

The design is an extension of the equipment presently used in low temperature, low pressure operations. The modifications include placing the collector within a refractory lined pressure vessel. This vessel also serves as an initial stage cyclone collector so that the collector is actually a two stage device with the second stage based on the existing Aerodyne Series "SV" Dust Collector (License: System Siemens).

The predicted collection efficiency is shown on Figure 4. Calculations indicated that two of these collectors operating in series are required for each PFB combustor. The predicted performance is shown in Table 9 for a Ca/S ratio of 1.0 and in Table 10 for a Ca/S ratio of 3.0. The dust loading entering the gas turbine in the critical 2 to 10 micron size range is projected to be 1/3 of the allowable level.

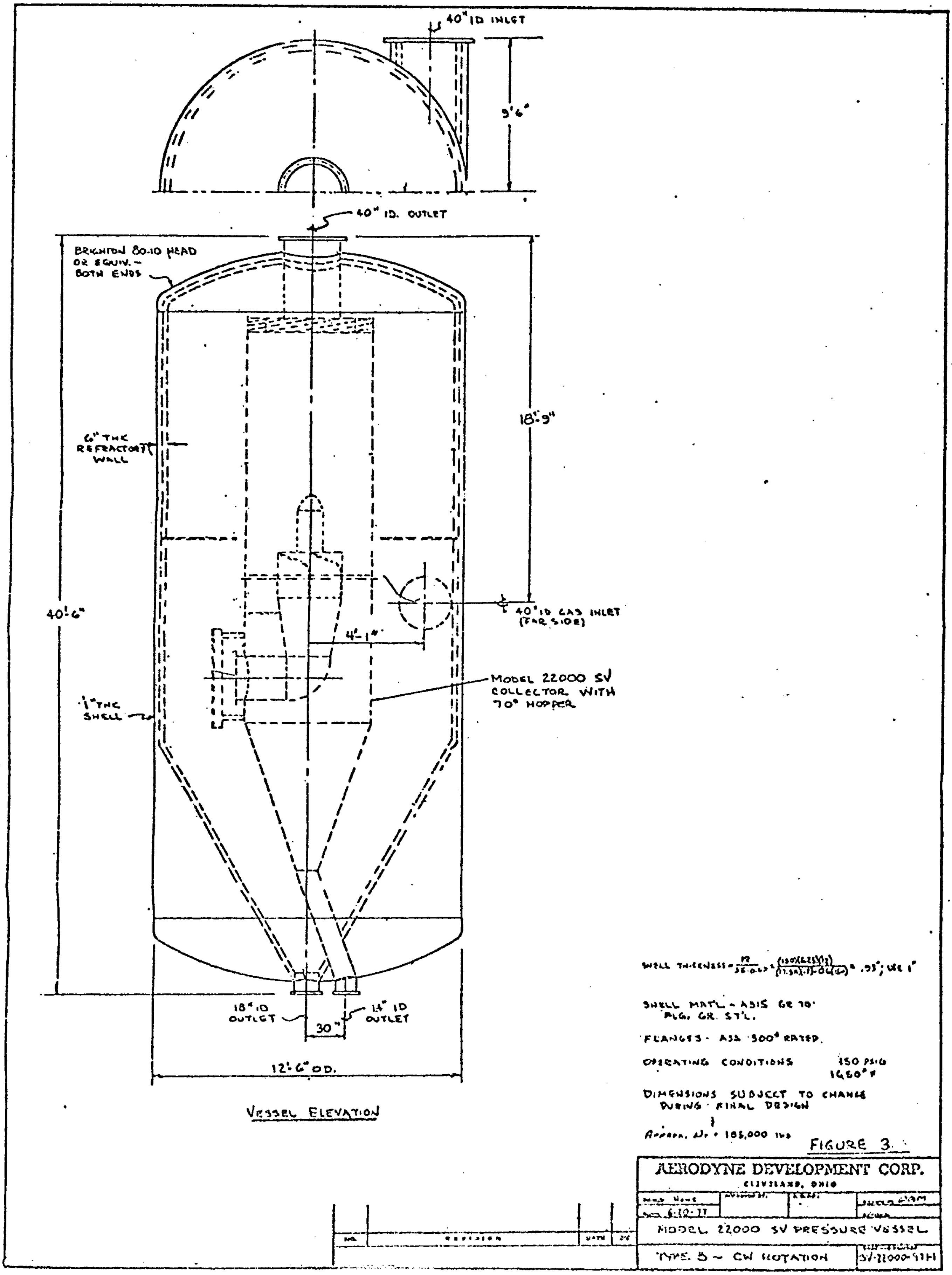
4.1.4.2 AFB Particulate Emissions and Control

As in the case of the PFB combustor, little data is available concerning the material size distribution for the various particles elutriated from the bed; therefore; assumptions were made. The spent sorbent and coal ash were treated in a manner similar to the analysis of the PFB particulate removal system. In addition, assumptions were made for the sizing of the char based upon the coal feed sizing. These assumed size distributions are shown in Table 11.

Two dust collector systems are incorporated into the AFB system, one for the Main Fluid Bed Unit (MFB) and one for the Carbon Burn-up Unit (CBB).

The primary function of the dust collectors following the MFB is to capture the char (unburned coal) elutriated from the MFB unit. This captured char is combusted in the CBB to improve the system combustion efficiency. The primary function of the dust collector following the CBB is to reduce the total solids loading of the gases entering the electrostatic precipitator which is used for final particulate removal to comply with environmental requirements.

Conventional multiclone dust collectors are used for both systems. A number of vendors offer similar equipment; however, the equipment of the Air Correction Division of UOP, Inc. is used in the conceptual design. To achieve a high char collection efficiency the MFB dust collector uses 6" diameter tubes. The projected char collection efficiency for the assumed size distribution is 90% while the corresponding efficiency for the spent sorbent and coal ash are 99+% and 75% respectively. The same high collection efficiency is not required for the CBB dust collector; therefore, a less expensive collector utilizing 10" diameter tubes is provided. The collection efficiencies for



40" ID INLET
9'6"
40" ID. OUTLET
BRIGHTON 80-10 HEAD OR EQUIV. — BOTH ENDS
18'-9"
6" THK REFRACTORY WALL
40'-6"
40" ID GAS INLET (FAR SIDE)
4'-1"
MODEL 22000 SV COLLECTOR WITH 70° HOPPER
1" THK SHELL
18" ID OUTLET
14" ID OUTLET
30"
12'-6" OD.
VESSEL ELEVATION
SHELL MATL - A515 GR 70 FLG. GR. ST'L.
FLANGES - ASA 300# RATED.
OPERATING CONDITIONS 150 PSIG 1650°F
DIMENSIONS SUBJECT TO CHANGE DURING FINAL DESIGN
AERODYNE DEVELOPMENT CORP.
CLEVELAND, OHIO
MODEL 22000 SV PRESSURE VESSEL
TYPE 3 ~ CW ROTATION
SV-22000-97H

FIGURE 3

246

AERODYNE SV-FBC SEPARATOR

EFFICIENCY CURVES

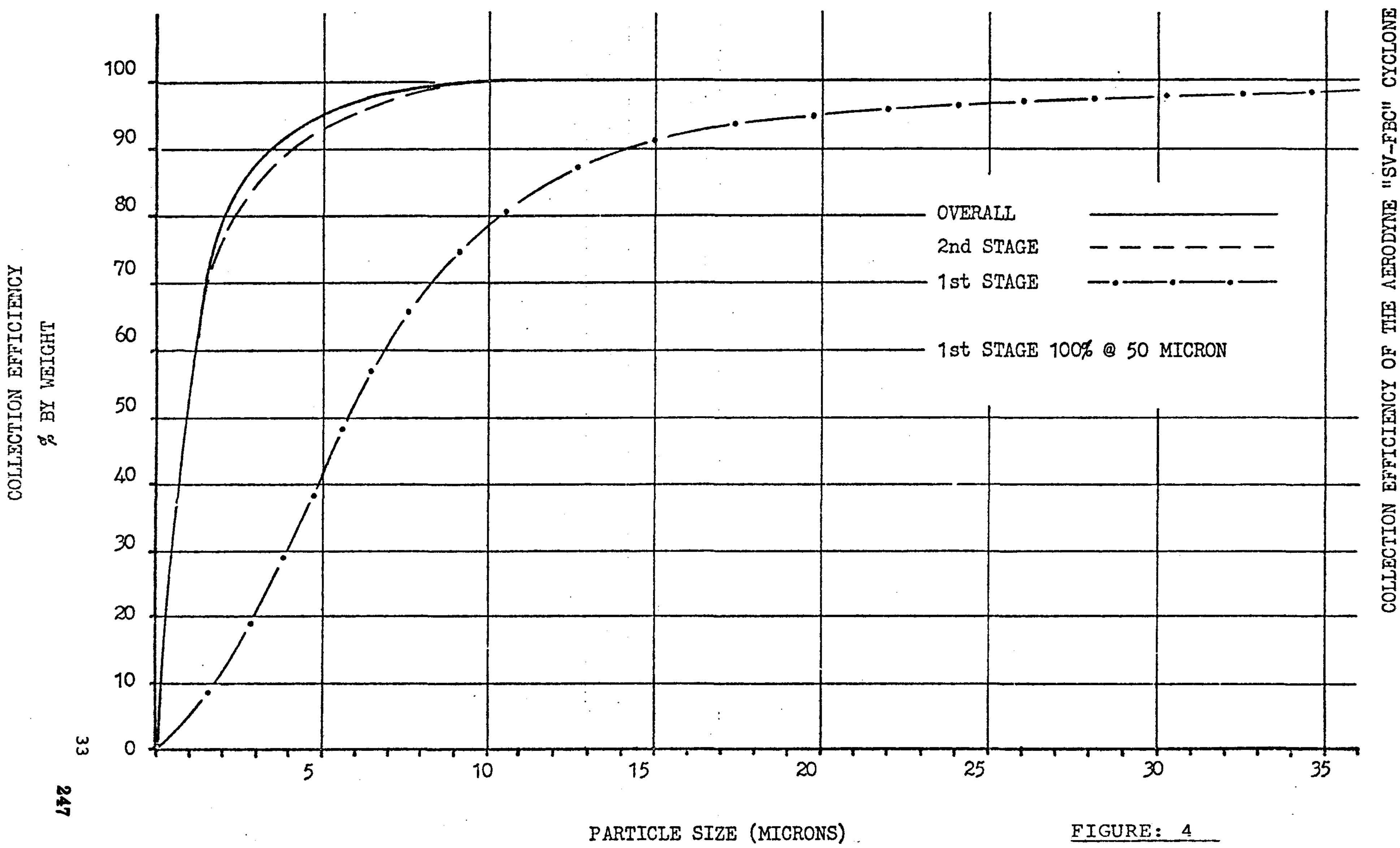


FIGURE: 4

TABLE 9

AERODYNE PARTICULATE COLLECTION SYSTEM
PREDICTED PERFORMANCE FOR Ca/S = 1.0

Dust flow entering the first collector = 6474 lbm/hr

	First Collector	Second Collector
Particulates removed in first stage (lbm/hr)	5291.5	27.68
Particulates removed in second stage (lbm/hr)	991.4	107.71
Dust flow leaving the collector (lbm/hr)	191.1	55.71

Dust entering the turbine 0.0207 grains/SCF

*particle distribution entering the turbine:

Particle diameter, d (microns)	particle concentration (grains/SCF)
d<2.0	0.0176
2.0<d<10.0	0.0031
d>10.0	0.0000

*This flow rate is for each PFB combustor.

TABLE 10

AERODYNE PARTICULATE COLLECTION SYSTEM
PREDICTED PERFORMANCE FOR Ca/S = 3.0

Dust flow entering the first collector = 10372.35 lbm/hr

	First Collector	Second Collector
Particulates removed in first stage (lbm/hr)	9172.75	27.81
Particulates removed in second stage (lbm/hr)	1008.34	107.74
Dust flow leaving the collector (lbm/hr)	191.26	55.71

Dust entering the turbine 0.0207 grains/SCF

*particle distribution entering the turbine:

particle diameter, d (microns)	particle concentration (grains/SCF)
$d<2.0$	0.0176
$2.0<d<10.0$	0.0031
$d>10.0$	0.0000

* This flow rate is for each PFB combustor.

