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Experimental/Engineering Support for EPA's FBC Program: Final Report - Volume IV. Engineering Studies

by

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PREFACE

The Westinghouse R&D Center is carrying out a program to provide experimental and engineering support for the development of fluidized-bed combustion (FBC) systems under contract to the Industrial Environmental Research Laboratory (IERL), U. S. Environmental Protection Agency (EPA), at Research Triangle Park, NC. The contract scope includes atmospheric (AFBC) and pressurized (PFBC) fluidized-bed combustion processes as they may be applied for steam generation, electric power generation, or process heat. Specific tasks include work on calcium-based sulfur removal systems (e.g., sorption kinetics, regeneration, attrition, modeling), alternative sulfur sorbents, nitrogen oxide (NO_x) emission, particulate emission and control, trace element emission and control, spent sorbent and ash disposal, and systems evaluation (e.g., impact of new source performance standards (NSPS) on FBC system design and cost).

This report contains the results of work defined and completed under technical directives issued by the EPA project officer. Work on these tasks was performed from January 1976 to January 1979 and is documented in the following EPA contract reports:

- The present report, which presents the results of four technical directives on systems evaluation
- Report on an engineering assessment of intimate coal/sorbent mixtures for SO_2 control in FBC applications which is reported in our 1978 EPA report, EPA-600/7-78-005[1]
- Report on the "Effect of SO_2 Emission Requirements on Fluidized-Bed Combustion Systems: Preliminary Technical/Economic Assessment," issued in August 1978 (EPA-600/7-78-163, NTIS PB 286 871/7ST).[2]

Work on the other tasks performed under this contract is reported in:

- Experimental/Engineering Support for EPA's FBC Program: Final Report Volume 1, Sulfur Oxide Control, EPA-600/7-80-015a, January 1980
- Experimental/Engineering Support for EPA's FBC Program: Final Report Volume II, Particulate, Nitrogen Oxide, and Trace Element Control, EPA-600/7-80-015b, January 1980
- Experimental/Engineering Support for EPA's FBC Program: Final Report Volume III, Solid Residue Study, EPA-600/7-80-015c, January 1980
- Alternatives to Calcium-Based SO_2 Sorbents for Fluidized-Bed Combustion: Conceptual Evaluation, EPA-600/7-78-005, January 1978
- Regeneration of Calcium-Based SO_2 Sorbents for Fluidized-Bed Combustion: Engineering Evaluation, EPA-600/7-78-039, NTIS PB 218-317, March 1978
- Disposal of Solid Residue from Fluidized-Bed Combustion: Engineering and Laboratory Studies, EPA-600/7-78-049 (NTIS PB 283-082), issued in March 1978, which presented the results of work performed from January 1976 to January 1977
- Evaluation of Trace Element Release from Fluidized-Bed Combustion Systems, EPA-600/7-78-050, NTIS PB 281-321, March 1978.

ABSTRACT

Engineering studies addressing several aspects of fluidized-bed combustion (FBC) system design and performance are reported. An evaluation on the impact of SO_2 emission requirements on FBC system performance and cost is reviewed (EPA-600/7-78-163). Stringent SO_2 emission requirements can be satisfied economically if proper selection of design and operating parameters is made. Another study on the feasibility of feeding coal/sorbent mixtures to FBC units is also reviewed (EPA-600/7-78-005). Critical data gaps exist for this concept. Moreover, general economic feasibility would not be expected. An alternative SO_2 control concept for pressurized fluidized-bed combustion (PFBC), that is, pressurized scrubbing of the products of combustion with water, is evaluated. The concept is not economically competitive because of the requirement for recuperative heating and reduced plant efficiency. A potential reduction in solid waste is realized with the concept, but the SO_2 control efficiency may be limited.

An evaluation of PFBC examining the technical and economic tradeoffs between the level of particulate control achieved and the frequency of gas-turbine blade replacement is described. The evaluation incorporates models of PFBC particulate carry-over, particulate control device efficiency, and turbine erosion. Also, an indirect air-cooled PFBC concept is evaluated and compared with other PFBC concepts. The indirect air-cooled concept provides significant particulate control advantages over the adiabatic combustor PFBC concept, while resulting in about a 4 percent lower plant efficiency and a 1 percent higher cost of electricity.

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NOMENCLATURE

AFBC - atmospheric-pressure fluidized-bed combustion

Ca/S - calcium-to-sulfur ratio

CBC - carbon burnup cell

CF - capacity factor

COE - cost of electricity

DOE - Department of Energy

ECAS - Energy Conversion Alternatives Study

EPA - Environmental Protection Agency

ERDA - Energy Research and Development Agency

FBC - fluidized-bed combustion

GBF - granular-bed filter

HHY -

HRSG - heat recovery steam generator

IERL - Industrial Environmental Research Laboratory

ISO - International Standards Organization

NO_x - nitrogen oxide

NSPS - New Source Performance Standard

O&M - operating and maintenance

PFBC - pressurized fluidized-bed combustion

SO_x - sulfur oxide

TDC - total direct costs

TGA - thermogravimetric analysis

ACKNOWLEDGMENT

We want to express our high regard for and acknowledge the contribution of Mr. D. B. Henschel who served as the EPA project officer. Mr. P. P. Turner and Mr. R. P. Hangebrauck, Industrial Environmental Research Laboratory, EPA, are acknowledged for their continuing contributions through discussions and support of the program.

We gratefully acknowledge the contributions of the following Westinghouse personnel: K. D. Weeks for his assistance in evaluating the air-cooled PFBC cycle and Dr. D. H. Archer, Manager, Chemical Engineering Research, for his program consultation and continued support.

1. INTRODUCTION

This volume documents five systems evaluation tasks that were performed during the contract as technical directives from the EPA project officer. One study, the effect of sulfur dioxide (SO_2) emission requirements on fluidized-bed combustion (FBC) systems, was issued as a separate report in 1978 (EPA-600/7-78-163). Another study, an engineering assessment of intimate coal/sorbent mixtures for SO_2 control by FBC applications, was also issued in 1978 (EPA-600/7-78-005). Results from three studies, the feasibility evaluation of pressurized-water scrubbing for SO_2 emission control with PFBC (completed in 1977), a particulate control/turbine life trade-off study for PFBC systems (completed in 1977), and an evaluation of indirect air-cooled PFBC concepts (completed in 1978) were not previously issued as separate contract reports.

A summary of the sulfur oxide (SO_x) control report is presented in Section 4, the intimate coal/sorbent mixture study is summarized in Section 5, and results of the remaining three studies are reported in Sections 6 through 8.

2. CONCLUSIONS

The primary conclusions from the five studies follow.

EFFECT OF EMISSION REQUIREMENTS ON FBC SYSTEMS

- AFBC and PFBC systems can economically meet the New Source Performance Standards (NSPS) for utility power plants: 90 percent sulfur removal, 12.9 ng/J (0.03 lb/MBtu) particulate emission, and 258 ng/J (0.6 lb/MBtu) nitrogen oxide (NO_x) emission.
- The selection of FBC design and operating parameters to minimize the sorbent feed requirement is critical for realizing economical systems.

INTIMATE COAL/SORBENT MIXTURES FOR SO_2 CONTROL

- Sufficient technical data on the performance of intimate coal/sorbent mixtures, such as pellets consisting of coal and limestone powders, do not exist to project FBC performance reliably.
- The performance of intimate coal/sorbent mixtures in FBC is unlikely to be economically competitive with conventional FBC concepts.

FEASIBILITY EVALUATION OF FLUIDIZED-BED COMBUSTION USING PRESSURIZED-WATER SCRUBBING

- A PFBC plant using the cold pressurized-water scrubbing concept for SO_2 control is not economically competitive with calcium-based PFBC or conventional steam power plants with stack-gas cleaning.

PARTICULATE CONTROL TRADE-OFF FOR PFBC SYSTEMS

- A methodology has been developed and is available for evaluating trade-offs between fluid-bed combustor, gas-cleaning, and turbine design and operating parameters.

INDIRECT AIR-COOLED FLUIDIZED-BED COMBUSTION CONCEPT SYSTEMS EVALUATION

- Indirect air-cooled PFBC concepts will have
  - Lower performance (higher heat rates) than PFBC boiler concepts
  - Similar performance with PFBC adiabatic combustor concepts.
- The cost of electricity for indirect air-cooled PFBC concepts is essentially the same as for PFBC adiabatic combustor concepts.
- Indirect air-cooled AFBC concepts will have lower performance (higher heat rates) than indirect air-cooled PFBC concepts. The AFBC concept provides for turbine reliability using a clean gas.
- Environmental emissions standards can be met with all indirect air-cooled FBC concepts.

3. RECOMMENDATIONS

The primary recommendations from work carried out under the technical directives are presented here, followed by recommendations for extended systems studies to evaluate and guide the development of economical FBC systems operating within environmental constraints.

- Investigate the ability of AFBC and PFBC processes to achieve more stringent emission standards, with specific focus on the relationship of performance to combustor design and operating parameters.
- Evaluate and develop advanced FBC sulfur removal concepts (e.g., sorbent pretreatment, sorbent regeneration, alternative regenerable sorbents). This will be particularly important when solids procurement or disposal represents a constraint.
- Develop understanding of NO_x minimization alternatives, perform system evaluation to select economic options, and demonstrate capability.
- Carry out experimental test programs to obtain performance data on particulate control equipment applicable to AFBC and PFBC systems. The primary need is an understanding of high-temperature, high-pressure particulate control equipment performance - e.g., cyclones, granular-bed filters, fabric filters, and other advanced filter and cyclonic concepts. Data will be important for process constraints (e.g., turbine erosion/deposition) and environmental constraints (leading and potential fine particle emission criteria).

- Carry out experimental test programs to obtain understanding of turbine tolerance - e.g., erosiveness of particulate, response characteristics of turbine materials, effect of turbine design and operating parameters.
- Extend and apply methodology for evaluating fluid-bed combustor/gas-cleaning/turbine design and operating parameter trade-offs to identify optimal fluidized-bed combustor systems for given application and environmental requirements.

Additional FBC systems studies are recommended to

- Assess the effects of potential NSPS on industrial FBC systems to aid EPA in developing standards.
- Project and evaluate the environmental performance of FBC system designs currently proposed by Department of Energy (DOE) contractors or commercial vendors to understand the status of these designs.
- Evaluate the impact of variable coal sulfur content and variable sorbent properties on the control of SO_2 emissions from FBC systems in order to quantify the effect of variable properties on sorbent consumption and system economics.
- Evaluate FBC unit start-up and turndown techniques with respect to environmental performance in order to identify superior techniques and performance sensitivity.
- Assess the technical/environmental performance of alternative FBC operating regimes (e.g., turbulent fluidization, circulating fluidized bed, fast fluidization, multisolids systems) to understand their potential and limitations.

4. SULFUR OXIDE CONTROL

Westinghouse evaluated the impact of up to 90 percent sulfur removal on the capital and energy costs of conventional dense-phase, fluid-bed, AFBC and PFBC power plants in a previous report as part of this contract. A brief summary of that report is presented here. The full study is presented in EPA-600/7-78-163.

Two levels of emissions standards were considered:

- The current (1978) EPA NSPS for large coal-fired boilers: SO_x, 516 ng/J (1.2 lb SO_2 MBtu); particulates, 43.0 ng/J (0.1 lb/MBtu); and NO_x, 301 ng/J (0.7 lb NO_2 MBtu)
- A set of more stringent degrees of control: SO_x, 90 percent removal of coal sulfur content; particulates, 12.9 ng/J (0.03 lb/MBtu); and NO_x, 258 ng/J (0.6 lb/MBtu).

These levels were selected for the study because they represent one set of values considered during the planned revision of the NSPS for utility boilers.

Projections of AFBC and PFBC power plant performance and economics have been developed through the assimilation of previous FBC power plant design studies, FBC performance models, and data assessments. The key parameters in the evaluation are the sorbent Ca/S ratio, the coal sulfur content, and the fluid-bed combustor design and operating conditions.

The projections of FBC power plant energy costs indicate that for both the existing SO_x emission standard and for 90 percent sulfur removal FBC is potentially cost competitive with conventional coal-fired power plants using lime-slurry scrubbing. The competitiveness of FBC depends upon proper selection of fluid-bed combustor operating conditions--i.e., sufficiently long gas residence in the bed (sufficiently low gas velocity and sufficiently deep beds) and sufficiently small sorbent particle size. This selection of variables will result in a larger combustor, but the cost savings resulting from decreased sorbent requirements would more than compensate for increased combustor costs.

In the design of FBC power plants one should emphasize maximization of fluid-bed combustor performance rather than minimization of the combustor cost through compact design. The combustor cost represents a small portion of the FBC power plant investment and is also relatively insensitive to changes in design and operating conditions. On the other hand the overall FBC power plant cost of electricity is strongly dependent on the combustor performance.

The Ca/S molar ratio--that is, the moles of sorbent calcium fed to the fluid-bed combustor divided by the moles of sulfur fed in the coal--is the single, most important performance factor relative to FBC power plant cost and performance for high-sulfur eastern coals (2 to 5 wt % sulfur). The Ca/S ratio has a dramatic impact on the FBC power plant thermal efficiency, capital investment, and cost of electricity. An increased Ca/S ratio, if required for lower SO_x emissions, results in increased auxiliary power consumption for solids handling and significant sorbent calcination energy losses. The resulting reduced net plant efficiency and slightly increased equipment costs for solids handling, crushing, drying, feeding, and spent solids disposal lead to increased capital investment and energy costs. In addition, the increased cost of raw sorbent at increased feed rates significantly increases the energy cost.

The all-important projection of sorbent feed requirements was accomplished in this study by using a kinetic model for SO_x capture Westinghouse had developed. This model--using rate constants measured in laboratory thermogravimetric analysis (TGA) equipment and confirmed where possible by using available data from experimental fluidized-bed combustors--is capable of projecting sorbent requirements, where TGA data have been generated, as a function of key combustor operating/ design conditions.

While the cost and performance of several subsystems in the FBC power plants are uncertain (for example, solids feeding and particulate control), these are expected to be resolved through proper design and specification of materials and operating conditions and maintenance and operating procedures. The overall financial impact of these cost/ performance uncertainties will probably be small relative to the uncertainties in such site factors as sorbent availability, sorbent cost, coal cost, solid waste disposal feasibility or utilization markets, local emission standards, and so on.

For low-sulfur western coals and lignites the impact of an increased Ca/S ratio is greatly reduced because of the relatively small quantities of sorbent involved. Uncertainties associated with sorbent selection and cost are also less significant.

Projections of particulate control and emissions of NO_x for FBC power plants indicate that the more stringent emission requirements considered here of 12.9 ng/J (0.03 lb/MBtu) and 285 ng/J (0.6 lb/MBtu), respectively, are economically feasible and of lower cost impact than the more stringent SO_x requirement. Conventional fabric-filter (baghouse) techniques should permit achievement of this requirement, depending on particle size and future environmental standards. We expect PFBC plants to require two stages of particulate control equipment operating at the combustor temperature and pressure: e.g., conventional cyclones followed by a filter system. Nitrogen oxide levels from the assessment of FBC experimental results have been shown to be generally lower than 258 ng/J (0.6 lb/MBtu) without special control efforts.

On the basis of available information, the projections developed indicate that both AFBC and PFBC should be able to achieve the higher levels of control considered in this evaluation economically if proper combustor design and operating conditions are selected. Development programs should focus on developing large-scale information on the relationship between combustor operating conditions and FBC plant emissions, while engineering evaluation should assess FBC pollution control capabilities.

The detailed conclusions and recommendations developed in this report are as follows:

- On the basis of available information, the more stringent emission requirements considered in this study (SO_x, 90% sulfur removal; particulates, 12.9 ng/J (0.03 lb/MBtu); NO_x, 258 ng/J (0.6 lb NO_2/MBtu) should be economically achievable for both AFBC and PFBC power plants.
- The proper selection of fluid-bed combustor design and operating conditions is critical to the economical realization of these environmental goals. The gas residence in the bed, in particular (as determined by gas velocity and bed height), should be sufficiently long, and sorbent particle size should be sufficiently small. In this assessment residence of 0.67 to 2.0 s (gas velocities of 1.5 to 1.8 m/s) and particle sizes averaging 500 μm appeared to offer effective SO_x removal performance, although these conditions are not necessarily optimal.
- The high level of SO_x emission control considered has a greater impact on the FBC power plant energy cost than do

the revised particulate and NO_x standards considered. The most critical process parameter with respect to FBC power plant cost and performance is the Ca/S ratio.

- The fluid-bed combustor cost does not depend strongly on changes in design and operating conditions. The fluid-bed combustor should be designed to minimize the cost of plant energy rather than cost of the combustor. For example, low - rather than high - fluidization velocities will probably result in lower FBC power plant energy cost.
- Particulate control to levels as low as 12.9 ng/J (0.03 lb/MBtu) should be economically achievable for AFBC using commercially available techniques. Baghouses seem most suitable for this duty. No testing of any type of final-stage particle control device on an AFBC unit, however, has yet been conducted.
- Particulate control to levels below 12.9 ng/J (0.03 lb/MBtu) may be dictated for PFBC by turbine protection requirements, depending on particle size. Projections indicate that 0.03 lb/MBtu should be achievable, but the technology to meet this control at high temperature and pressure has not yet been demonstrated.
- Oxides of nitrogen will generally be emitted by FBC at levels below the 258 ng/J (0.6 lb NO_2/MBtu) requirement considered in this evaluation. No direct control techniques for NO_x have been clearly demonstrated on fluidized-bed combustors to date, although several options are under study.
- The greatest FBC power plant uncertainties presently involve reliability questions - e.g., solids feeding, particulate control (especially for PFBC), material erosion/corrosion/deposition, and process control. The impact of emission standards averaging time basis and system reliability has not been evaluated.

- AFBC and PFBC development programs should focus on more stringent emission standards and their relation to combustor design and operating conditions.
- Advanced FBC sulfur removal concepts, for example, sorbent precalcination, sorbent regeneration, sorbent fines reconstitution, additives for improved sorbent utilization, alternative metal oxide sorbents, should be evaluated with respect to more stringent emission standards.

5. INTIMATE COAL/SORBENT MIXTURES FOR SO_2 CONTROL IN FLUIDIZED-BED COMBUSTION

Westinghouse performed a conceptual evaluation of the use of intimate coal/sorbent mixtures (e.g., pellets consisting of powdered coal and limestone) as part of this contract. This evaluation has been reported previously, in EPA-600/7-78-005, and a brief summary of that report is included here.

The study was carried out to investigate the technical and environmental feasibility and economic potential of "intimate coal/sorbent mixtures" when used in an FBC system for power generation. Various classes of intimate coal/sorbent mixtures were first qualitatively screened for feasibility on the basis of their probable performance assessment. Intimate coal/sorbent mixtures selected as potentially feasible in the initial screening were then subjected to an engineering assessment of technical and environmental performance. Areas such as SO_x and NO_x control, trace metal and particulate control, solid waste and plant efficiency, and design factors for the fluidized-bed combustor were considered. Because no actual performance or kinetic data exist for the intimate coal/sorbent mixtures, only potential performance could be addressed and problem areas identified.

Economic potential was examined by using optimistic performance assumptions for the intimate coal/sorbent mixture. Process alternatives for the preparation of the mixtures were identified and cost projections for the preparation systems were generated.

The major conclusions reached are as follows:

- The only technically feasible intimate coal/sorbent mixture that could be identified for the current fluidized-bed combustion design concept is the consolidated coal/sorbent particle concept.
- Attrition of the consolidated particle is the most critical factor influencing the performance and feasibility of the concept. Modifications to the combustor design would probably be required in order to apply the consolidated particle concept.
- The performance (technical and environmental) cannot be estimated without initiating a test program. The overall technical and environmental performance of the consolidated particle concept could conceivably by worse than or better than the conventional fluid-bed combustor, but it is highly unlikely that any significant improvement in performance is to be realized.
- Except under very extreme conditions, the consolidated particle concept will not be economically competitive with conventional FBC concepts.
- Washing the pulverized coal during consolidated particle preparation could reduce trace elements, ash, sulfur, and the sorbent requirement. The economics of this option have not been investigated.
- The most attractive consolidated coal/sorbent particle from the standpoint of technical and environmental impact would utilize a binder to maintain the coal-ash and sorbent particles in discrete, consolidated particles following combustion. A binder that will effect this behavior has not been identified.

6. FEASIBILITY EVALUATION OF FLUIDIZED-BED COMBUSTION USING PRESSURIZED-WATER SCRUBBING

INTRODUCTION

The national and private development efforts for fluidized-bed combustion are based on SO_x absorption by calcium-based sorbents (limestone or dolomite) at high temperatures. Both AFBC and PFBC concepts are being pursued with either regenerative or once-through sorbent operation. We believe that once-through sorbent operation represents only the first-generation of FBC systems, but even with sorbent regeneration, if it is eventually realized, FBC will produce significant quantities of dry, granular, sulfated limestone or dolomite that must be disposed of or utilized in an environmentally satisfactory manner.

We have evaluated an alternative FBC concept that may be applicable to PFBC. This concept is a cold gas-cleaning scheme that uses water to scrub the pressurized combustion products without additives for controlling SO_2. The potential advantage of this alternative is the reduction of solid-waste emissions.

A feasibility study has been conducted to better define the concept and to estimate its cost and performance. Approximate material and energy balances, conceptual equipment designs, and process economic estimates were performed in order to determine concept problem areas and process economic and environmental feasibility.

CONCEPT AND PROCESS OPTIONS

The basic PFBC concept with pressurized water scrubbing is shown in Figure 1 with various process options indicated. An understanding of these options is important if one is to select the best process to be evaluated.