TABLE 11

TABULATION OF AFB DUST LOADING CONTRIBUTION TO EP
(Excludes dust in turbine exhaust gas)

Constituents	Particle Size (Micron)	Particle Size Apparent (Micron) @SG=2.5*	% In Inlet Gas Stream %	Wt. in Inlet Gas Stream (Lb/Hr) Main	Wt. in Inlet Gas Stream (Lb/Hr) CBB	% Removal % Main 3.4"wg	% Removal % CBB 3.5"wg	% of Original in Exit Main	% of Original in Exit CBB	Wt. in Exit Gas (Lb/Hr) Main	Wt. in Exit Gas (Lb/Hr) CBB
CHAR											
	+60	37.9	60	23927	2735	100	100	0	0	0	0
	-60 +40	31.6	3	1196	137	100	99	0	1	0	1
	-40 +30	22.1	2	798	91	100	97	0	3	0	3
	-30 +20	15.8	2	798	91	100	93	0	7	0	6
	-20 +15	11.1	2	798	91	98	83	2	17	16	15
	-15 +10	7.9	3	1196	137	94	73	6	27	72	37
	-10 +7.5	5.5	3	1196	137	87	59	13	41	155	56
	-7.5	2.4	25	9970	1140	60	33	40	67	3988	764
										4231	882
SPENT SORBENT											
	+60	62.9	93.2	19533	20520	100	100	0	0	0	0
	-60 +40	52.4	3.1	650	682	100	100	0	0	0	0
	-40 +30	36.7	1.3	272	286	100	99.5	0	0.5	0	1
	-30 +20	26.2	1.15	241	253	99.8	98	0.2	2	.5	5
	-20 +15	18.3	.47	99	103	99.5	95	0.5	5	.5	5
	-15 +10	13.1	.74	155	163	99	88	1.0	12	1.5	20
	-10	5.2	.04	8	9	84	56	16	44	1.3	4
										2.8	35
ASH											
	+40	40	52	10422	11831	100	100	0	0	0	0
	-40 +20	30	8	1603	1820	100	99	0	1	0	18
	-20 +10	15	7	1403	1593	99	91.5	1	8.5	14	135
	-10 +4	7	5	1002	1138	92	68	8	32	80	364
	-4 +2	3	4	802	910	67	39	33	61	265	555
	-2 +1	1.5	5	1002	1138	40	25	60	75	600	853
	-1	0.5	19	3808	4323	14	7	86	93	3265	4020
										4224	5945

*Apparent size = average size x $\left(\frac{SG}{2.5}\right)^{1/2}$ SG = specific Gravity the char, spent sorbent and coal ash are projected to be 80%, 99+% and 75% respectively. The collection efficiency curves for these two collectors are shown on Figures 5 and 6 and the projected performance and dust flow to the electrostatic precipitator (EP) are shown in Table 11.

4.1.4.3 Final Particulate Collection State

The flue gas from the AFB boiler after passing through high efficiency multiclones goes through a high temperature electrostatic precipitator (EP).

The electrostatic precipitator is designed for a maximum temperature of 800°F. The total volume of flue gas handled by the EP is 3.4 x 10^6 ACFM. The EP has four electric fields in series. The total particulates emission to the plant stack is 0.09 lb per million BTU of heat input.

In order to reduce the emissions to the level of 0.03 lb/10^6 Btu as per the anticipated future requirements, a dust collection efficiency of over 99% would be required on the part of the final collection device. While this is within the capability of currently available collection equipment some changes to the equipment included in this conceptual plant might be required to achieve the higher efficiency. However, it is expected that the effect of these changes on the estimated cost of the plant would be insignificant.

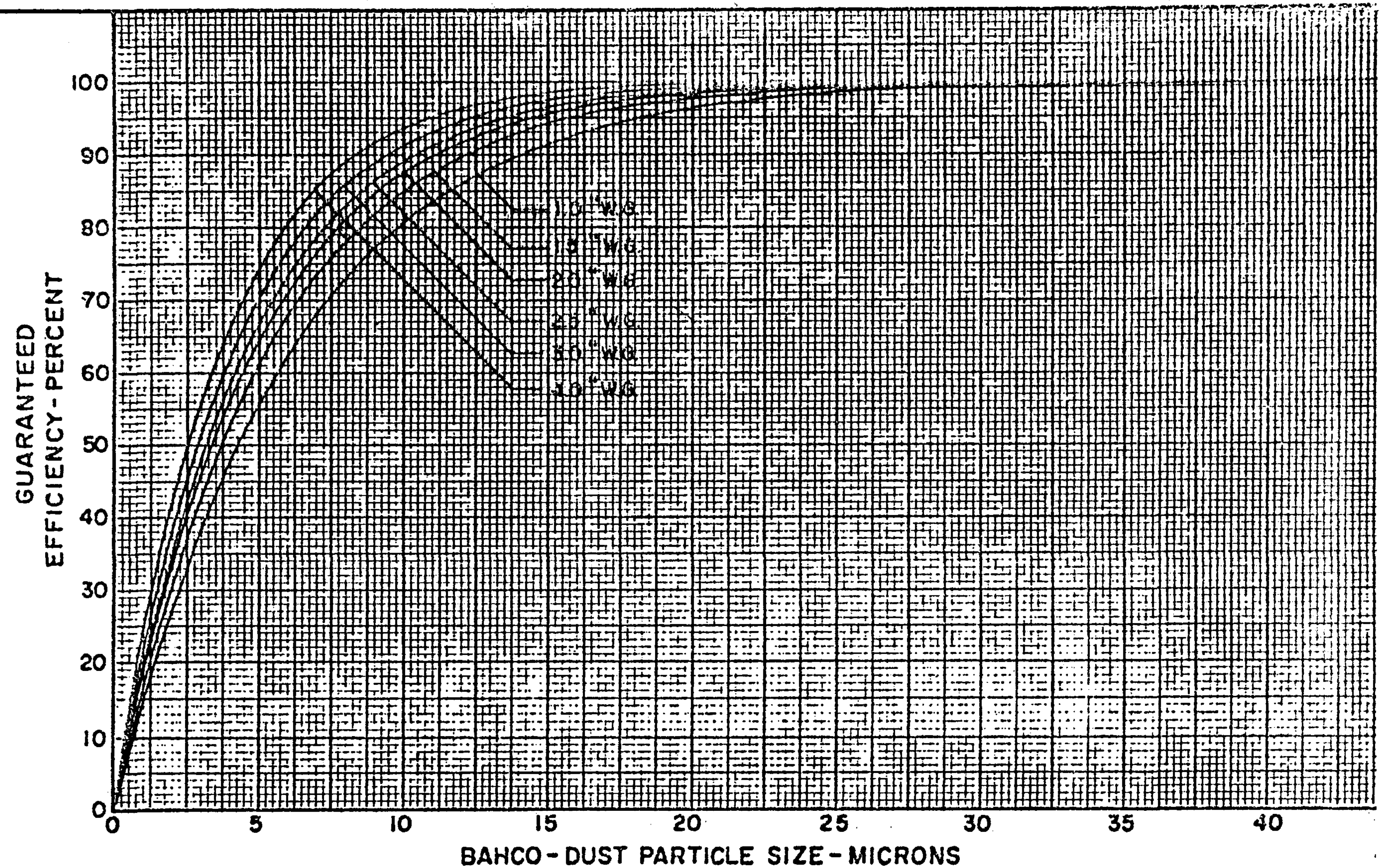
4.1.5 Trace Elements Emissions

As indicated in section 3.2.1, the process of coal combustion releases trace elements to the environment as vapors and also in association with particulate emissions that are sufficiently small to escape the particulate collection devices. For instance, there are indications that certain trace elements such as lead and cadmium, concentrate on particles that are less than a few microns in diameter. Such small particles are of environmental concern because they are difficult to remove from the flue gas and, once emitted, they can be readily embedded in the lung.

To better understand the fate of trace elements in a fluid bed combustor, it has been suggested [11] that these elements be classified into four geo-chemical groups. The four groups are: (I) lithophile, (II) chalcophile, (III) volatile elements and (IV) unclassified elements exhibiting the properties of either Class I or Class II. This classification is shown in the following table.

THE SEPARATION OF ELEMENTS IN THE GEOCHEMICAL CLASSIFICATION SCHEME

Class I		Class II	Class III	Class IV
Al	Mn	As	Hg	CR
Ba	Rb	Cd	Cl	Cs
Ca	Sc	Cu	Br	Na
Ce	Si	Ga	F	Ni
Co	SM	Pb		U
Eu	Sr	Sb		V
Fe	Tu	Se		
Hf	Th	Zn		
K	Ti			
La				
Mg				



Dust Collector For
Main Fluid Bed

FIGURE: 5

CONDITIONS OF CURVE

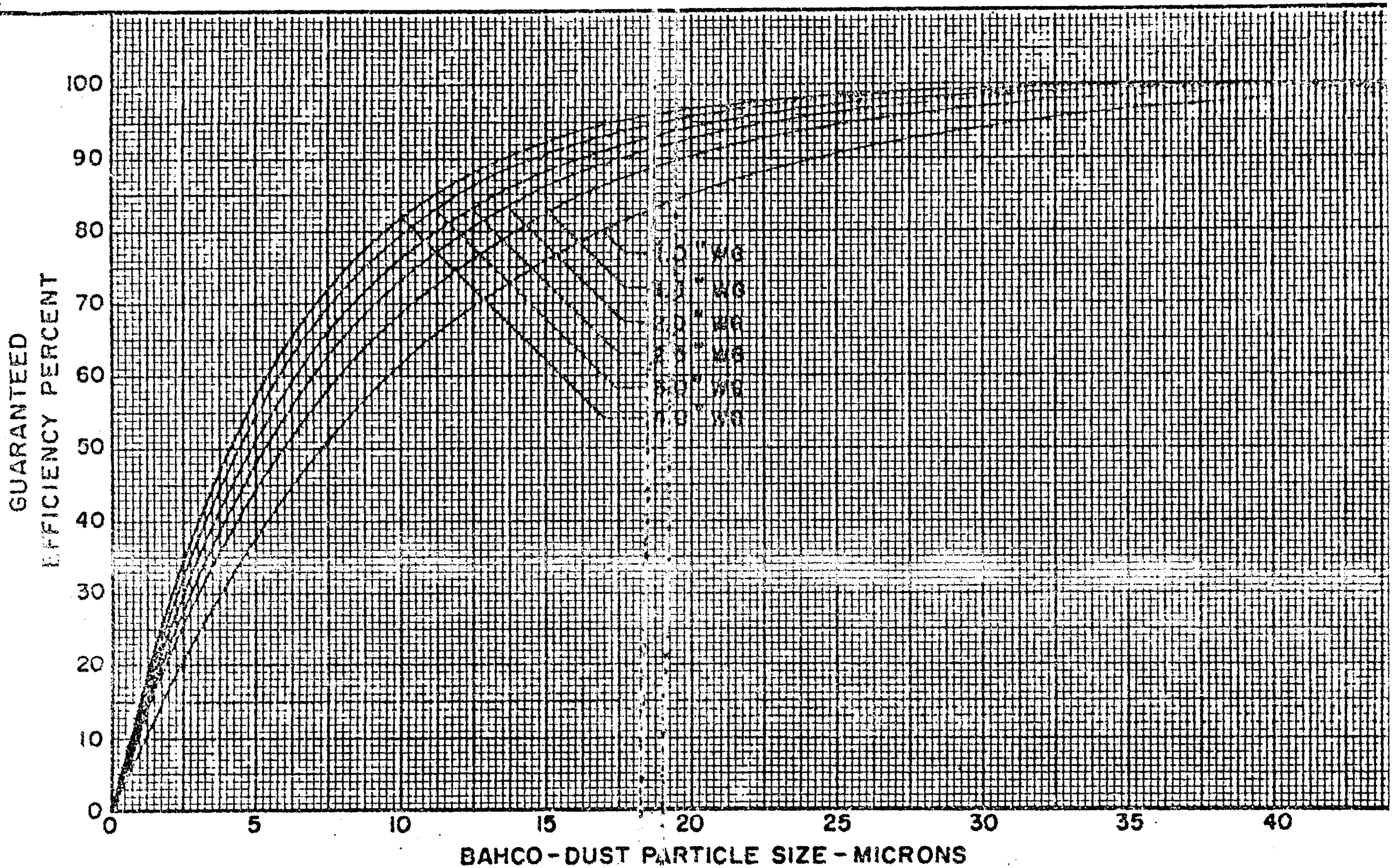
DUST SPECIFIC GRAVITY	-	2.5
DUST CONCENTRATION	-	2 TO 5 GR./CU.F
GAS TEMPERATURE	-	70-600 °F

Air Correction Division
Universal Oil Products Company
Tokeneke Road • Darien, Connecticut 06820 203-655-8711
UOP

MICRON EFFICIENCY CURVES
TUBULAR DUST COLLECTOR
DESIGN 106 OR 106A

COMPUTED BY	DATE	APPROVED BY	DATE	APPROVED BY	DATE

39



253 Dust Collector For
Carbon Burn-Up Bed

FIGURE: 6

CONDITIONS OF CURVE

DUST SPECIFIC GRAVITY - 2.5
DUST CONCENTRATION - 2 TO 5 GR./CU.F
GAS TEMPERATURE - 70 - 800 °F

Air Correction Division
Universal Oil Products Company
Tokeneke Road · Darien, Connecticut 06820 203·655-8711
UOP

MICRON EFFICIENCY CURVES
TUBULAR DUST COLLECTOR
DESIGN 104B OR 105A

COMPUTED BY	DATE	APPROVED BY	DATE	APPROVED BY	DATE

Trace elements listed in Class I are lithophiles and are associated with aluminosilicate minerals in coal. As such, they are high boiling compounds and do not decompose on combustion. They usually melt and coalesce to form fly ash and slag. Elements in this class are not enriched during combustion.

Class II elements are generally present in coal as sulfides. These sulfides themselves may be fairly volatile or, upon combustion, decompose and the elements themselves are produced in the vapor phase. These volatile sulfides or elements then condense on the extensive surface area presented by particulates thus leading to a surface enrichment. This enrichment is usually most prevalent in the fine particle fraction (i$<$ 3 um) of the total particulate loading." Generally, elements could be placed in Class II if:

$$\frac{\text{Wt \% of constituent in fly ash}}{\text{Wt \% of constituent in coal}} > 3$$

Class III elements boil below the furnace and flue gas temperatures and exit from the stack as vapors.

Of the Class IV elements, only Cr and Ni tend to show chalcophile (or volatile) characteristics.

A study of the factors affecting trace element emissions indicates that fluid bed combustion might emit lower concentrations of these elements than a conventional boiler. The potential reduction of trace element emissions is related to the bed temperature, coal size and the fact that a sulfur sorbent is used. The low fluid bed temperature should oxidize or volatilize fewer trace elements than a PC boiler. With the use of sulfur sorbent, the fluid bed system has the advantage of a sorbent that acts like a sink for certain trace metals, such as lead.

The preceding discussion indicated that the elements of Class I should not be enriched or volatilized during fluid bed combustion. Therefore, the worst case analysis would assume that these trace elements are elutriated from the fluid bed combustors and that their emission levels are governed by the performance of the particulate control devices. The trace element emissions approaching the particulate control devices for the fluid bed systems are higher than for the conventional boiler since, in the latter case, a portion of these elements is contained in the coal ash slag.

Trace element emissions for elements in Classes II, III, and IV are more difficult to predict because of their volatility, which could lead to vapor phase emissions or to enrichment of fine particulates which are inefficiently collected. Indeed, these elements may pose a problem for some PFB systems. The ideal PFB system would include only a hot particulate collection system intended to both meet emission standards and protect the gas turbine. Some of these trace elements may be vapors when passing through the hot particulate collection system and, hence, they will not be removed from the gas. Any PFB system, regardless of cycle configuration, may require some type of relatively cool, atmospheric pressure particulate collection system to capture these elements after they have condensed. The commercial plant concept studied in Subtask 1.2 of this contract already includes such a system.

In summary, it must be noted that the amount of trace elements that ultimately escape any combustion process is greatly dependent on the performance of the final particulate system and on combustion temperature. Until the performance of electrostatic precipitators and other collection devices is established, little can be done except to speculate that elements will leave in the stack gas.

While there is some uncertainty concerning the performance of particulate control devices such as fabric filters and electrostatic precipitators on full scale fluidized bed systems, it seems likely that such devices can be used to maintain the emission of trace element bearing particulates at a level which is equivalent to a conventional coal-fired plant. The emission of vaporized trace elements should be lower in fluid bed systems due to the lower combustion temperatures that exist relative to the conventional plant. Therefore, it is expected that the PFB/AFB combined cycle plant will have a lower total trace element emission level than a conventional plant. A comparison of projected stack gas emissions of selected trace elements from conventional and fluidized-bed combustors is presented in Table 12. It is evident from this data that FBC units have good potential for reducing trace element emissions.

However, there is little definitive data available about the trace element emissions from conventional boilers or fluid bed units. Hence, definitive field measurements from both complete FBC systems and conventional power plants must be made before any final conclusions can be drawn concerning the relative merits of the FBC systems with respect to trace element emissions.

4.2 EMISSIONS FROM SOLID MATERIAL STORAGE AND TRANSPORT SYSTEMS

Detailed descriptions of the coal and sorbent storage and preparation systems are given in Reference 2. Briefly, coal is unloaded from unit trains and sent to storage. Upon reclamation from storage, the coal is crushed and dried prior to being sent to the feed bins for the fluid bed combustors. Similarly, the sorbent (limestone/dolomite) is unloaded from rail cars, stacked for storage, reclaimed, crushed and sent to feed bins. During these handling steps, there is an opportunity for emissions of coal and stone dust, carbon monoxide, and hydrocarbon gases.

4.2.1 Fugitive Dust Emissions

4.2.1.1 Dust Emissions from Coal and Sorbent Handling Systems

The handling and preparation of coal result in atmospheric discharges of particulates which significantly contribute to the total emissions of the overall power plants. For example, the nominal 600-MW PFB/AFB combined cycle power plant which meets the EPA new source particulate standard (0.1 lb particulate/10^6 Btu coal burned) has a stack emission of approximately 600 lb/hr. The particulate emissions from the storage, handling and drying of the coal necessary to fire this plant adds an additional 142 lb/hr or about 20 percent of the stack emissions. Assuming that the sorbent handling emits approximately the same percentage dust, fines, etc. as the coal, a further 25 lb/hr

TABLE 12

PROJECTED STACK EMISSIONS OF SELECTED TRACE ELEMENTS FROM CONVENTIONAL AND FLUIDIZED-BED COMBUSTORS EXPRESSED AS A PERCENTAGE OF THE ELEMENT ENTERING THE SYSTEM (Ref. 12)

Element	Conventional Combustion[(a)]	Fluidized-Bed Combustion ANL[(a)]	Exxon[(b)]	BCL[(c)]
Mercury	90	80	No Data	75
Fluorine	90-100 (estimated)	40	No Data	98
Bromine	100 (estimated)	65-82	79	90
Arsenic	50-60	15	14	59
Lead	0-60	0-20	No Data	21
Beryllium	No Data	20-40	No Data	98
Scandium	10	0-3	15	0
Chromium	0	25	0	(d)
Cobalt	10-20	0-20	No Data	(d)
Sodium	20	4-5	12	(d)
Potassium	30	0-10	25	25-54
Iron	0	0	20	(d)
Manganese	0	0	4	(d)

(a) Source: Argonne National Laboratories
(b) Source: Exxon Research and Engineering
(c) Source: Battelle-Columbus Laboratories Spark Source Mass Spectometer Data (SSMS)
(d) Data Suspect Due to Accuracy Limitations of SSMS

is emitted. The sources of the emissions, estimates of their magnitude, potential control technologies and an estimate of the controlled emissions are summarized in Fig. 7.

In addition to the control technologies listed on Figure 7, other methods are used to minimize the fugitive dust emissions. For instance, to control the release of dust from the storage area, the coal is stockpiled in successive layers not more than one-foot thick and compacted to eliminate air spaces. To minimize the loss of fines due to the wind, the top and sides of the coal storage piles are compacted with stack-size coal with the sides of the piles having a shallow slope. The storage pile is sprayed with commercial products available for control of dusting. The top and sides of the piles are periodically trimmed and the tops of the piles recompacted. Dust suppression equipment is provided at the silos, feeders, bunkers, and all transfer points. Similar measures are used for sorbent storage.

4.2.1.2 Dust Emissions from Solid Waste Handling and Storage Systems

The power plant has various sources of solid wastes which are identified in Section 3.2.2. Provision has been made to handle the approximately 68 ton/hr of solid wastes in an environmentally acceptable manner. A positive-pressure pneumatic transfer system moves the wastes to storage silos. There will be some emissions of fugitive dust due to imperfect separation of solids and transport air at the silos. In addition, some dust will escape when the stored wastes are transferred to railcars or trucks for ultimate disposal in a landfill. Through the use of dust collectors (fabric filters) and dust suppression sprays, the emissions can be maintained at 0.1 percent of the total load or approximately 136 lb/hr.

4.2.1.3 Total Fugitive Dust Emissions

As shown in Table 13, the total particulate emissions to the air due to coal and stone storage and handling and solid waste disposal are estimated to be 303 lb/hr, or approximately 50 percent of the allowable stack emissions.

4.2.2 Water Pollutant Emissions

When rain falls on coal piles, certain elements are washed out and become part of the runoff. The runoff characteristics can differ widely depending upon the coal type, local weather and contact time between the coal and water. Characteristics of runoff at seven different power plants are given in Table 14. Using national averages for coal storage areas and rainfall, a value of approximately 40,000 gal/yr-MW can be derived for coal pile runoff. Applied to Table 14, this would mean that 1 mg/liter would be roughly equivalent to 0.33 lb/yr-MW. In the commercial PFB/AFB plant, all storage pile runoff is routed to the in-plant wastewater treatment facility for processing prior to discharge. This facility is described in Section 4.5.