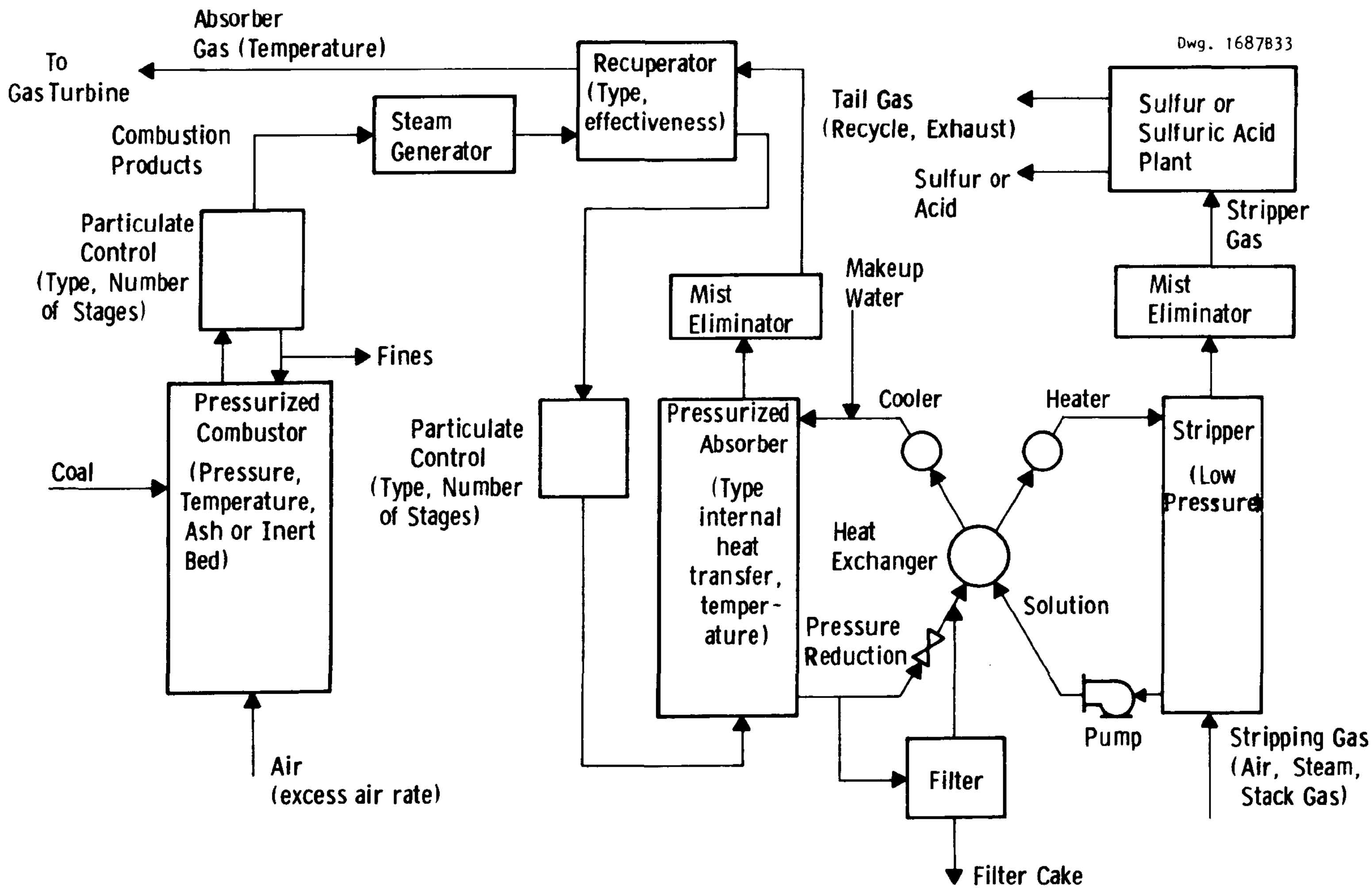

Figure 1 - Concept and Process Options

Coal is combusted with air in the pressurized fluidized-bed combustor. The bed consists of either coal ash or an inert bed material such as alumina. The bed temperature (760-1040°C) and pressure (620-1600 kPa) are important process variables relating to the combustor performance and the cycle efficiency. The excess air rate, also, is a critical process variable since it defines the quantity of gas that must be handled by the pressurized water scrubbing system. The excess air may range from 10 to 100 percent for fluidized-bed boilers and may be about 300 percent for an adiabatic fluidized-bed combustor (no heat transfer surface in the bed). The combustor fluidization velocity and heat transfer rates are assumed to be very similar to those of the calcium-based combustor, as are the attrition and elutriation rates, although they could be lower with proper selection of the inert bed material.

High-temperature particulate removal equipment (cyclones, filters) could be situated so as to operate before the combustion products are cooled, and/or low-temperature removal equipment (filters, scrubbers, electrostatic precipitators) could be placed to operate after the cooling step. Captured bed material (coal ash, inert material) could be recycled to the combustor or removed from the system.

Cooling and reheating the combustion products would be a critical step. A recuperator or a convection-type steam generator followed by a recuperator would cool the combustion products to a level suitable for the absorber (<150°C) and would reheat the absorber gas to a temperature resulting in an economical, combined-power cycle. Various types of recuperators could be used: a shell-and-tube heat exchanger constructed from high-alloy tube materials (bare or finned) and designed for high thermal expansion conditions, the more conventional, stove-type or packed-bed-type heat exchanger requiring cyclic heating and cooling of parallel vessels containing refractory material (packed-bed or checker structure) with gas flow controlled by high-temperature valves, or a circulating pebble-bed heat exchanger requiring continuous circulation of a refractory heat transport medium between parallel vessels.

Conventional countercurrent absorber and stripper towers would be used to remove the SO_x from the combustion products at pressure and to generate at atmospheric pressure an SO_2 gas suitable for elemental sulfur or sulfuric acid (H_2SO_4) recovery. Packed columns or a plate-type design could be applied with proper construction for the highly corrosive environment. An internal heating or cooling surface could be placed in the columns to control the column temperatures. Mist eliminators might be required in order to protect downstream equipment from corrosion.

The stripping gas could be either air, steam, or stack gas. Each would have advantages in terms of power requirements, oxygen content, and capital investment.

Elemental sulfur or H_2SO_4 could be recovered from the stripper gas. The composition of the stripper gas is critical to this step. A commercial sulfur recovery process such as Allied Chemical's could be applied. The Allied Chemical process requires the use of a clean fuel, such as methane (CH_4), for SO_2 reduction. Alternatively, a developmental process such as the Foster Wheeler RESOX Process, which uses coal as a reductant, could be applied. The tail gas from the sulfur plant could be exhausted or recycled to the absorber.

The circulating solution system requires heat exchange, cooling and heating with conventional devices in order to control the absorber and stripper temperatures. In addition to a pump to circulate the solution a means of pressure reduction such as a pressure reduction valve or a power recovery turbine would be required, since the absorber is operated at elevated pressure and the stripper is at low pressure.

Particulate material trapped in the absorber must be removed in order to maintain the absorber performance. Various commercial devices that will permit the filtration of a side stream of the circulating solution are available.

Makeup water would be fed to the system to account for filter cake losses and evaporation losses.

SELECTION OF BASE CASE DESIGN CONCEPT

A selection of a base design concept for the PFBC with pressurized water scrubbing from the options presented in the previous section has been made. We have judged, on the basis of preliminary considerations, that the selected base-concept would probably be the most successful of all of the concepts presented.

A previous cycle study for a PFBC concept that used low-temperature venture scrubbing as an alternative to high-temperature particulate control was applied to reach the following conclusions:[3]

- A combustor temperature of about 927°C (1700°F), resulting in a combustion product temperature of about 871°C (1600°F) to the recuperator, and a combustor pressure of about 1034 kPa (150 psia) are suitable combustor operating conditions for this concept.
- A recuperator effectiveness of at least 0.86 (resulting in a turbine inlet temperature of about 760°C (1400°F)) and an excess air rate of no more than 20 percent should be used for economic feasibility. These conditions will yield a plant heat rate of about 10,000 kJ/kWh (9,500 Btu/kWh), including the boiler efficiency increase due to the elimination of sorbent calcination energy losses and energy losses in the gas cleaning system. Using a steam generator prior to using the recuperator or using the high excess air fluidized-bed boiler or adiabatic combustor will not be economically feasible with this concept.

Other equipment and process selections are as follows:

- An inert ceramic bed (alumina) in the combustor because it should result in superior combustor performance in terms of particle elutriation and potential ash fusion
- Two stages of high-temperature particulate removal equipment (cyclones) located directly after the combustor. The first stage would recycle coarse material (alumina and carbon) to the combustor. The second stage would remove fines from the combustor products (coal ash and alumina) in order to protect the recuperator from erosion and deposition. These fines would be removed from the system. No low-temperature particulate control equipment would be used before water scrubbing, and we assumed that the absorber and stripper could tolerate a relatively high particulate content.

Because the recuperator is a critical process component, both the shell-and-tube recuperator and the cyclic stove recuperator have been evaluated. The packed bed concepts were not considered because of the possibility of particle elutriation and plugging. An effectiveness of 0.90 was selected with a turbine inlet temperature of 788°C (1450°F).

Valve-tray columns were selected for the absorber and stripper to improve performance under conditions of high particulate content and to permit simplified periodic maintenance of the columns. Plastic lining was specified to protect against corrosion. Preliminary calculations indicate that internal heating or cooling would not be required in the columns.

Stack gas would be used for stripping rather than steam or air. Steam consumes a large quantity of power and results in a large water loss. Stack gas contains a lower oxygen content than does air and results in less reductant consumption for sulfur recovery. Preliminary cleaning of the stack gas would be required in order to protect the blower.

Elemental sulfur would be recovered by using the Allied Chemical Process with CH_4 as the reductant. Developing technologies such as the Foster-Wheeler RESOX process have not yet been demonstrated and may not be very efficient or capable of producing commercial-grade sulfur. The sulfur recovery process tail gas would be exhausted on the basis of an assumed 90 percent sulfur recovery efficiency. On the basis of economic projections, an SO_2 content of at least 4 mole % would be required in the stripper off-gas.

A hydraulic turbine would be used for power recovery from the circulating scrubber solution. The solution would be cooled by a cooling-water exchange and heated by clean fuel (heating oil) combustion. Low-grade steam was considered for heating the solution, but the steam requirements exceeded the availability in the plant.

A typical design philosophy for large fluidized-bed combustion plants calls for modular design with four parallel combustors in a 600 MW_e power plant. This philosophy has been followed in this design evaluation that specifies parallel gas-cleaning trains.

PLANT BASIS

The following plant basis was selected for the evaluation:

- 594 MW_e power plant net output (635 MW_e conventional PFBC) power plant net output
- Four boiler modules
- 17.5 percent excess air in primary combustors
- 4 wt % sulfur coal with 10 wt % ash and a heating value of 30×10^6 J/kg (13,000 Btu/lb)
- SO_2 emission controlled to 0.5 kg SO_2/GJ (1.2 lb SO_2/10^6 Btu), equivalent to about 81 percent sulfur removal

- Single sulfur recovery plant with 90 percent sulfur recovery efficiency
- Absorber 89.5 percent efficient in removing SO_x.

This basis provides direct comparison with previous PFBC designs using calcium-based, high-temperature gas cleaning.

MATERIAL AND ENERGY BALANCES

Material and energy balances were performed for the base case described and are summarized in Figure 2 and Table 1.

An iterative approach was used for the absorber and stripper system material and energy balances in order to provide reasonably near optimum designs for these columns. The minimum absorber operating temperature possible [based on normal cooling water temperatures of 27-30°C (80-85°F), and considering the absorber inlet gas temperature of 121°C] is about 38°C (100°F). This temperature was selected for the design in order to yield the most efficient SO_x absorption. An operating temperature of 66°C (150°F) was selected for the stripper on the basis of maximizing the SO_2 concentration in the stripper gas and minimizing evaporative water losses. The maximum SO_x content of the stripper gas for this process operated with realistic temperature conditions is about 6 mole %. A value of 5 mole % was selected for the design in order to give reasonable column dimensions.