FIGURE 7

PARTICULATE EMISSIONS FROM COAL PREPARATION AND HANDLING

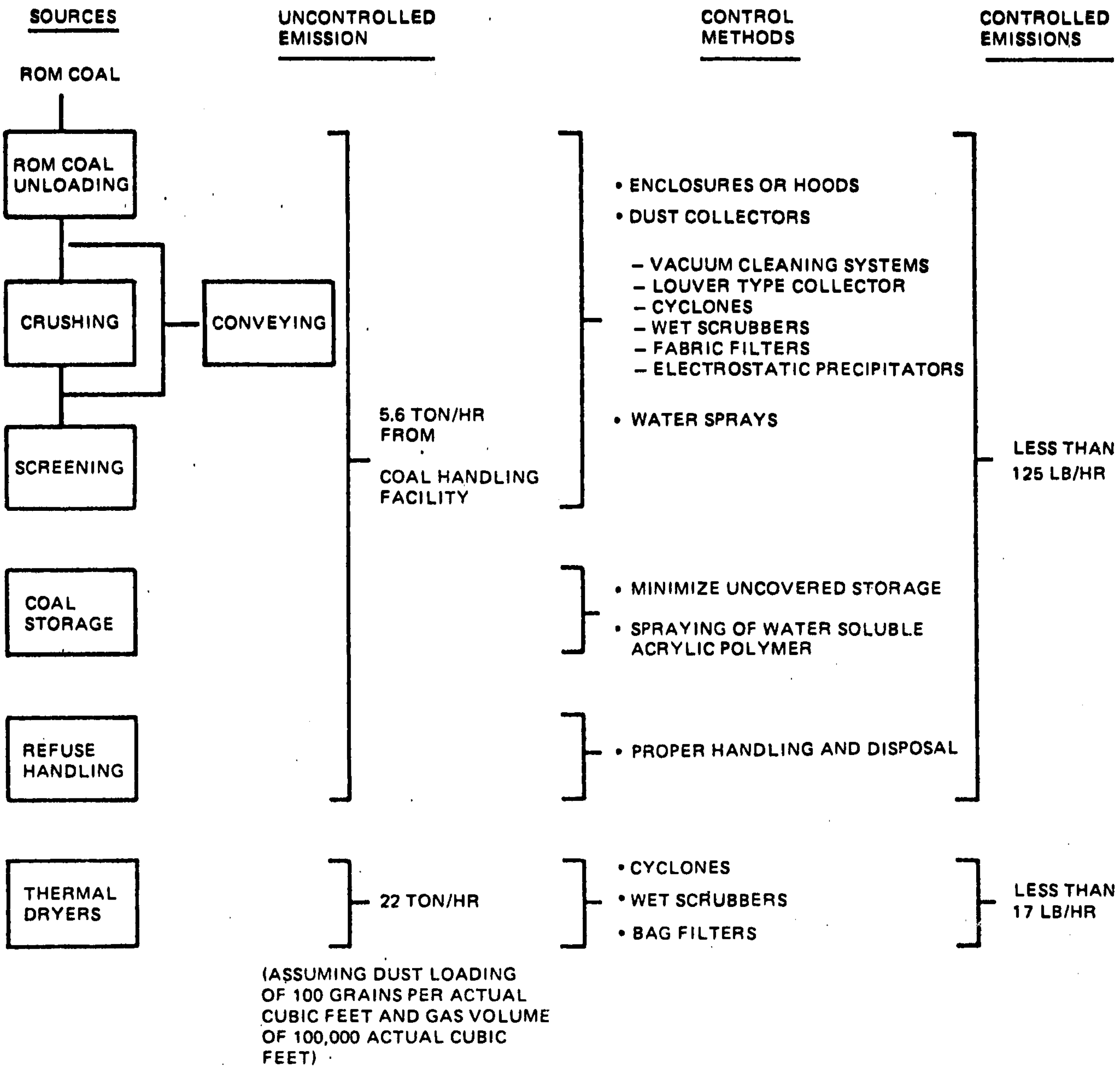


TABLE 13

AIR EMISSIONS FROM COAL, STONE AND SOLID WASTE HANDLING

Source		Amount lb/hr
Coal Handling and Storage		125
Coal Drying		17
Stone Handling and Storage		25
Solid Waste Handling		136
	Total	303

TABLE 14
COAL-PILE RUN OFF ANALYSIS AT SELECTED PLANTS (REF. 11)
(MG/1)

PLANT	A	B	C	D	E	F	G
Alkalinity	6	0	-	-	14.32	36.41	-
Total Solids	1,330	9,999	-	-	-	-	6,000
TDS	720	7,743	-	28,970	-	-	5,800
TSS	610	22	-	100	-	-	200
Ammonia	0	1.77	-	-	-	-	1.35
Nitrate	0.3	1.9	-	-	-	-	1.8
Phosphorus	-	1.2	-	-	-	-	-
Turbidity	505	-	-	-	2.77	6.13	-
Acidity	-	-	-	21,700	10.25	8.84	-
Total Hardness	130	1,109	-	-	-	-	1,851
Sulfate	525	5,231	6,837	19,000	-	-	861
Chloride	3.6	481	-	-	-	-	-
Aluminum	-	-	-	1,200	-	-	-
Chromium	0	0.37	-	15.7	-	-	0.05
Copper	1.6	-	-	1.8	-	-	-
Iron	0.168	-	0.368	4,700	1.05	0.9	0.06
Magnesium	0	89	-	-	-	-	17.4
Zinc	1.6	2.43	-	12.5	-	-	0.006
Sodium	1,260	160	-	-	-	-	-
pH	2.8	3	2.7	2.1	6.6	6.6	4.4

4.3 EMISSIONS FROM SOLID WASTE DISPOSAL AREAS

4.3.1 Disposal Problems

The solid wastes from the fluid bed system consist of the spent material from the beds and the particulates captured in the particulate collection system. The total quantity of solid waste to be discharged is 68.5 tons/hr at full load. This material is unique in that it is dry as contrasted with scrubber sludge.

The pressurized fluid bed system utilizes dolomite as a sulfur sorbent in part because limestone is not expected to calcine at the selected operating conditions of the bed. The solids from the PFB system are therefore expected to be $CaSO_4$, $CaCO_3$ and MgO at the full load operating conditions. The $CaCO_3$ however may calcine during the depressurizing process of the ash letdown system. In addition, at lower loads the bed performance favors the calcining of the $CaCO_3$ due to the reduced partial pressure of CO_2 resulting from the PFB turndown concept. Hence, consideration for disposal of the PFB solid wastes must, as for the AFB solid waste, consider the potential for the presence of CaO.

Several studies are progressing on the use of fluid bed waste as landfill. The testing has included leachate and shake test by Westinghouse, and Lysimeter and field cell tests by Ralph Stone and Company. Results reviewed to date have indicated trace metal concentrations in the leachate are within acceptable limits for discharge purposes. Problem areas, which may prevent direct discharge, are the high total dissolved solids (calcium) and the high pH (10.5-11.6) of water passing through the spent bed material. A future problem may arise with the concentrations of Ni, Pb, As, Hg, and Cd when large quantities of material are placed in disposal sites. These trace element concentration problems are expected to be generic for coal combustion processes and are not unique to fluid bed systems.

Although no actual set of standards exists for leachate, threshold limits are to be established by the Federal EPA as soon as the data is compiled. With this set of standards for comparison, the trace element concentration in an actual landfill may exceed the allowable limits. It should be noted that the solubility of trace metals is related to pH. The high pH of the fluid bed waste material may cause trace metals to precipitate in the limestone mixture, thereby preventing most of the trace metals from leaving the landfill in the leachate. See section 4.4.3 for further discussion on trace element solubilities. Monitoring of a test landfill area would provide the information needed to evaluate the effectiveness of this mechanism.

4.3.2 FBC Waste Material Utilization

A number of other disposal choices are being tested, some of which may help pay disposal costs. Agricultural possibilities are being explored for use of the spent bed material as a source of sulfur and calcium for crops. The waste material, when mixed with compost, forms a nutrient fertilizer. A program now in progress in Alabama has corn growing in a soil mixture containing the fluid bed waste. Results, however, will not be known until after the full growing season ends in March, 1978. Peanuts have already been successfully grown in an agricultural mix of the fluid bed waste material.

Crop yields for the peanuts were similar to the yields obtained when a commercial calcium source was applied. A series of feeding tests will be conducted in which crops grown with the waste material will be fed to animals to see if trace metals are passed through the feed cycle via the plants. These tests will be carried out in the fourth and fifth year of the agricultural program.

Although the future outlook is positive concerning farming usage of fluid bed material, two main drawbacks remain. The first is the limited scope of the agricultural area. The cost of transporting the material restricts its use to the farms surrounding the power plant. To attempt to ship the material over long distances would raise the cost beyond that of a commercial fertilizer. Secondly, the amount of bed waste which can be disposed of in this manner is small compared to the huge amounts that would be generated with commercial usage of fluid bed power plants. Other uses must be developed which will consume large amounts of the waste.

As more samples of the waste material become available, they are to be shipped to Pennsylvania, New York, and Ohio. It is hoped that sewage waste can be stabilized and neutralized by the fluid bed material additive. The Philadelphia Sewage Authority of Pennsylvania plans to investigate the treatment of sewage wastes with the bed material. Compressibility and load factor results have been encouraging enough to interest officials of New York and Ohio. New York City is interested in building blocks of the bed material. The blocks are to be applied in the building of jetties in the harbor area. In Ohio, the bed material is to be tested in place of the usual limestone in road bases.

Babcock & Wilcox Company has mixed different amounts of fluid bed ash with wet scrubber sludge to form a cementitous type of landfill material. Results show a 20-25% fluid bed concentration can withstand pressures necessary to support light construction equipment. Additional tests have revealed no problems with leachate contamination other than high pH levels.

There has been speculation that fluid bed material could be used as a limestone replacement in a scrubber due to the high percentage of unreacted CaO. Naturally, this idea has fair economic potential, but then the problem of sludge disposal does not disappear. It is merely displaced from one plant to another.

Another question to consider is whether active CaO and MgO in the solid waste will cause any problems in a landfill without first being treated or slaked. When water is added to active CaO or MgO, an exothermic reaction occurs. This hydration reaction can be very violent if too little water is added to the oxide particles. For example, in such a case the CaO could become "burned". The temperature of the reaction can reach as high as 400 to 500F, causing dehydration of nearby hydrated particles which can make them very unstable.

Should large amount of water be added too quickly to the oxide particles, another unstable condition will occur. The oxide particles become "drowned". In this instance, the outer layer of the particle reacts to form an impervious layer which prevents water from reaching the interior. Thus, the reaction does not go to completion, forming an unstable product that is liable to hydrate later, causing unwanted heat generation. It may be that neither of these reactions will cause environmentally adverse effects. However, further efforts are required to confirm this.

Until it is certain no adverse environmental effects occur, some precautions are required in order to use the landfill method for disposal of solid wastes from fluid bed systems. The high pH of the leachate samples implies the need to use an impervious liner before the dry effluent is placed in a landfill. It may also be necessary to cover the material because the dry bed material has a tendency to air slake, forming very fine dust particles which could add to the fugitive emissions of the area. An alternative to the landfill method is ponding. Again, an impervious liner is required to prevent the formation of an unacceptable leachate of high pH. In both disposal methods, the containment of the leachate is important as the pH levels of 10-13 are above the new limits developed by Tennessee (6-9 pH). The treatment of the leachate water should not cause any special problems.

4.3.3 Methods for Reducing the Quantity of Solid Wastes

For a given FBC rating and fuel, the coal firing rate is fixed. Thus, in order to reduce the quantity of solid wastes generated in the FBC process, it is necessary to reduce the sorbent feed rate while maintaining the desired level of sulfur retention in the bed. Reducing sorbent feed requirements results in a less severe solid waste problem and improved FBC efficiency due to reduced sorbent calcination requirements and sensible heat losses in the extracted bed material.

The following approaches are being pursued to reduce sorbent make-up rates.

4.3.3.1 Enhancement of Sorbent Effectiveness

Enhancement of sorbent reactivity or increased sorbent utilization may be affected by controlled calcination, catalyst addition, sizing operations and modification of the FBC operating conditions (i.e., gas residence time, etc.).

4.3.3.2 Alternate Sorbents

Although calcium-based sorbent materials (limestones and dolomites) have been employed thus far in FBC test units, materials other than these may exhibit properties suitable for FBC sorbent application.

Potential alternate sorbent materials are being investigated for suitability under FBC operating conditions. A listing of these potential materials is presented in Table 15.

TABLE 15

EVALUATION OF ALTERNATE SORBENTS (Ref. 13)

Reaction Conditions: 870-900° C, 0.1% SO_2, 5% O_2

Simple Oxides: Sulfated	Simple Oxides: Did Not Sulfate		Aluminates (Sulfated)
Na_2O (1)	MgO	Ce_2O_3	$Li_2Al_2O_4$
BaO	Mn_3O_4	ThO_2	$BaAl_2O_4$
SrO	CoO	Bi_2O_3	$SrAl_2O_4$
CaO	NiO	Y_2O_3	$Ca_3Al_2O_6$
La_2O_3	ZnO		

Other Materials: Sulfated	Other Materials: Did Not Sulfate	Titanates: Sulfated	Titanates: Did Not Sulfate
$BaCO_3$	$(CaO)_3SiO_2$	Li_2TiO_3	$PbTiO_3$
$CaCO_3$		$BaTiO_3$	
$CaSiO_3$		$SrTiO_3$	
$BaSiO_3$		$CaTiO_3$	
$BaZrO_3$			

CaO - Containing Composites (All Sulfated) (2)

$(CaO)_3Al_2O_3$

$(CaO)_3Al_2O_3$-7% Na_2O

$(CaO)_3Al_2O_3$-1.1% Na_2O

$(CaO)_3(SiO_2Al_2O_3)_{1/2}$-14.6% Na_2O

$(CaO)_3(SiO_2Al_2O_3)_{1/2}$-0.5% Na_2O

$(CaO)_3SiO_2$-3% Na_2O

85% CaO/10% SiO_2/5% Na_2O

CaO + Portland Type 1 Cement

CaO + Calcium Aluminate Cement

Notes: (1) melted on sulfation
(2) formulas indicate stoichiometric proportions of starting materials only, not composition of final sorbent after heat treatment

Substances found to exhibit higher sorbent reactivity than limestone or lime are composites of $CaO-SiO_2-Al_2O_3$ and especially Cao-calcium aluminate cement (CAC). Calcium titanate, barium titanate, and barium carbonate also display higher reactivity than most other sorbents.

4.3.3.3 Sorbent Regeneration

The objective of sorbent regeneration is to reconvert the spent sorbent ($CaSO_4$) to CaO or $CaCO_3$ for reinjection into the FBC units with minimum sorbent deactivation, and to economically recover elemental sulfur or sulfuric acid. The effectiveness of a sorbent regeneration process is a function of the concentration of sulfur in the regenerator off-gas, the regenerator $CaSO_4$-to-CaO conversion efficiency, and the effective number of regeneration cycles possible without significant degradation of sorbent reactivity.

A rather complete technical and economic study of sorbent regeneration systems is reported under Subtask 1.3 of this program. This study concludes that, while sorbent regeneration is feasible, a considerable amount of developmental effort is required to successfully integrate the regeneration process into a combined cycle FBC power plant and to demonstrate the technical and economic viability of sorbent regeneration for commercial applications (Ref. 14).

4.4 EFFLUENTS FROM WATER TREATMENT SYSTEMS

4.4.1 Water Supply and Treatment

The expected overall yearly average water demand for the conceptual plant is estimated to be 5100 gpm, of which approximately 5000 gpm of water will be drawn from the North River for make-up to compensate for blowdown and evaporation losses. Approximately 100 gpm of water will be drawn from the city water system for make-up feedwater for the AFB boiler. An analysis of the North River water is shown in Table 16. An analysis of the city water is shown in Table 17. The make-up feedwater analysis is presented in Table 18.

The make-up water treatment plant consists of two (2) activated carbon purifiers and two trains of ion exchange regenerators. The average backwash water demand is approximately 110 gpm. The activated carbon purifier backwash demand is based on backwashing each of the two units once per week for 14 minutes at 357 gpm. Ion exchange regeneration demand is based on regenerating each cation and anion unit once per day and each mixed bed unit once per week.

4.4.2 WASTE WATER TREATMENT PLANT DESCRIPTION

The wastewater treatment plant is divided into two parallel units, one of which is used while the other is on standby service available to be used for any unusual spillage or excessive waste flow condition of operation.

TABLE 16
COOLING TOWER MAKEUP WATER
ANALYSIS

Constituted	Ppm As	North River Makeup	Cooling Tower 3X Conc.	Cooling Tower Acid Feed & 3X Conc.
Cations				
Ca^{++}	$CaCO_3$	192	576	576
Mg^{++}	$CaCO_3$	82	247	247
Na^+ & K^+	"	512	1537	1537
Total Cations	"	786	2360	2360
Anions				
HCO3	"	136	408	5
Cl	"	527	1581	1581
Fl	"	1	3	3
NO_3	"	3	9	9
SO4	"	118	359	762
Total Anions		786	2360	2360
pH	"	8.0	8.3	6.5
Free CO_2	CO2	3.0	3.0	3.0
Silica	SiO2	17.3	52.0	52.0

TABLE 17
CITY WATER COMPOSITION

Item	Concentration
pH	9.0
Total Hardness as $CaCO_3$	100 mg/l
Calcium Hardness as $CaCO_3$	78
Magnesium Hardness as $CaCO_3$	22.0 mg/l
Total Alkalinity as $CaCO_3$	35.6 mg/l
Carbonate Alkalinity as $CaCO_3$	23.0 mg/l
Bicarbonate Alkalinity as $CaCO_3$	12.0 mg/l
Chloride as Cl	28.0 mg/l
Color	0.5
Iron as Fe	0.4 mg/l
Ammonia Nitrogen (N)	- mg/l
Nitrate Nitrogen (N)	0.90 mg/l
Dissolved Oxygen	6.0 mg/l
Phosphate (PO_4)	-
Phosphate (Metapoly)	1.0 mg/l
Sulfate as SO_4	40.0 mg/l
Total Solids	110.0 mg/l
Total Dissolved Solids	110.0 mg/l
Turbidity JTU	1.2
Silica (SiO_2) *	5.0 mg/l

* Colloidal silica is also present on an intermittent basis. It has averaged .15 ppm, but has been as high as .67 ppm.

TABLE 18
MAKEUP BOILER FEED WATER ANALYSIS
(CITY WATER AFTER TREATMENT)

Item	Concentration
Hardness	*
Organics	*
Chloride	100 ppb
Total Silica (as SiO_2)	5 ppb max.
Total Iron (as Fe)	10 ppb max.
Total Copper (as Cu)	2 ppb max.
Total Dissolved Solids	500 ppb max.
Total Suspended Solids	50 ppb max.
Conductivity	1.0 micromho

* Below detectable limits

The water treatment plant will be sized for 150 gpm minimum, 300 gpm maximum.

The blended waste streams flow into one of two collection ponds. Oil rises to the top of the pond level by means of gravity separation and is removed periodically by manual skimming. A skimming device is located between the two ponds so that oil can be removed from either pond and placed in a storage tank for off-site disposal.

Water flows from the collection pond to a pH trim tank, where the pH is measured and automatically adjusted to predetermined limits by the addition of acid or caustic solutions.

The pH adjusted water then flows through a clarifier where coagulant and coagulent-aid chemicals are injected automatically and in proportion to the flow being treated. The clarifier is designed to accomplish complete coagulation and remove suspended solids from the stream being treated.

The treated water is continuously monitored for pH, suspended solids, oil, chlorine and temperature. In the event of an excursion from acceptable Federal Register limits, the water of unacceptable quality is automatically recycled back to the collection ponds by gravity flow.

Sludge blowoff from the clarifier unit is routed to one of two separate sludge drying ponds. These ponds are located adjacent to the collection ponds.

4.4.3 Quantity, Characteristics and Treatment of Wastewater Streams

Table 19 identifies the waste streams along with flow quantities and frequency. The totals obtained from this tabulation have been used to determine the criteria for the design capacity of the Central Wastewater Treatment Plant.

Ion exchange regeneration solutions from the condensate demineralizers and makeup water treatment ion exchangers are neutralized before discharge. The combined regenerants and rinses may contain 20 to 200 ppm of total iron, manganese, copper, and zinc and 100 to 400 ppm of calcium, magnesium, aluminum, potassium, and chloride in a 5000 ppm solution of sodium sulfate.

Steam/condensate cycle blowdown contaminants principally consist of soluble sodium salts, and small concentrations of iron, copper, nickel, and chromium. Ammonia and hydrazine may be present at concentrations up to 1.5 and 0.5 ppm, respectively, with a total solids concentration of approximately 5 ppm.

Chemical cleaning wastes are generated in amounts approximately equal to the corresponding filtered water usage. Large quantities of waste are generated in a year during which boiler cleaning operations are accomplished, and much of this waste is generated within a period of 2 to 4 days. This peak generation of waste is expected to occur at 4-year intervals. Boiler cleaning wastes are high in iron, copper, nickel, zinc, chromium, calcium and magnesium.

Floor drain and miscellaneous wastes such as equipment drains are generated on a continuous (leakage) and a periodic (floor wash and maintenance

TABLE 19
ESTIMATED WASTER TREATMENT STREAMS TO
CENTRAL WASTE TREATMENT SYSTEM

Source	Instant. Flow	Batch Vol.	Frequency	Max. Vol. 24 hrs.	Aver. Volume Annually
Boiler Blowdown	30 gpm	400,000 gal.	continuous	43,200 gal.	15,800,000 gal.
Boiler Chemical Cleaning		400,000 gal. (2-day cleaning cycle)	once/4 yrs.	200,000 gal.	200,000 gal. at 4 yr. intervals
Boiler Fireside Wash	200 gpm	36,000 gal. (one day cleaning cycle)	annually	36,000 gal.	36,000 gal.
Condenser & Heater Acid Cleaning		300,000 gal. (one day cleaning cycle)	once/4 hrs.	300,000 gal.	300,000 gal. at 4 yr. intervals
Boiler Area Drains	75 gpm		periodic		225,000 gal.
Turbine Floor Drains	75 gpm		periodic		225,000 gal.
Turbine Floor Drains (Oil)	50 gpm		periodic		150,000 gal.
Service Bldg. Floor Drains	50 gpm		periodic		150,000 gal.
Coal Pile Reclaim Pits	400 gpm		@ rainfall	51,900 gal.	465,000 gal.
Sorbent Pile Reclaim Pits	400 gpm		@ rainfall	51,900 gal.	465,000 gal.
Coal Pile Runoff	800 gpm		@ rainfall	550,000 gal.	5,000,000 gal.
Yard Area Runoff	4,000 gpm		@ rainfall	190,000 gal.	1,700,000 gal.

TABLE 19

Source	Instant. Flow	Batch Vol.	Frequency	Max. Vol. 24 hrs.	Aver. Volume Annually
Transformer Pits	350 gpm		@ rainfall	16,500 gal.	150,000 gal.
Oil Storage Area	300 gpm		@ rainfall	15,000 gal.	130,000 gal.
Chemical Treatment Area	235 gpm	38,000	daily	38,000 gal.	14,000,000 gal.
Chemical Lab Drains	10 gpm		continuous	1,000 gal.	365,000 gal.
Roof Drains	3,000 gpm		@ rainfall	136,000 gal.	1,200,000 gal.
Clarifier Blowoff	480 gpm	960 gal. (2 min.)	75 times a day	72,000 gal.	26,300,000 gal.
					66,651,000 gal.

Waste Treatment System to be desing
for oper. range from 150 gpm to 300 gpm
maximum flow.

$$\frac{66.7 \times 10^6}{1440 \times 365} = 127 \text{ gpm total average flow}$$

drainage) basis, with housekeeping practices having a significant effect on the quantity of waste produced. Therefore, total dissolved and suspended solids vary, depending upon operations in progress. Local oil separators are provided at sources with potential for oil contamination, but some oil is expected in the combined waste stream. At times, detergent and high suspended solids are also present.

Coal pile runoff results from the percolation of rain and snow water through the pile. To prevent contamination of ground water, coal is stored on an impervious layer of clay, and runoff is routed via the coal pile settling basin to the equalization basins of the central waste treatment system. Figures 8 and 9 show the average annual and ten-year 24-hour precipitation of the United States from which the design basis was established. Based on this data, design runoff quantities are calculated for 36 inches of precipitation per year and 5 inches in 24 hours. Both dissolved and suspended solids (coal fines) enter the runoff. The pyritic content of the coal is particularly important in determining acidity of the runoff since the reaction of iron sulfides with oxygen produces the sulfate and acid. The acid dissolves many other complexes and releases metals and other pollutants. Data on pollutant concentrations are given in Tables 14 and 20. Wide ranges are attributable to both coal properties and precipitation and drainage conditions. Drainage rates vary greatly from day to day, depending upon precipitation, and the lowest pH and highest pollutant concentrations are associated with lower drainage rates. The composite analysis of elements in coal, given in Tables 6 and 21, indicates the full spectrum of potential elemental contaminants, but their release to runoff depends strongly on their chemical form in the coal. It should be noted that the analyses shown on Tables 6 and 21 are illustrative only and do not necessarily apply specifically to the design coal and sorbents.