Energy balances around the absorber and stripper indicate that the solution circulation rate would be so great (361,725 kg-moles/hr) that heat of absorption effects, heat of evaporation effects, and sensible heats of entering gas streams would be negligible and the columns would operate isothermally.

A small amount of CO_2 would also be absorbed from the combustion products and released into the stripper gas (about 180 kg-moles/hr). The particulate content of the circulating solution was assumed to build

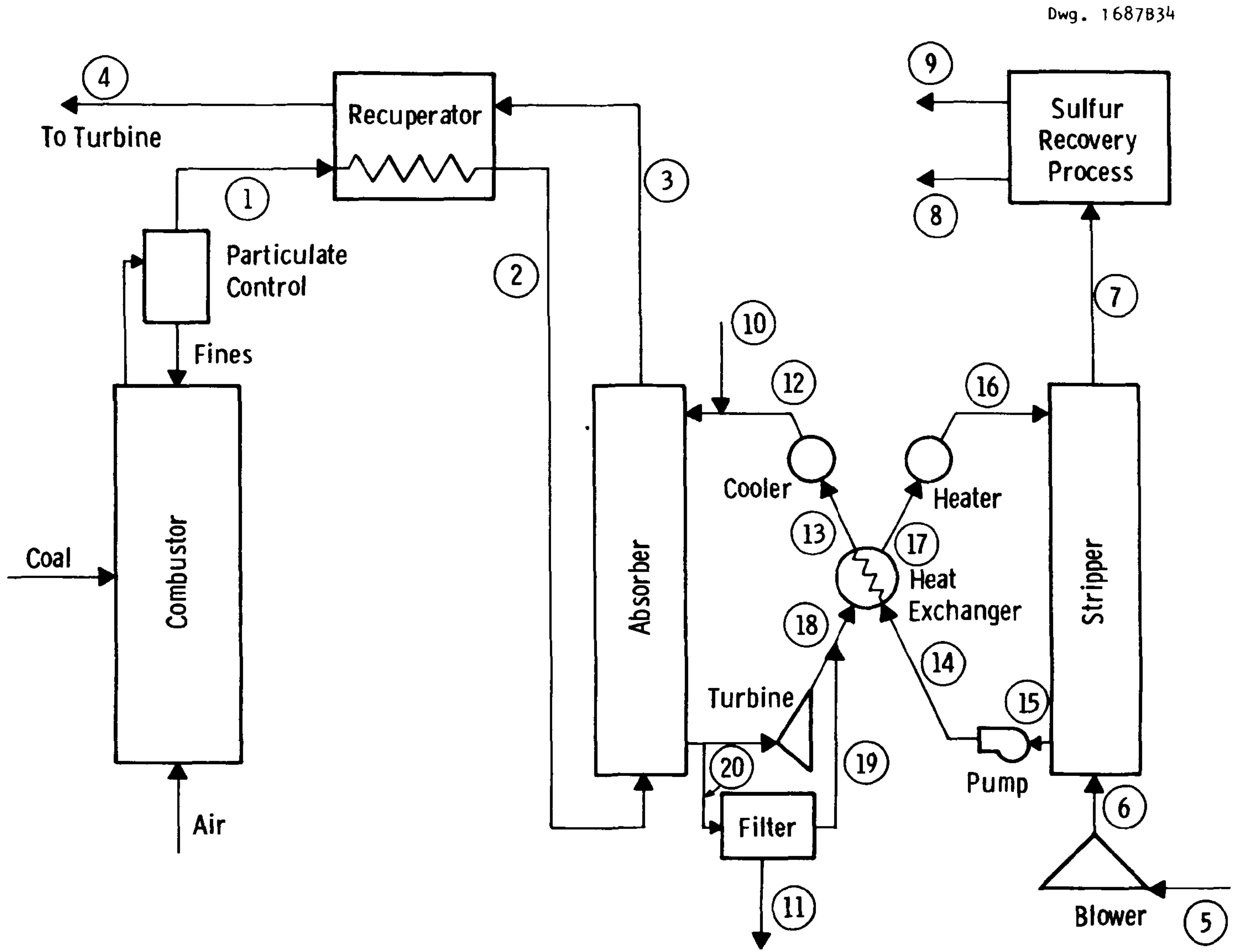

Figure 2 - Material and Energy Balances

Table 1

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MATERIAL AND ENERGY BALANCES
(594 MW_e Power Plant)

	Stream	Flow Rate, kg-moles/hr	Composition, mole %	Temperature, °C (°F)	Pressure, kPa (psia)
1	Combustion products	89,255	15% CO_2, 0.284% SO_2	871 (1600)	1034 (150)
2	Combustion products	89,255	15% CO_2, 0.284% SO_2	121 (250)	1014 (147)
3	Absorber gas	88,847	15% CO_2, 0.030% SO_2, 0.5% H_2O	38 (100)	965 (140)
4	Absorber gas	88,847	15% CO_2, 0.030% SO_2	788 (1450)	952 (138)
5	Stripper stack gas	3,454	4% O_2, CO_2, H_2O, N_2	121 (250)	103 (15)
6	Stripper air	3,454	4% O_2, CO_2, H_2O, N_2	177 (350)	172 (25)
7	Stripper gas	4,543	5% SO_2, 16% CO_2, 15% H_2O, 3% O_2, 61% N_2	66 (150)	110 (16)
8	Sulfur	204	Commercial grade sulfur	38 (100)	103 (15)
9	Tail gas	~5,260	0.43% SO_2	38 (100)	103 (15)
10	Makeup water	1,200	H_2O	27 (80)	965 (140)
11	Filter cake	2,722 kg/hr	95 wt% particulate, 5 wt% acid solution	38 (100)	103 (15)
12	Solution	361,725	0.004% SO_2 in water (0.5 wt% solids)	35 (95)	965 (140)
13	Solution	361,725	0.004% SO_2 in water (0.5 wt% solids)	43 (110)	1000 (145)
14	Solution	361,725	0.004% SO_2 in water (0.5 wt% solids)	66 (150)	1296 (188)
15	Solution	361,725	0.004% SO_2 in water (0.5 wt% solids)	66 (150)	172 (25)
16	Solution	361,725	0.067% SO_2 in water (0.5 wt% solids)	68 (155)	110 (16)
17	Solution	361,725	0.067% SO_2 in water (0.5 wt% solids)	60 (140)	117 (17)
18	Solution	333,773	0.067% SO_2 in water (0.5 wt% solids)	38 (100)	310 (45)
19	Filtrate	27,952	0.067% SO_2 (0 wt% solids)	38 (100)	310 (45)
20	Filter solution	27,952	0.067 SO_2 in water (0.5 wt% solids)	38 (100)	1034 (150)

up to about 0.5 wt % of particulate material. This would require that 8 percent of the circulating solution be continuously filtered in order to maintain a steady particulate level in the solution. The effects of particulate levels of this order on the absorber, stripper, hydraulic turbine, pump, and heat exchangers requires further investigation.

For the 4 wt % sulfur coal and the 90% sulfur recovery efficiency assumed for the sulfur plant, the overall process sulfur removal efficiency would be 80.5%, if we assume an absorber efficiency of 89.5 percent. Coals with a higher sulfur content would require higher sulfur removal efficiencies and greater solution circulations rates. Coals with less sulfur would result in less SO_2 in the stripper gas.

The auxiliaries (power, fuel, water) required for the pressurized water scrubbing process are listed in Table 2. Methane was used as the reductant in the sulfur recovery process and fuel oil for the circulating solution heater.

Table 2

GAS-CLEANING AUXILIARIES

Makeup Water	365 ℓ/min (96 gal/min)
Cooling Water	72,000 ℓ/min (19,000 gal/min)
Methane	5,200 m^3/hr (183,200 scf/hr)
Fuel Oil	5,000 kg/hr (11,000 lb/hr)
Power	3,870 kW
Pump	2,190 kW
Blower	2,520 kW
Turbine	-840 kW

EQUIPMENT DESIGN

The major equipment items for the gas-cleaning system are described in Table 3. Following the philosophy of maximum use of shop fabrication, we used a modular design which resulted in a single recuperator

Table 3

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Equipment	Number per Combustor	Description
Absorber	3	Plastic-lined vertical shell column, 3.7 m (12 ft) diameter, 32 m (105 ft) tall, 40 valve trays, 0.76 m (2.5 ft) tray spacing, 1400 kPa (200 psia) design pressure, with mist eliminators
Stripper	3	Plastic-lined vertical shell column, 4 m (12 ft) diameter, 37 m (120 ft) tall, 76 valve trays, 0.46 m (1.5 ft) tray spacing, 345 kPa (50 psia) design pressure, with mist eliminators
Recuperator	1	Shell-and-tube horizontal heat exchanger, floating-head design, 4 m (13 ft) diameter, 23 m (75 ft) long; 8,000, 2.5 cm (1 in) OD finned tubes with 0.089 cm (0.035 in) wall thickness, 15 m (50 ft) tube length, finned-tube-area-to-base-tube-area ratio = 10, Inconel or Incoloy construction, 1400 kPa (200 psia) design pressure
Alternative Recuperator	1	Two parallel vessels with internal refractory checker structure connected by high-temperature valves (4 per vessel); each vessel 6.7 m (22 ft) in diameter and 46 m (150 ft) long; each vessel containing 2.9×10^6 kg (6.5×10^6 lb) of refractory checker structure; 1-hr cycle time assumed
Circulating Solution Filter	3	Continuous-pressure drum filter, handles 42,000 kg (93,000 lb) of solution/hr, 29 m^2 (310 ft^2) filter area, 1000 kPa (150 psia) inlet pressure, 690 kPa (100 psi) pressure drop
Circulating Solution Turbine	3	Hydraulic turbine, handles 7,600 ℓ/min (2,000 gal/min), recovers 67 kW (90 HP)
Circulating Solution Pump	3	Centrifugal pump, handles 7,600 ℓ/min (2,000 gal/min), consumes 182 kW (244 HP)
Circulating Solution Heat Exchanger	3	Shell-and-tube heat exchanger, heat duty of 1.4×10^7 W (4.78×10^7 Btu/hr), tube surface of 1800 m^2 (19,000 ft^2)
Circulating Solution Cooler	3	Shell-and-tube heat exchanger, heat duty of 5.26×10^6 W (1.79×10^7 Btu/hr), tube surface of 560 m^2 (6,000 ft^2)
Circulation Solution Heater	3	Shell-and-tube heat exchanger, oil-fired, heat dury of 1.4×10^7 W (4.8×10^7 Btu/hr), tube surface of 1860 m^2 (20,000 ft^2)
Stripper Stack-Gas Blower	3	Centrifugal unit, stack-gas rate of 11,000 kg/hr (24,200 lg/hr), 216 kW (290 HP) motor power
Stack-Gas Recycle System	1	Baghouse, screw conveyor, airlock, lockhopper, valves, and booster fan; 570,000 ℓ/min (20,000 acfm) stack gas.