Sorbent pile runoff results from precipitation percolating through the storage pile. Runoff is again based on 36 inches of precipitation per year and 5 inches in 24 hours as discussed above. Runoff is routed to the equalization basins where its alkalinity helps offset the acidity of the coal pile runoff. The spectrum of potential elemental contaminants is indicated by the composite analysis of dolomite and limestone given in Table 21. The actual concentrations in the runoff depend on the solubilities of the chemical species present in the limestone and dolomite. The major constituent of the sorbent pile runoff is expected to be calcium and magnesium hydroxide with pH in the 8 to 9 range.

Ash and spent sorbent are delivered to an emergency ash and spent sorbent area only when off-site shipment is temporarily restricted. Runoff caused by rain or snow is routed to the equalization basins where its alkalinity tends to offset the acidity of the coal pile runoff. Runoff quantities also are based on 36 inches of precipitation per year and 5 inches in 24 hours. Based on small scale investigations carried out to date, the runoff has a pH between 10 and 13, with potential for

TABLE 20

COMPOSITION OF DRAINAGE FROM COAL PILES
(Ref. 11)

	Concentration, mg/l
Alkalinity	15 - 80
BOD	3 - 10
COD	100 - 1,000
Total Solids	1,500 -45,000
Total Suspended Solids	20 - 3,300
Total Dissolved Solids	700 -44,000
Ammonia	0.4 - 1.8
Nitrate	0.3 - 2.3
Phosphorus	0.2 - 1.2
Turbidity	6 - 505
Acidity	10 -27,800
Total hardness	130 -20,000
Sulfate	20 - 480
Chloride	825 - 1,200
Aluminum	0 - 16
Chromium	1.6 - 3.9
Copper	0.4 - 2.0
Iron	90 - 180
Magnesium	160 - 1,260
Sodium	2.2 - 8.0

TABLE 21

TYPICAL VALUES OF TRACE ELEMENTS IN DOLOMITE, LIMESTONE AND COAL (ppm)
(Ref. 11)

Element	Argonne Dolomite	Tymochtee Dolomite	Limestone	Lignite	Average or Typical Bituminous
As	1.9	0.566 ± 0.17	<6	8	30
Bs	5		30-300	280	100
Be	2		<2	1.5	2.5
Br	2	6.75 ± 1.4	<0.3		15
Cd	14		<0.3	0.2	0.4
Ce	0.9		<3		
Cu		1.03 ± 0.21	<2	3	4
Cr		4.23 ± 0.85	<20	7	14
Cs		0.439 ± 0.091	<0.06		
Dy					
Eu		0.0598 ± 0.013	<1		
Fe	5.6×10^3	3240 ± 650	200-2000	6344	1.86×10^4
Hf					
Hg					0.2
K	4.6×10^3	2180 + 440	100-1000	0.1	-
La	3.4		0.3-3	551	1927
Mn	55	42 ± 8.4	6-60	38	50
Na	368	303 ± 61	10-100	1×10^4	481
Ni			<6	7	14
Rb		12.2 ± 2.5	<2		
Pb			<3	7	9
Sb		0.0527 ± 0.015	<0.3	0.4	0.5
Sc	1.5	0.952 ± 0.19	<0.3		
Se			<3	1.3	3
Sm		0.658 ± 0.13	<1		
Sr		130 ± 29	100-1000		
Ta					
Te			<0.3	0.11	0.3
Tb		2.81 ± 0.63	<0.2		
Th		0.58 ± 0.12		<0.1	0.1
Yb					
Zn			<30	12	8
U		2.23 ± 0.45	<0.6	150	15
V			0.06-0.6	16	30

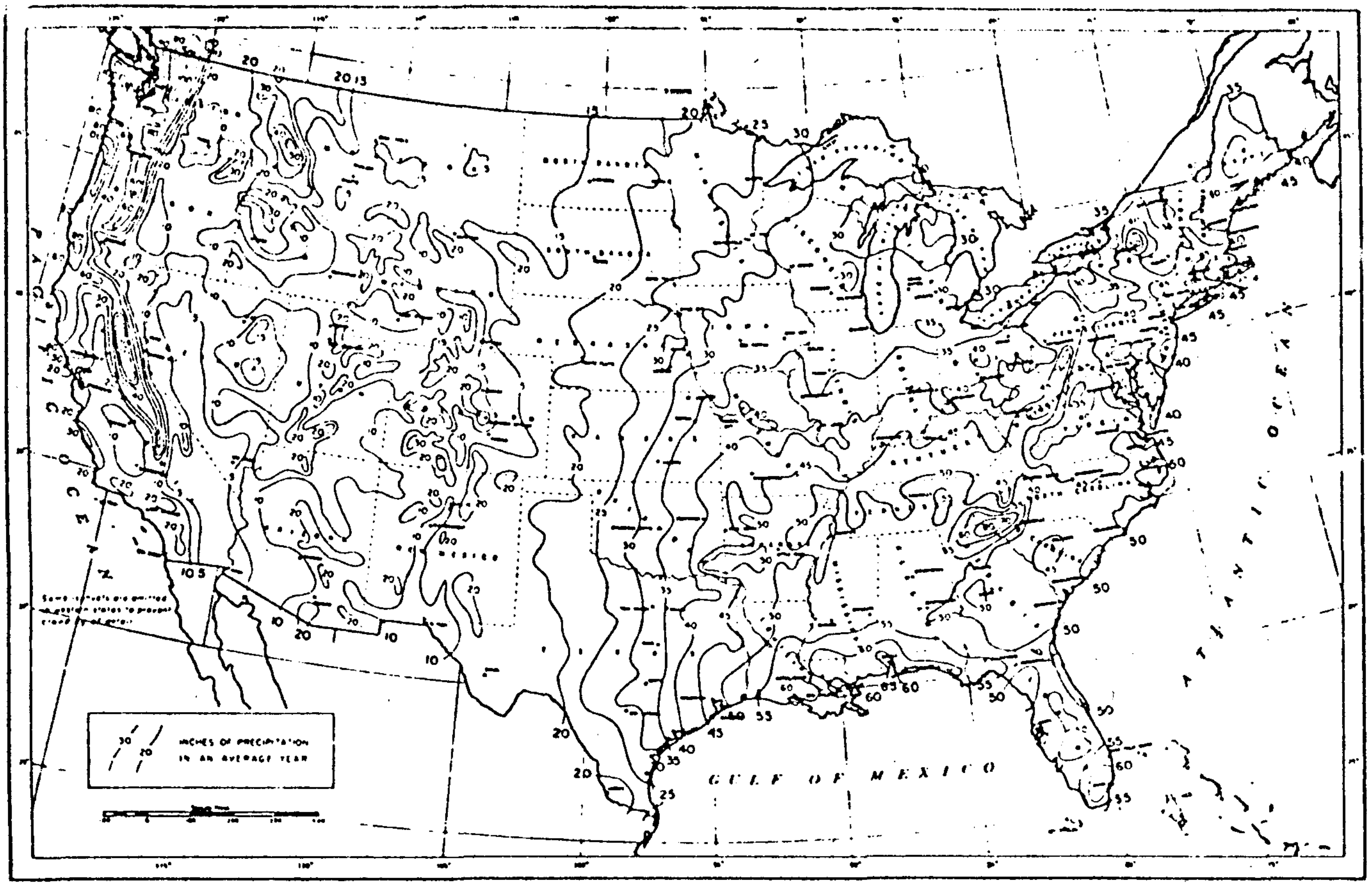


Figure 8 AVERAGE ANNUAL PRECIPITATION OF THE UNITED STATES
(Ref. 15)

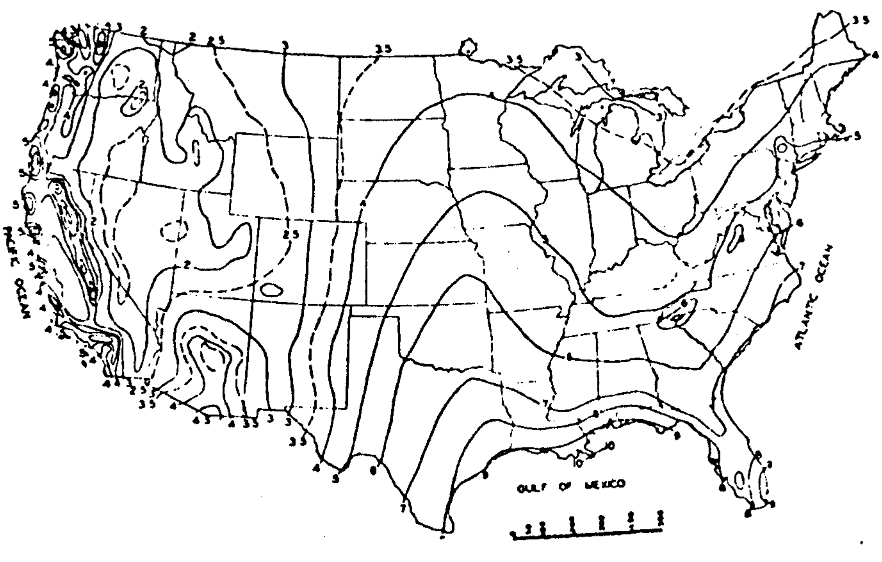


Figure 9 TEN-YEAR 24-HOUR RAINFALL (INCHES)
(Ref. 15)

high extraction of calcium and sulfate. Furthermore, the runoff from the waste material is diluted by runoff from the empty portion of the ash and spent sorbent storage area. Thus, ash/spent sorbent pile runoff is expected to be significantly lower in pH and sulfate than test data suggests.

As indicated in section 4.3, trace element leaching is a major pollution concern because of the large inventory in the ash. It is probably advantageous that the ash and spent bed material are mixed since, in general, trace elements display greatly decreasing solubilities with increasing pH as provided by the bed material. Figure 10 indicates this decreasing solubility with increasing pH. The trace elements of major concern appear to be arsenic, cadmium, chromium, lead and mercury as indicated in Table 22. Vanadium has also been indicated to be a major concern as has the potential for increased solubilities due to complex formation. Relative solubilities are presented in Table 23.

Oil storage area runoff can be contaminated by oil spills or leakage from pump or valve seals, etc., as well as by suspended solids. Local oil separators are provided to remove most of the oil before discharge to the central waste treatment system.

Paved area runoff, which is potentially high in oil and suspended solids, is routed to the central waste treatment system. Oil interceptors are provided to reduce oil contaminated runoff from major vehicle parking and maintenance areas.

Dewatering system filtrate can be relatively high in suspended solids and pH, both governed by the pretreatment clarifier underdrains. This alkalinity, together with that from sorbent pile and emergency ash and spent sorbent pile runoff (if any) normally exceeds the acidity of the coal pile runoff so that the wastewater in the equalization basins is normally basic. This is a desirable condition, since trace metals tend to precipitate as hydroxides.

There are two waste sources that are not treated in the central waste treatment system. These are runoff from other improved areas and sewage. Improved area runoff during both construction and after plant operation can be high in total suspended solids. Rates are based on 36 inches of precipitation per year and 5 inches rain in 24 hours, as indicated above. The sewage rate is based on toilet facilities for 100 persons and 100 gallons per day per person. Sewage will be treated in an extended aeration treatment plant rated for 10,000 gallons per day. The unit consists of aeration, clarification and sludge compartments, complete with air blowers and distributors and chlorinator. Clarifier effluent is periodically monitored for biological oxygen demand and residual chlorine. Sludge removal from the treatment plant is periodically transported to a disposal site.