(4 boilers per 594 MW_e plant) and three parallel absorber/stripper gas cleaning trains per fluidized-bed boiler. A single sulfur recovery plant was used for the 594 MW_e power plant.

The absorber and stripper columns were designed by following design techniques and recommendations presented in the literature.[4-6] For simplicity we applied design relationships for dilute gas mixtures that assumed the validity of Henry's Law. These assumptions should be excellent for the absorber and reasonable for the stripper. Henry's Law constants of 40 and 80 were assumed for the absorber and the stripper, respectively.

The greatest uncertainty in design concerns the recuperators. Designs for two types of recuperators (shell-and-tube and refractory stove) were developed and are described.

The remaining items are essentially conventional devices modified for the corrosion protection required and for the particulate content of the acid solution they must handle. The large number of modules required for the process indicates potential economic limitations.

CAPITAL INVESTMENT

Equipment and total gas-cleaning process capital investments were estimated on the basis of the descriptions in Table 3. The results of these estimates are presented in Table 4 and based on mid-1977 dollars. Again, the greatest uncertainty surrounds the recuperator costs. We estimate that the total direct cost for the pressurized-water scrubbing system would be \$172/kW with the shell-and-tube and \$144/kW with the refractory stove recuperator design.

A breakdown of the boiler plant equipment costs for the dolomite-based PFBC plant and the water-scrubber-based PFBC plant is given in Table 5. A breakdown of the total power plant cost is given in Table 6. Both plants have the same coal-feed rate, identical combustor designs,

Table 4

CAPITAL INVESTMENT FOR PRESSURIZED WATER SCRUBBING[a]

Dwg.7718A30

Equipment	Purchased Equipment, $\$ \times 10^6$	Cost of Installed Equipment, $\$ \times 10^6$
Absorbers	2.9	8.9
Strippers	2.2	6.6
Recuperators		
Shell-and-tube	21.6	49.8
Refractory stove	16.8	31.2
Solution Filters	1.5	3.4
Solution Turbines	0.7	1.8
Solution Pumps	0.3	0.7
Solution Heat Exchangers	3.3	7.6
Solution Coolers	1.2	2.8
Solution Heaters	0.6	1.3
Stack Gas Blowers & Recycle System	1.2	2.2
Sulfur Recovery Plant	--	17.0
TOTAL DIRECT COST		102.2 (shell-and-tube), 83.6 (refractory stove)

[a]Basis: mid-1977 dollars; 594 MW_e plant

and identical combustion product flow rates. The dolomite-based PFBC plant produces 635 MW_e of electrical energy, but the water-scrubber-based PFBC plant produces 594 MW_e of electrical energy because of lower plant efficiency. Costs have been taken from previous Westinghouse PFBC cost studies and updated to include particulate cleaning equipment and escalation.[3,7]

We estimate that the total power investment for the PFBC with calcium-based, high-temperature gas cleaning would be \$423/kW. This

Table 5

PFBC BOILER PLANT EQUIPMENT COSTS

Equipment	Dolomite-Based System, $/kW	Water-Scrubber Based Systems, $/kW[a]
Steam Generator	20.89	22.33
Draft System		
Particulate removal	45.78	19.66
Draft flues and ducts	2.39	2.55
Piping	3.99	4.27
Stack and foundation	0.68	0.73
Coal- and Sorbent-Handling and Feeding Equipment	21.54	13.03
Ash- and Dust-Handling Systems	2.23	1.23
Stack-Gas Cleaning System	---	172.05 (140.74)
Instrumentation and Controls	4.47	4.78
Miscellaneous Equipment	1.36	1.45
	103.33	242.08 (210.77)
Net Plant Electrical Output	635 MW_e	594 MW_e

[a]System with refractory stove recuperator is in parentheses; system with shell-and-tube is shown to its left.

cost is based on mid-1977 dollars, 635 MW plant capacity, once-through operation with dolomite, 17.5 percent excess air, and three stages of particulate control equipment (final-stage, granular-bed filter).

The PFBC with pressurized-water scrubbing for SO_2 control would cost about $639/kW with the shell-and-tube recuperator and $594/kW with the refractory stove recuperator. This estimate is based on mid-1977 dollars, 594 MW plant capacity, 17.5% excess air, two stages of particulate control equipment (high-temperature cyclones), a combustor

Table 6

PFBC POWER PLANT COST BREAKDOWN

Item	Limestone-Based System, \$/kW	Water Scrubber Based-System, \$/kW[a]
Land and Land Rights	1.63	1.74
Structures and Improvement	28.21	30.16
Boiler Plant Equipment	103.33	242.00 (210.77)
Gas Turbine-Generator Equipment	21.33	22.80
Steam Turbine Generator Equipment	63.62	68.01
Electric Plant Equipment	22.93	24.51
Misc. Plant Equipment	5.13	5.48
Undistributed Costs	40.86	43.68
Other Plant Costs	4.19	4.48
Subtotal	291.23	442.94 (411.63)
Normal Contingency	17.47	26.58 (24.70)
Subtotal	308.70	469.52 (436.33)
Escalation	57.88	88.03 (81.81)
Subtotal	366.58	557.55 (518.14)
Interest during Construction	48.13	73.20 (68.03)
General Items and Engineering	7.93	7.93
TOTAL CAPITAL COST	422.64	638.68 (594.10)

[a]System with refractory stove recuperator is in parentheses; system with shell-and-tube is shown to its left.

cost identical with the high-temperature, gas-cleaning case, and a reduction of $10/kW to account for the elimination of dolomite-handling equipment.

A conventional, coal-fired steam power plant with limestone scrubbing for SO_2 control would probably cost from $500 to 570/kW.

The alternative PFBC system using venturi scrubbing for particulate control is estimated to cost between $459 and 502/kW based on 17.5 percent excess air, two stages of high-temperature particulate removal equipment, and mid-1977 dollars (Reference 3, Appendix A).

The option of applying the RESOX process for sulfur recovery to the PFBC with water-scrubbing for SO_2 control in place of the commercially available Allied Chemical process would probably increase the capital investment further because of the low sulfur-recovery efficiency expected with a 5 percent SO_2 gas. The basic RESOX plant would cost about the same as the Allied Chemical process, but the tail-gas cleaning plant for the RESOX process (Beavon process, for example) could easily cost another $20 to 30/kW based on a sulfur recovery efficiency of 50 to 60 percent.

An estimate of the most optimistic case for the pressurized-water scrubbing concept for PFBC was also developed. If the minimum modular design is used (a single gas-cleaning train per combustor module) with no increase in plant construction time, the refractory stove recuperator design, a coal-fired solution heater, and a RESOX sulfur recovery plant (assumed to have 90 percent sulfur-recovery efficiency), the total power plant capital investment would be reduced to $577/kW instead of the more realistic case of $591/kW to 635/kW.

In these capital cost estimates we have assumed off-site disposal of waste solids. In the dolomite-based PFBC system, the bed overflow and collected fly ash would be conveyed dry to on-site storage silos. A similar system for collected fly ash would be used in the water-scrubber-based PFBC system, where no accumulation of coal ash in the inert combustor bed was assumed. The filter cake is also disposed of off site, handled with slurry techniques similar to those used on FGD sludge. Disposal cost is counted as an operating cost, accumulated within the cost of electricity.

COST OF ELECTRICITY

The costs of electricity generated by PFBC with calcium-based SO_2 control and with pressurized-water scrubbing SO_2 control are developed and compared in Table 7. The energy cost of a conventional coal-fired power plant with limestone scrubbing is also shown.[3,7] The basis on which these costs are derived is listed in the table.

The energy cost associated with the pressurized-water scrubbing concept is projected to be 3.8 to 5.0 mills/kWh greater than the calcium-based fluidized-bed combustion power plant energy cost and 2.7 mills/kWh greater to 0.4 less than a conventional power plant energy cost. For the most optimistic case previously defined the total energy cost would be 23.7 mills/kWh.

The cost of disposing of waste solids and liquids does not contribute significantly to the cost of electricity for the disposal costs assumed. If higher costs should occur in the future (say >$20/Mg) the water-scrubber PFBC concept could result in competitive costs of electricity.

Also, should the cost of sorbent increase significantly (say to >$20/Mg) then the water-scrubber PFBC concept could provide economic incentive for development.

ENVIRONMENTAL COMPARISON

The environmental performance of calcium-based PFBC and of PFBC with pressurized-water scrubbing for SO_2 control are compared in Table 8. The concepts are expected to be comparable with respect to

Table 7

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COMPARISON OF COST OF ELECTRICITY[a]

Item	PFBC: Calcium-Based SO_2 Control	PFBC: Pressurized-Water Scrubbing SO_2 Control		Conventional Power Plant: Limestone Scrubbing
		Shell-and-Tube Recuperator	Refractory Stove Recuperator	
Capital Investment, $/kW	423	639	594	500 - 570
Energy Cost, mills/kWh				
Capital charges	10.4	15.6	14.5	12.2 - 13.9
O & M	1.1	1.7	1.6	1.5 - 1.7
Fuel (coal)	6.8	7.2	7.2	7.2
Sorbent	1.5	--	--	0.6
Auxiliary fuel	--	0.6	0.6	-
Makeup water	--	0.1	0.1	<0.1
Cooling water	--	<0.1	<0.1	-
Solid/liquid waste disposal	0.6	0.2	0.2	1.2
TOTAL	20.4	25.4	24.2	22.7 - 24.6

[a]Basis: Capital charges 15% of capital investment per year
Capacity factor 70%
O & M 2.36% of capital investment per year
Sorbent (dolomite and limestone) at $10/Mg
Ca/S of 2.0 for fluid-bed combustion, 1
Coal at $0.80/GJ ($0.76/10^6 Btu)
Cooling water at 0.5¢/10^3 ℓ (24¢/M gal)
Process water at 5¢/10^3 ℓ (20¢/M gal)
Methane and fuel oil at $1/GJ ($0.95/10^6 Btu)
Dry solid disposal $4/Mg
Sludge disposal at $10/Mg

NO_x, particulates, and heat rejection. The high-temperature, calcium-based, gas-cleaning process, however, has the greater potential for SO_2 emissions lower than the current standard of 0.5 kg/GJ (1.2 lb/10^6 Btu). The PFBC water-scrubber concept is probably limited to sulfur removal efficiency less than 90 percent because of limited SO_2 solubility in water and the limited efficiency of sulfur recovery in commercial and developmental sulfur recovery processes. NO_x emissions from the inert-bed combustor could conceivably be less than or greater than the calcium-based combustor NO_x emissions; factors such as catalytic effects from calcium compounds or alumina particles and the influence of high-versus-low combustion gas SO_x content may affect the formation/decomposition of NO_x in the combustor. Differences in particulate emissions between the two cases are also possible because of the elutriation of sorbent fines from the combustor in the dolomite-based PFBC. Particulate standards should be achievable with both concepts with properly selected equipment.

The solid wastes associated with the high-temperature, calcium-based, gas-cleaning process would be larger in mass than those for the pressurized-water-scrubbing concept by a factor of 2.4 if all forms are considered, by a factor of 6 to 20 if the waste sorbent is compared to the filter cake and attrited alumina only. The difficulty of handling the waste solids, however, and the environmental impact of these wastes would not necessarily be directly proportional to mass, and further processing of the filter cake waste would be required. The environmental impact and exact nature of the filter cake material is unknown, but we expect that this material could be handled by methods applied for corrosive wastes in the chemical industry, with neutralization being potentially acceptable.