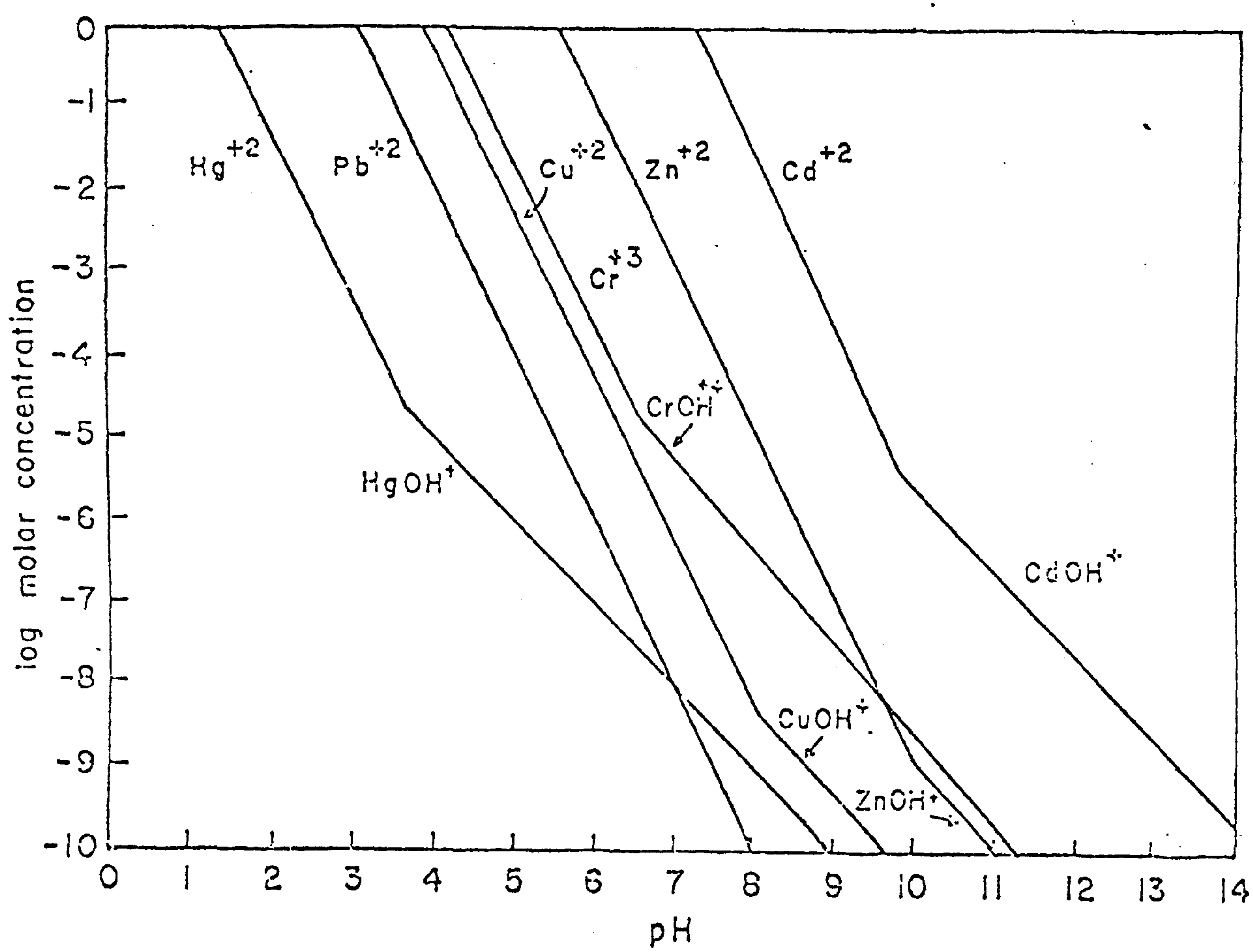


Figure 10 SOLUBILITIES OF TRACE METALS - FREE AQUEOUS AND MONO-HYDROXO COMPLEXES ONLY CONSIDERED (Ref. 11)

TABLE 22

TRACE METALS IN ASH, FGD SLUDGE POND, LIQUORS/SUPERNATANTS, AND FBC ASH AND SPENT STONE (EXCEPT pH, CONCENTRATIONS IN PPM) (REF. 12)

	Ash Pond		FGD Sludge Pond		FGD Sludge Liquors		FBC Ash Leachates		FBC Spent Stone Leachates		EPA Proposed Standards Public Water Supply Intake
	Mean	High	Mean	High	Mean	High	Mean	High	Mean	High	
pH	10.9[a]	12.5	8.9	9.7	7.4	9	11.4	12.2	12.16	12.5	5 to 9
Antimony	0.017	0.33	0.021	0.035	-	-	-	-	0.3	0.3	-
Arsenic	0.036	0.084	0.011	0.03	0.068	0.20	0.68	2.5	5.0	5.0	0.1
Barium	8.24	40.	0.866	2.0	-	-	-	-	-	-	-
Beryllium	0.0011	0.003	0.002	0.002	0.041	0.18	-	-	0.01	0.01	-
Boron	3.66	16.9	3.286	6.3	-	-	0.39	0.61	0.37	0.83	1.0
Cadmium	0.0031	0.01	0.0012	0.002	0.038	0.10	0.0025	0.01	0.1	0.2	0.01
Chromium	0.267	1.0	0.0043	0.011	0.087	0.21	ND	0.1	0.1	0.1	0.05
Copper	0.031	0.092	0.027	0.045	0.070	0.20	-	-	0.1	0.1	1.0
Fluorine	4.88	17.3	15.93	31.5	-	-	-	-	-	-	-
Germanium	0.01	0.01	0.013	0.02	-	-	-	-	-	-	-
Mercury	0.0033	0.015	0.008	0.001	0.045	0.12	3.8	6.2	8.8	13.2	0.002
Lead	0.0088	0.024	0.005	0.0061	0.072	0.18	1.3	2.5	0.92	2.4	0.05
Manganese	0.002	0.002	0.002	0.002	-	-	-	-	0.05	0.05	-
Molybdenum	0.169	0.69	0.066	0.075	-	-	9.7	17	5.8	12	-
Nickel	0.037	0.05	0.05	0.05	-	-	-	-	0.1	0.1	-
Selenium	0.10	0.47	0.023	0.045	0.75	2.5	-	-	-	-	0.01
Vanadium	0.12	0.2	0.1000	0.1	-	-	-	-	0.05	0.05	-
Zinc	0.055	0.19	0.0270	0.052	0.14	0.30	0.028	0.08	0.4	0.4	5.0
Samples	5		5		5		4		4		

(a) Underline indicates value higher than EPA and WHO standards.

Sources: Data compiled by Battelle from information supplied by Aerospace Corporation, R. Stone and Company, Westinghouse Research Laboratory, and EPA.

TABLE 23

RELATIVE SOLUBILITIES IN WEAKLY ALKALINE SOLUTIONS (Ref. 12)

	Toxic anions		Major anions			
Cations	AsO_3^-	$SeO_3^=$	$CO_3^=$	OH^-	$SO_3^=$	$SO_4^=$
Major						
Ca^{2+}	Slightly Soluble	Insoluble				
Mg^{2+}		Insoluble				
Toxic						
Be^{2+}			Soluble	Insoluble		Insoluble
Cd^{2+}			Insoluble	Insoluble	Slightly Soluble	Soluble
Cr^{2+}				Insoluble		Insoluble
Cu^+			Insoluble	Insoluble	Slightly Soluble	Soluble
Hg^+			Insoluble	Insoluble	Insoluble	Slightly Soluble
Pb^{2+}			Insoluble	Slightly Soluble	Insoluble	Insoluble
Zn^{2+}			Very slightly Soluble	Insoluble	Slightly Soluble	Soluble

4.4.4 Options Available for Compliance with Future Effluent Limits

The wastewater treatment systems described previously are designed to meet present EPA effluent limits for 1983. However, there is a potential for more restrictive discharge limits being enacted before 1985. The first goal of Public Law 92-500 is to eliminate the discharge of pollutants into navigable waters by 1985, conceivably leading to the requirement of essentially zero release plants by that same year. Although zero release plants are possible with present technology, they may not be environmentally and economically sound due to the increases in solid waste, energy usage, and cost.

In the conceptual design, raw water, high in suspended solids and relatively low in dissolved solids, is used for makeup. Processed wastewater, high in dissolved solids and low in suspended solids, is discharged. In certain cases, such discharge can have a beneficial effect on the receiving water body, such as when the waste stream can partially neutralize the receiving stream. If an essentially zero release plant design is deemed necessary or desirable, however, the problem is then one of removing dissolved solids from the waste streams so that the water can be reused in plant water systems. Rainwater runoff could be processed and used as makeup, thereby reducing the average raw water demand to less than that needed to make up for evaporative losses which are mainly due to cooling tower evaporation.

Cooling tower blowdown could be treated to allow reuse as cooling tower makeup and as feed to the makeup water treatment plant.

Given the same liquid effluent limitations, wastewater treatment systems for a conventional pulverized coal-fired plant and a combined cycle FBC plant are expected to be very similar in function, capacity and cost.

4.5 THERMAL DISCHARGES FROM THE CCFBC POWER PLANT

The thermal efficiency of the CCFBC power plant is expected to be about 10% higher than a conventional pulverized coal-fired power plant with flue gas desulfurization. Therefore, the thermal effluent characteristics of the CCFBC plant should be less severe. The quality of thermal discharges from the various sources is indicated below:

Cooling Towers	2352 X 10^6 Btu/hr
Stack and Miscellaneous	700 X 10^6 Btu/hr
Total	3052 X 10^6 Btu/hr

These discharges are based on plant operation at 100% load. No unusual environmental problems are anticipated due to the plant cooling systems.

5.0 PLANT INFLOWS, EFFLUENTS, AND LAND REQUIREMENTS

This section defines the inflows, effluents, and land requirements of the commercial CCRBC plant.

5.1 PLANT INFLOWS AND EFFLUENTS

Table 24 summarizes the inflow of raw materials to the plant. Plant effluents are summarized on Table 1 and compared with applicable current environmental limits.

5.2 LAND USAGE

5.2.1 Area Requirements

In determining the land requirements for this installation some of the items considered were:

- Land availability and cost
- Plant costs vs. land cost
- Minimal environmental impact on immediately surrounding areas
- Creation of suitable buffer zone between plant and local residences
- Adequate space around equipment to insure proper performance and maintainability
- Future expansion of facility
- Adequate space for the installation and operation of the 100 car coal unit trains.

The land usage is shown on the area site plan, Figure 11, and is summarized in Table 25. The areas designated include space required for access, maintainance, and, where applicable, operation of the equipment and facilities.

5.2.2 Options for Reducing Land Requirements

Significant reductions in land usage can only be accomplished in facilities which utilize large land areas. Only five improved areas indicated in Table 25 are more than four acres. The potential options for reducing these five areas, as well as the much larger non-dedicated area, are discussed later. Although additional areas may have potential for some reduction, the impact on the overall land area of the site would be negligible.

One large designated area is the plant island in which essentially all of the buildings, equipment, and apparatus associated with power generation are located. Because of the size and configuration of the equipment, very little reduction can be made in this area.