With respect to waste liquids, makeup water consumption, clean fuel consumption, and the plant heat rate the high-temperature gas-cleaning process appears superior to the pressurized-water-scrubbing concept for

Table 8

Dwg. 1711B92

ENVIRONMENTAL COMPARISON (600 MW_e Power Plant)

	PFBC with Calcium-Based SO_2 Control	PFBC with Pressurized-Water-Scrubbing SO_2 Control
SO_2, kg/GJ (lb/10^6 Btu)	< 0.5 (< 1.2)	0.5 (1.2)
NO_x, kg/GJ (lb/10^6 Btu)	< 0.3 (< 0.7)	Probably < 0.3 (< 0.7)
Particulate, kg/GJ (lb/10^6 Btu)	< 0.04 (< 0.1)	< 0.04 (< 0.1)
Heat Rejection, % less than conventional plant	~ 15	~ 14
Waste Liquids	None	Small amount with filter cake; highly corrosive acid solution
Waste Solids, Mg/hr/MW (fraction of coal feed mass)		
Total	0.142 (0.458)	0.059 (0.177)
Coal ash	0.031 (0.10)	0.033 (0.10)
Sorbent	0.111 (0.358)	None
Sulfur	None	0.011 (0.032)
Others	None	0.005 (0.015) filter cake, 0.001 - 0.01 (0.003 - 0.030) attrited alumina
Makeup Water Consumption, ℓ/min	None	365
Clean Fuel Consumption		
Methane, 10^3 ℓ/hr	None	5,200 } Equivalent to 7% reduction in fuel efficiency
Fuel oil, kg/hr	None	5,200 }
Plant Heat Rate, kJ/kWh (Btu/kWh)	9380 (8892)	10,023 (9,500) - not including auxiliary fuel consumption

SO_2 control. The amount of process water circulated in the water-scrubber concept is about ten times as much as in a conventional plant with limestone scrubbing.

For the most optimistic case the consumption of clean fuels would be reduced to zero, significantly more coal would be consumed (~7%) and the emission of particulates and SO_2 would be expected to increase slightly. The solid waste generation would increase by about 0.003 Mg/hr/MW due to increased coal ash.

The high energy consumption of the PFBC water-scrubber concept is of particular concern.

CONCLUSIONS

The PFBC power plant utilizing the cold gas-cleaning (water scrubbing) concept for SO_2 control is not economically competitive with calcium-based PFBC or conventional steam power plants unless the cost of sorbents and/or waste solids disposal is substantially increased above the costs assumed in this study. There appears to be no alternative that could significantly improve the cold gas-cleaning concept economics. The recuperative heat exchanger is technically an item of great uncertainty and may limit the concept feasibility.

Environmentally, the pressurized-water-scrubbing concept could eliminate the massive amount of sulfated dolomite waste generated by PFBC, but the nature of the environmental effect of the waste filter cake produced in the process is uncertain. The energy conversion efficiency of the water-scrubber concept is very poor, and the sulfur removal efficiency associated with the concept is limited to less than 90 percent.

7. PARTICULATE CONTROL TRADE-OFF FOR PFBC SYSTEMS

OVERVIEW

Work is reported in Volume 2 (EPA-600/7-80-015b) that provides perspective on determining the impact of emission requirements, fluidized-bed combustor design and operating conditions, and turbine performance constraints on PFBC particulate control requirements and plant economics. An EPA technical directive was performed in 1978 to assess the effect of final-stage filter performance and the effect of filter staging on particulate loading and size distribution emitted and the resulting implications on turbine life. This work provided a basis for the subsequent analysis presented in Volume 2.

The results from the technical directive study are reported since they illustrate the methodology used to project turbine blade life and electrical energy costs as a function of different particulate loadings to the turbine using the Westinghouse turbine erosion and particle profile models. The projections of the particulate loadings and size distributions used here were based on methods employed prior to the development of the particle profile model described in detail in Section 5 of Volume 2. The model originally used for particle profile projections was not capable of including the effects of a recycle cyclone. The PFBC configurations discussed in this analysis, therefore, are limited to the carbon burnup cell (CBC) concept for high carbon utilization. The turbine erosion model has similarly been extended with additional understanding and experimental determination of the important model parameters.

This work is important in that it documents the early analysis and represents the basis for developing the tools that can provide perspective on important trade-offs and permit the design of reliable PFBC plants that operate within environmental constraints.

BACKGROUND

A substantial amount of work has been performed at the Westinghouse R&D Center to analyze the flow trajectories and erosion effects of particulates entrained in the flow stream of gas turbines. Reference 8 predicts quantitative turbine erosion rates for a typical 65 MW utility gas turbine expander by combining three major complementary analyses. These are the calculation of the inviscid dimensional flow stream through the blading; the calculation of the trajectories of particles entrained in the blading flow stream; and the calculation of the erosive effects of those particulates whose calculated trajectories result in an impact with the blade surfaces.

We must emphasize that the analysis in Reference 8 did not include the effect of particulate deflection and velocity reduction due to profile-boundary layers on the blading surfaces. Note also that the author cautioned that the erosion model used was based on data available in the literature, which at best is sketchy.

On the basis of the observations of actual erosion patterns on experimental coal-burning gas turbines,[9,10] we have concluded that secondary flow phenomena have a substantial effect on the trajectories of small (i.e., <10 μm) particles. Accordingly, an analysis was carried out[11] that includes the effects of viscous boundary layers on the trajectory of small particles. The results of this analysis confirmed qualitatively the tendency of the boundary layer flows to concentrate the particulates in certain regions of the gas turbine. Because of the complexity of the problem, however, quantitative erosion rates were not calculated.

Another phenomenon that is of interest, particularly for very high-temperature turbine applications, is the temperature reduction of the particle as it penetrates the boundary layer next to a cooled blade surface. To determine the temperature history of a particle as it passes through the boundary layer one must also calculate the trajectory velocity history.

ESTIMATION OF PARTICLE LOADING/SIZE DISTRIBUTION

The works reported above are related to the erosion caused by particles after they enter the turbine. One must, of course, determine the concentration and size distribution of the particles that enter the turbine. These parameters are determined by the particulate removal equipment installed in the hot gas stream between the turbine and the pressurized fluidized-bed boiler in which are generated both the hot gas stream and the entrained particulates. (Essentially the same particulate removal equipment would be used if the hot gas stream source were an adiabatic fluidized-bed combustor or a fluidized-bed gasifier.) The performance of the particulate removal equipment depends upon the concentration, density, and size distribution of the entering particles, and the detailed design of the equipment would depend upon these factors. The general approach, however, is to remove the bulk of the large particles in the early stage(s) and remove the fines in the later stage(s). The performance and cost of a particulate removal system in a plant with a pressurized fluidized-bed boiler have been described in some detail.[12]

Figure 3 is a flow diagram of the separation equipment arrangement designed for the present study. Figure 4 shows a typical arrangement of the particulate removal equipment and the piping connecting the pressurized fluidized-bed boiler to the gas turbine. As Figure 3 indicates, the effluent from the pressurized fluidized-bed boiler first enters a cyclone separator, a relatively inexpensive component that can handle high particulate loadings and has a much higher removal efficiency for the larger particles than for the smaller.

The particles collected by the primary cyclone separator contain a large portion of char (carbon), which is fed to a CBC to recover the chemical heat of combustion. Additional combustion air is fed to the CBC to complete the reaction. The volume of flow through the CBC is small compared to that of the main flow stream, so the exit stream is

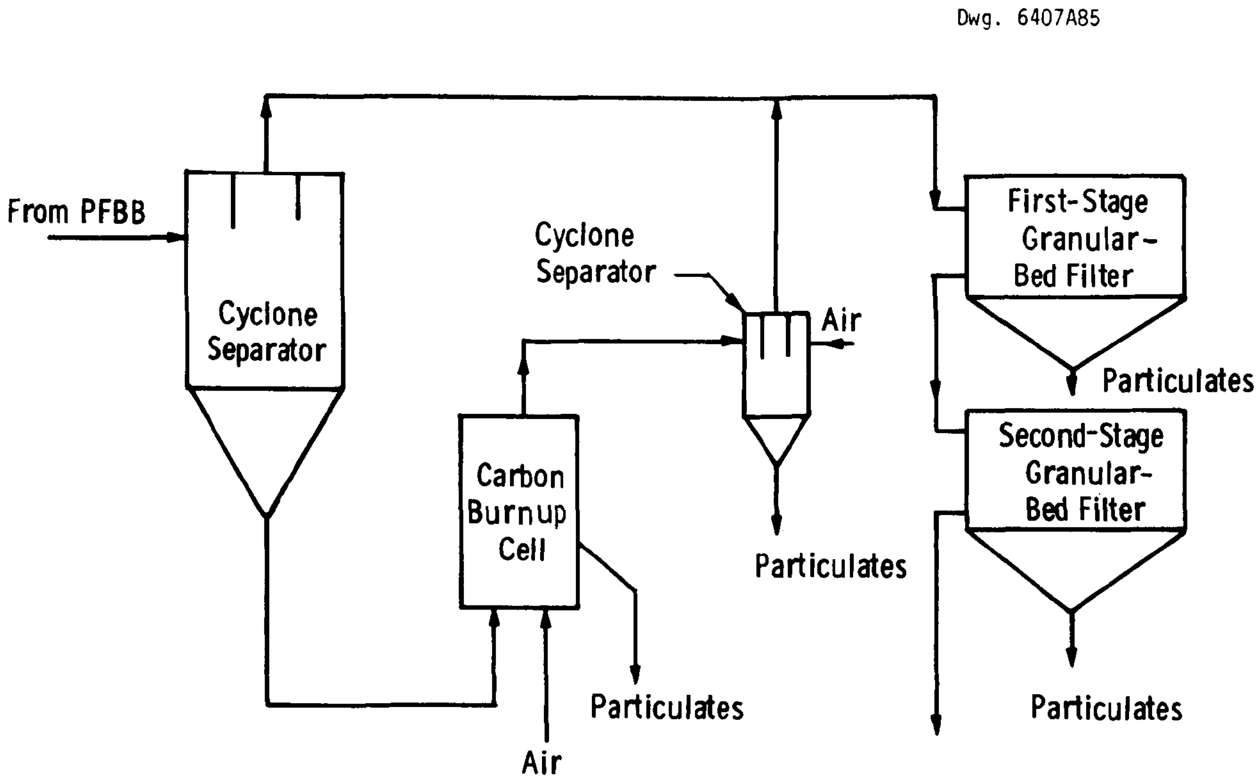

Figure 3 - Diagram of Particulate Removal System

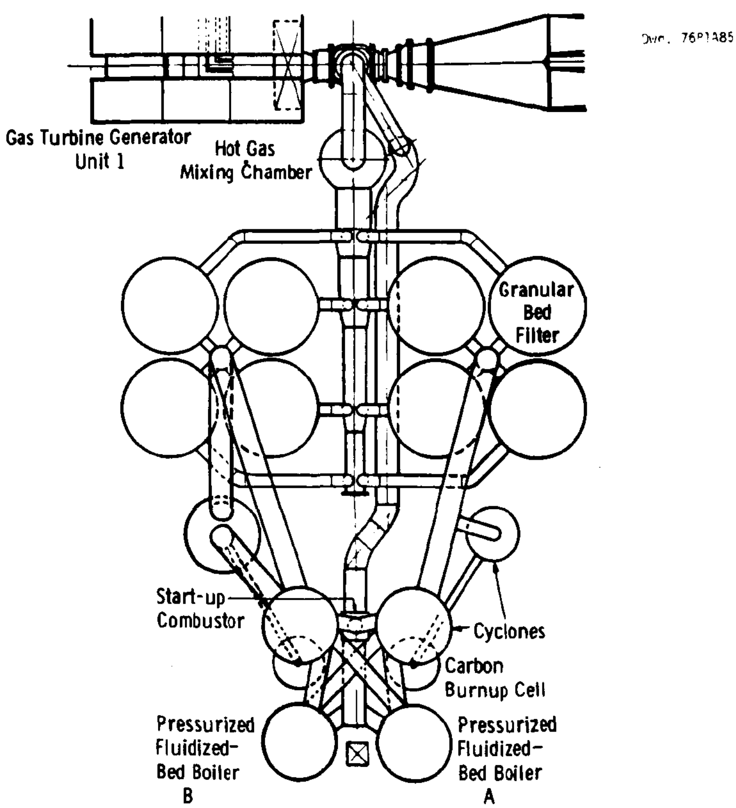

Figure 4 - Particulate Removal Equipment Arrangement[12]