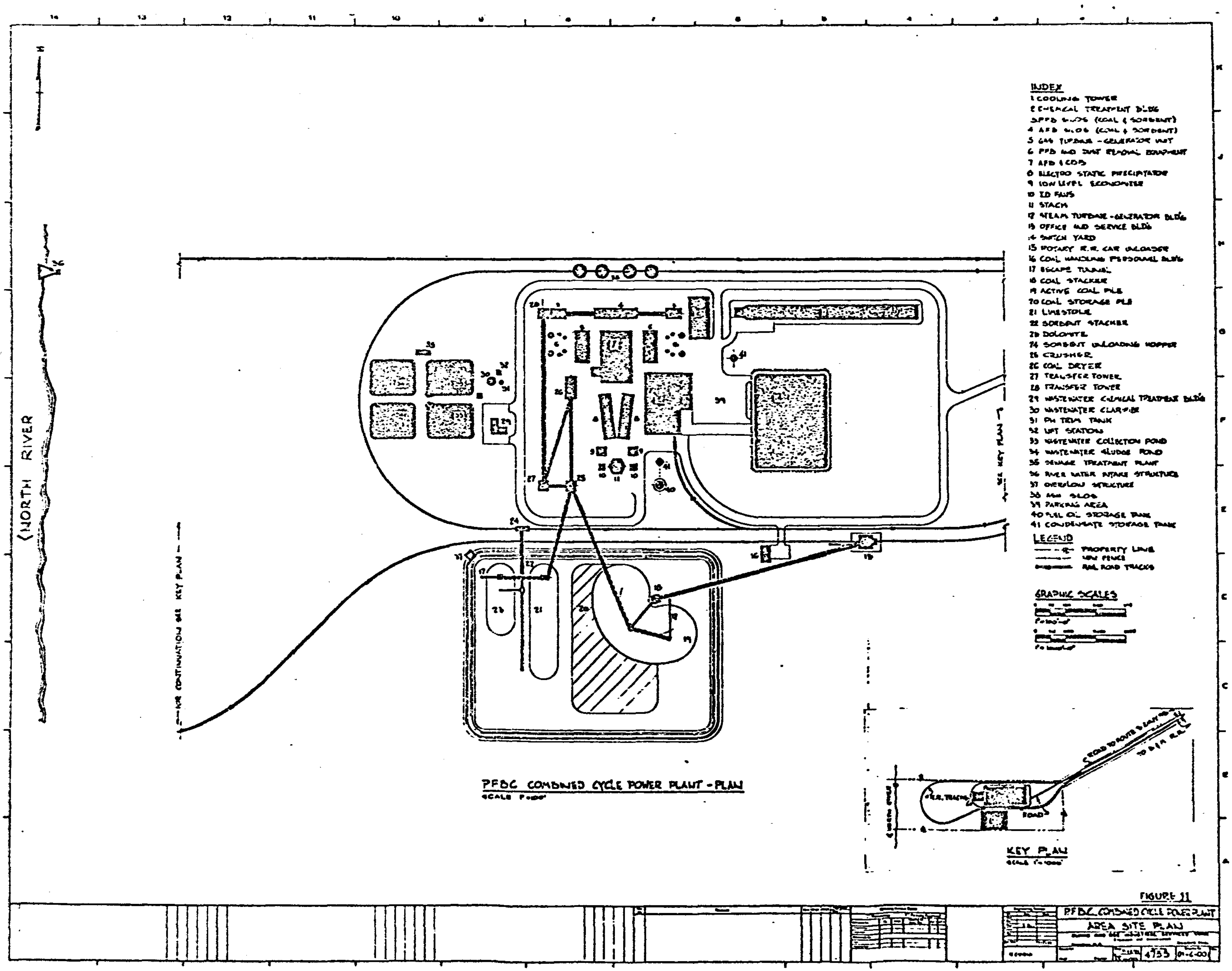

NORTH RIVER
FOR CONTINUATION SEE KEY PLAN
SEE KEY PLAN
PFBC COMBINED CYCLE POWER PLANT - PLAN
INDEX
1 COOLING TOWER
2 CHEMICAL TREATMENT BLDG
3 PFB SILOS (COAL & SORBENT)
4 AFB SILOS (COAL & SORBENT)
5 GAS TURBINE - GENERATOR UNIT
6 PFB AND DUST REMOVAL EQUIPMENT
7 AFB & CDS
8 ELECTRO STATIC PRECIPITATOR
9 LOW LEVEL ECONOMIZER
10 ID FANS
11 STACK
12 STEAM TURBINE - GENERATOR BLDG
13 OFFICE AND SERVICE BLDG
14 SWITCH YARD
15 ROTARY R.R. CAR UNLOADER
16 COAL HANDLING PERSONNEL BLDG
17 ESCAPE TUNNEL
18 COAL STACKER
19 ACTIVE COAL PILE
20 COAL STORAGE PILE
21 LIMESTONE
22 SORBENT STACKER
23 DOLOMITE
24 SORBENT UNLOADING HOPPER
25 CRUSHER
26 COAL DRYER
27 TRANSFER TOWER
28 TRANSFER TOWER
29 WASTEWATER CHEMICAL TREATMENT BLDG
30 WASTEWATER CLARIFIER
31 PH TRIM TANK
32 LIFT STATION
33 WASTEWATER COLLECTION POND
34 WASTEWATER SLUDGE POND
35 SEWAGE TREATMENT PLANT
36 RIVER WATER INTAKE STRUCTURE
37 OVERFLOW STRUCTURE
38 ASH SILOS
39 PARKING AREA
40 FUEL OIL STORAGE TANK
41 CONDENSATE STORAGE TANK
LEGEND
PROPERTY LINE
NEW FENCE
RAIL ROAD TRACKS
GRAPHIC SCALES
KEY PLAN
R.R. TRACKS
ROAD
FIGURE 11
PFBC COMBINED CYCLE POWER PLANT
AREA SITE PLAN
4753

TABLE 24

COMBINED CYCLE FBC POWER PLANT INFLOWS
(574 MW Net Output)

Item	
Water Requirements, gpm	5100*
gpm/MW	8.9
Fuel Type	Average Illinois Bituminous Coal
Fuel Consumption Rate, tons/hour	218.15*
Specific Fuel Consumption, lb/avg hr	0.76
Start-up Fuel (#2 Distillate Oil), gpm	60 (4 hours)
Sorbents	
Dolomite Consumption Rate, tons/hour (Ca/S = 1.0)	13.97*
Specific Dolomite Consumption lb/KWH	0.05
Limestone Consumption Rate, tons/hour (Ca/S = 2.5)	36.91*
Specific Limestone Consumption, lb/KWH	0.13

* At operating conditions and 100% load factor.

HHV coal = 12,453 Btu/lb as fired.

TABLE 25

PRINCIPAL LAND AREA ALLOCATIONS

Area	Length, Feet	Width, Feet	Area, Acres
Plant Island	900	650	13.43
Cooling Tower & Chemical Treatment	850	200	3.90
Switchyard	650	300	4.48
Wastewater Treatment Facility	550	400	5.05
Coal & Sorbent Yard	2,000	800	36.73
Parking Area	200	150	0.69
Access Railroad Right-of-Way	26,400	150	90.91
Balance of Site Acquisition	-	-	184.81
Total Area			340 acres

Another large designated area is the switchyard, which can be reduced by employing a compact substation design. This is done when land has a high value, such as in an urban area, or in areas where industrial or sea air contamination could reduce the reliability of air insulated equipment. (In a compact substation, air insulation is replaced by sulfur hexafluoride.) Land area could be reduced from approximately five acres to approximately two acres.

A third large designated area is the wastewater treatment facility. Very little can be done to reduce this area because of the sizes and configuration of the equipment and land required for its operation.

One of the largest designated areas is the coal and sorbent storage area. The storage area is based on an average depth of 35 feet and 30-day storage relative to the design coal firing rate at 100 percent power. The only significant parameters that are subject to change for a given power level are storage time and depth. The 35-foot average depth is considered near the upper limit and a reduction of storage below 30 days cannot be recommended.

The railroad right-of-way is the largest dedicated area. This area tends to decrease as overall land area decreases, but the percentage usage would remain relatively constant. The ability to lay out a railroad to handle unit trains becomes a problem if the overall designated area for railroad layout decreases.

The non-designated or balance of site aquisition area accounts for more than half of the total site area, and reduction of this area, therefor, provides the greatest potential for reduction of land usage. The overall site area could be reduced by reducing the open areas at the western and eastern corners of the plant. However, the elimination of open areas around the periphery of the plant reduces the buffer zone which would tend to increase problems with aesthetics and noise.

6.0 COMMUNITY NOISE LEVELS

The major sources of off-site noise generated by the CCFBC plant include the induced draft fans, gas and steam turbine-generator units, FBC units, main transformers, coal crushers and conveyors, and mechanical draft cooling towers. These pieces of equipment, with the exception of the FBC units and gas turbine generator units, are the same type generally installed in conventional coal-fired facilities. While this study did not incorporate a detailed acoustical analysis, it is anticipated that traditional acoustical treatments will insure compliance with applicable environmental criteria. For the gas turbine generator units, noise criteria established by the Occupational Safety and Health Act (OSHA) of 1970 specified stringent control of noise levels for personnel protection, and therefore governs the required acoustical treatment of this equipment. Allowance has been made in the plant cost estimates for inlet silencers and exhaust systems sound suppression treatment. A primary sound enclosure is provided around the gas turbine units and a secondary enclosure or building is also provided.

7.0 HEALTH AND SAFETY

This section covers the health and safety aspects of the FBC process and its related systems. Only the areas that differ from those of pulverized coal fired power generating plants are discussed here.

7.1 INVENTORY OF POTENTIAL HEALTH AND SAFETY CONCERNS

In evaluating the health and safety concerns of a CCFBC plant the following areas have been investigated:

- Coal Handling and Preparation System
- Sorbent Handling and Preparation System
- Induced Draft System
- Spent Sorbent System
- Gas Turbine Generator Units

7.1.1 Coal Handling and Preparation System

The CCFBC plant requires the delivered coal received to be dried, crushed, and sized before introduction to the FBC units. The system which performs these functions exposes the plant to potential hazards similar to those in a pulverized coal fired plant.

The potential hazards are dust, fire and explosion. The system design incorporates safety features which minimize if not eliminate these concerns.

The first step is to limit exposure of personnel to a dusty environment. To accomplish this, the system is designed to be controlled from a central control room. This eliminates the requirement of having equipment operators located in the process area while the system is in operation.

To minimize dust conditions in the process area all equipment is supplied with dust tight enclosures. In addition, in order to minimize dust in the process area, a ventilation system is provided. Filtration devices are provided in the ventilation system to remove dust from the air prior to discharge to atmosphere.

A CO_2 and water fire protection system is provided for this and other plant areas.

Coal drying and preparation systems of this type are widely used in the coal industry, and when properly designed pose no serious health and safety concerns.

7.1.2 Sorbent Handling and Preparation System

To overcome the potential dust problems, this system is encased in dust tight enclosures. All operations will be controlled from a central control room so that personnel are not required to be in the dust laden environment. Ventilation similar to that for the coal handling and preparation system is also provided.

7.1.3 Spent Bed Material Systems

The majority of the spent sorbent system is designed with well established and accepted pneumatic system concepts. This system poses no significant health or safety hazards.

One area, however, does require additional protective treatment to insure safe operation. This area includes the piping and equipment between the PFB and AFB units and the spent bed material coolers where spent sorbent temperatures exceed 1500^{o}F. The insulation of these components must be easily maintainable. In order to insure a high degree of safety around these devices, protective caging is provided.

7.1.4 Gas Turbine Generator Units

These units are provided with inlet air silencers and an exhaust system with sound suppression treatment. A primary sound enclosure is provided around the gas turbine units. In addition, a secondary enclosure or building is provided around the complete turbine-generator unit.

7.2 REQUIREMENTS FOR OSHA COMPLIANCE

The CCFBC plant design criteria includes compliance with all OSHA requirements.

As discussed previously in area requirements, a large portion of this installation is of conventional power plant design. The safety requirements for such facilities are clearly defined and commonplace in utility installations.

Plant areas and processes which are not consistent with a basic pulverized coal plant installation are treated individually to provide satisfactory personnel protection based on process design requirements.

7.3 OPTIONS FOR ADDITIONAL PROTECTION

As previously discussed, the design of the CCFBC facility provides a high degree of protection against hazards to personnel, equipment and the site in general. In the event of a mishap, however, systems have been incorporated to minimize the impact on personnel and equipment (i.e. fire protection system, emergency personnel showers and eyewashes, first aid facility).

Although the safety systems, included in the plant design, fall well within accepted utility practices, additional devices and systems might be incorporated to further increase plant protections.

These additional systems could include:

- Explosion suppression system for coal dryer
- Temperature monitoring of coal bunkers for fire detection
- Co_2 fire protection system for ash silo
- Temperature monitoring of ash silo for fire detection.

These systems, although not found in conventional pulverized coal fired utility plants, should be considered for further study to assess their value in a CCFBC installation.

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