passed through a small cyclone separator before being introduced into the main flow. This cyclone is designed to accommodate the very high particulate loadings leaving the CBC and is equipped with secondary inlet air jets that prevent fouling of the unit and provide angular momentum for the particle separation process.

The combined flow stream then enters the granular-bed filter, which is capable of high removal efficiency for particles smaller than 10 μm. Two stages of granular-bed filters are shown in Figure 3. Since the granular-bed filters are by far the most expensive component of the particulate removal system, this study will investigate whether the second stage is economically justified.

An analysis made specifically for the present study has been used to calculate the particulate concentration and size distribution at each location within the particuate removal system. The analysis considers the sorbent, the ash, and the char separately, since each of these contituents has a different effective density, an important factor in the efficiency of a particulate removal component.

The design conditions of the fluidized-bed boiler used to calculate the rate and size of the elutriated particles is as follows:

Superficial bed velocity	1.52 m/s (5 ft/s)
Bed depth	2.4 m (8 ft)
Pressure	1013 kPa (10 atm)
Bed temperature	1010°C (1850°F)
Sorbent	Dolomite
Sorbent feed size	3.2 x 0 mm (1/8 x 0 in)
Coal feed size	6.4 x 0 mm (1/4 x 0 in)
Excess air	Primary bed, 6%; CBC 36%
Ca/S atom ratio	1.5:1

Carry-over from the bed was estimated by calculating the size of the particle whose terminal velocity was equal to the superficial velocity in the combustor. All of the feed material below this size was assumed to be elutriated. Efforts to include sorbent attrition in the calculation involved the use of an average attrition rate that was a function of bed velocity and residence time. A more detailed account of the calculations involved in computing the carry-over characteristics is presented elsewhere.[2]

The performance of the primary conventional cyclone is shown in Figure 5 as a function of species density. These estimates were made on the basis of a commercial vendor's estimates of performance at pressure and temperature. The cyclone pressure drop is estimated at 165 cm wg at an inlet velocity of 17 m/s.

The performance of the CBC cyclone (Figure 6) is similarly based on the manufacturer's estimates for the elevated temperature and pressure. The specific device considered was a Donaldson Tan Jet(TM) cyclone operating with a primary flow pressure drop of about 180 cm wg. This device also employs a high-pressure, secondary flow of clean gas to impart the rotary motion to the dust-laden gases entering the device. A schematic of this device is presented in Figure 7.

The granular-bed filter concept used to establish the costs of the final filter system is basically the same Ducon filter that had been the basis of cost estimates for the previously published ECAS[12] report. Figure 8 presents a schematic of the filter system. In the Westinghouse design relatively few, but large (7-8 m dia), pressure vessels were used to effect a cost savings over a large number of small modules.

Although the performance of cyclone separators is known to a reasonable degree of accuracy, the performance of the granular-bed filter - because it is still in an early stage of development - is poorly defined. In the long range granular bed filters may achieve a removal efficiency equal to that currently achieved by conventional low-temperature fabric filters. At present, Westinghouse has bench-test results of a granular-bed filter whose performance is not as good as

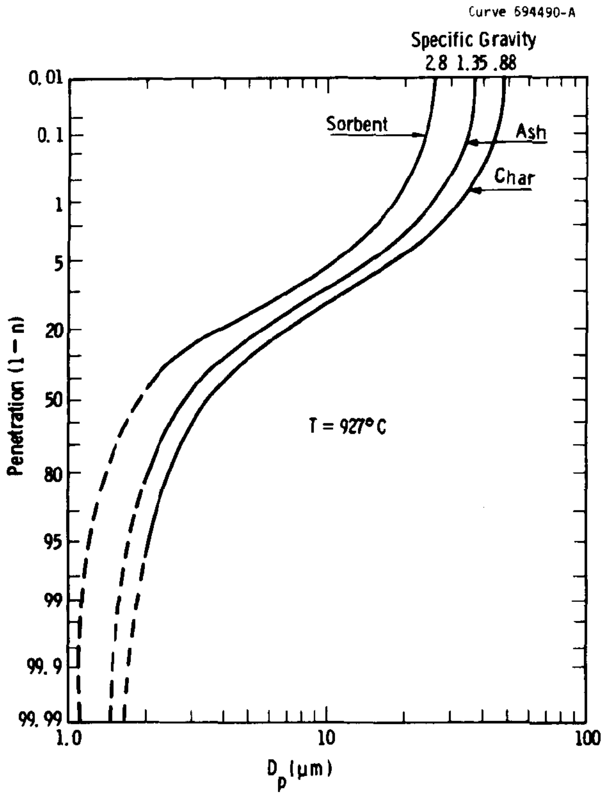

Figure 5 - Grade Efficiency of Primary Cyclone Separator

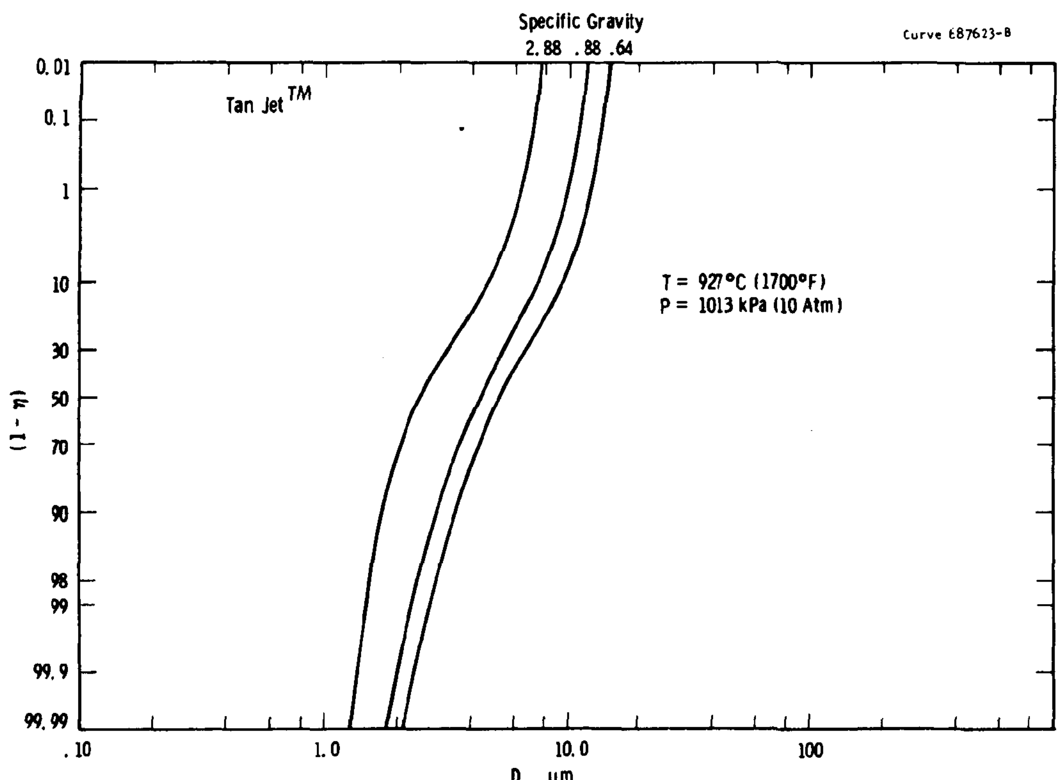

Figure 6 - Grade Efficiency Curve for Tan Jet

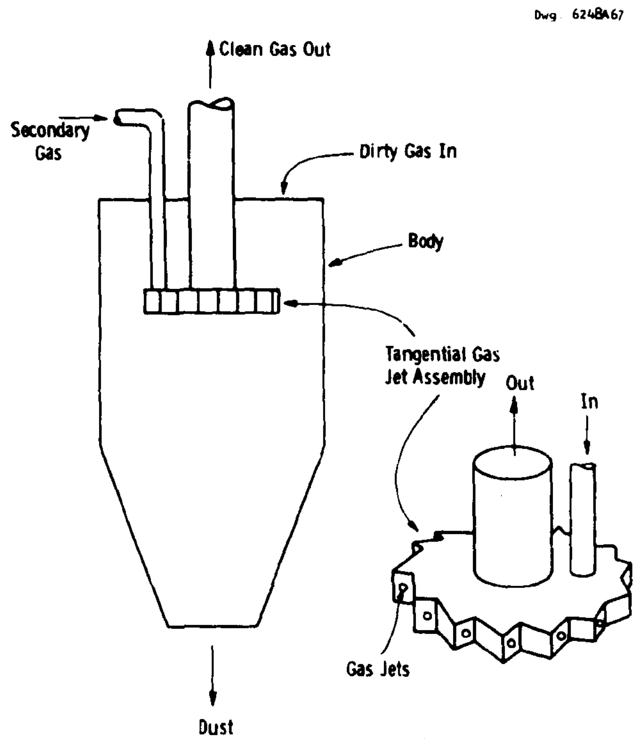

Figure 7 - Schematic of Tan Jet

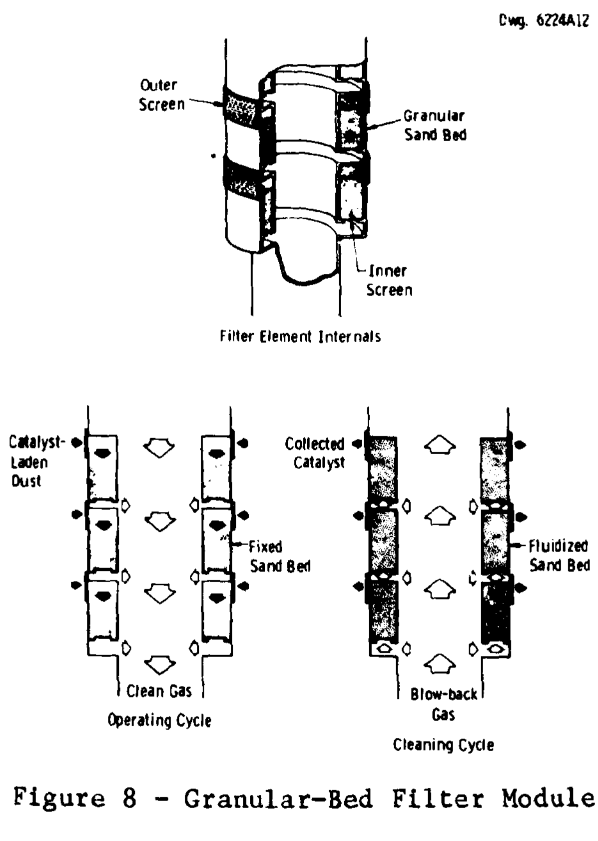

Figure 8 - Granular-Bed Filter Module

that of a fabric filter. As a third benchmark, results are available from the EPA[13] on an existing commercial unit (Rexnord) whose performance is substantially inferior to that of fabric filters.

In an attempt to bracket the actual situation, the calculations presented here have been based on the grade efficiencies of each of the three cases mentioned above. The actual grade efficiencies used for the fabric filter and the Westinghouse granular-bed filter performance are pre-sented in Figure 9 along with the Rexnord data.

Using these three efficiency criteria, we have calculated the concentration of particulate at the outlets of the first and second stages of granular-bed filters, and these are shown in Tables 9 and 10. A more graphic illustration of these performances is shown in Figures 10 and 11. Figures 12 through 20 show the size distribution of the particles leaving these filters. Similar information at the discharge of the cyclone separators has not been included since preliminary analysis showed that the turbine erosion rates corresponding to these particulate loadings were so high that a system not including granular bed filters would be impractical.

ESTIMATION OF GAS TURBINE IMPACT

On the basis of the information available in all of the above references, a relatively simple method was conceived with which to calculate the erosion rate of gas turbine blading as a function of particle distribution and concentration. A parametric, quantitative assessment of power generation cost penalties has been determined from those data on erosion rate, practical wear limits, and the cost of blading replacement (compared to the cost of particulate removal equipment as a function of its removal efficiency).

Reference 8 shows that maximum erosion occurs at the leading and trailing edges of rotors and at the trailing edge of stators. Although

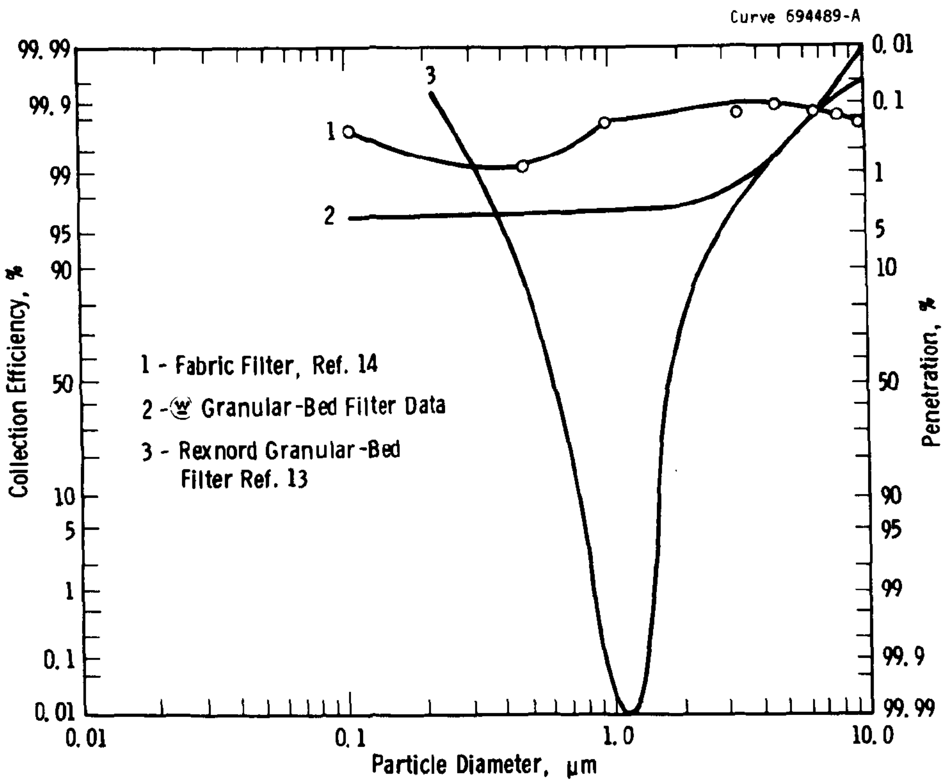

Figure 9 - Various Filter Performances Assumed for Final Cleanup Stage

these rates are roughly comparable in magnitude, the calculations neglect the effect of blade profile boundary layers. Since the boundary layer thicknesses are larger at the trailing edge than at the leading edge, the actual erosion rates will be reduced a greater amount at the trailing edge. This is particularly true for the particulates that pass through a granular-bed filter since they have a large proportion of very small particles that are readily deflected and slowed by the boundary layer velocity profile. This general reasoning is supported by the experimental evidence[9], which shows the maximum erosion at the leading edge of the rotor blades. On the basis of these considerations, we have made calculations for this analysis only for the rotor leading edge.

The calculation procedure is essentially a stepwise integration of the erosion of each particle size over the range of particle sizes entering the gas turbine for each of the three types of particulates.

Table 9

PROJECTED PARTICULATE CONCENTRATION LEVELS

Projected Based Upon	Location, Outlet of	Dolomite Concentration, g/sm^3 (gr/scf)	Ash Concentration, g/sm^3 (gr/scf)	Char Concentration, g/sm^3 (gr/scf)
Rexnord Granular-Bed Filter	First GBF	0.1326 (0.05795)	0.2179 (0.09518)	0.0398 (0.01737)
Rexnord Granular-Bed Filter	Second GBF	0.0866 (0.03783)	0.1243 (0.0543)	0.0244 (0.01065)
W Bench Tests Granular Bed Filter	First GBF	0.00787 (0.00344)	0.0138 (0.00601)	0.00240 (0.00105)
W Bench Tests Granular Bed Filter	Second GBF	0.000214 (0.0000935)	0.000290 (0.000127)	0.0000559 (0.0000244)
Conventional Fabric Filter*	First GBF	0.00103 (0.000449)	0.00203 (0.000886)	0.000353 (0.000154)
Conventional Fabric Filter*	Second GBF	0.000000 (0.000000)	0.000000 (0.000000)	0.000000 (0.000000)

*Particle concentrations indicated are based upon assumptions that GBFs perform as effectively as a conventional low-temperature fabric filter.

Table 10

PARTICULATE EMISSION LEVELS

Performance Characteristics	No. of GBF Stages	Total Particulates Emitted	
		g/sm^3 (g/scf)	g/MJ ($lb/10^6$ Btu)
Rexnord	1	(0.1705) 0.390	(0.137) 0.136
Rexnord	2	(0.1028) 0.235	(0.191) 0.0820
W Bench Tests	1	(0.0105) 0.0240	(0.0195) 0.00838
W Bench Tests	2	(0.00025) 0.000572	(0.000465) 0.000200
Fabric Filter	1	(0.0015) 0.00343	(0.00279) 0.00120
Fabric Filter	2	nil	nil

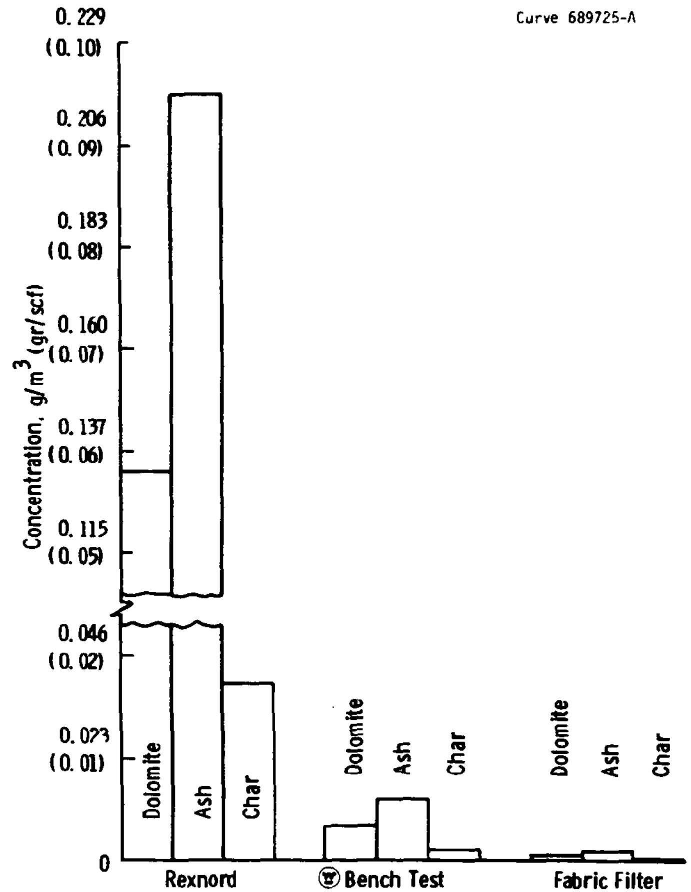

Figure 10 - Particulate Concentration at Outlet of First-Stage Granular-Bed Filter

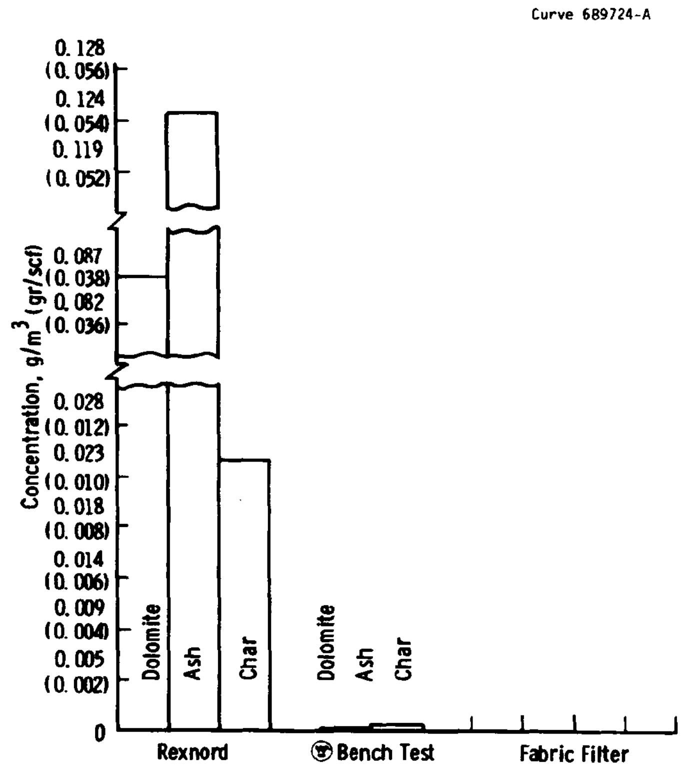

Figure 11 - Particulate Concentration at Outlet of Second-Stage Granular-Bed Filter

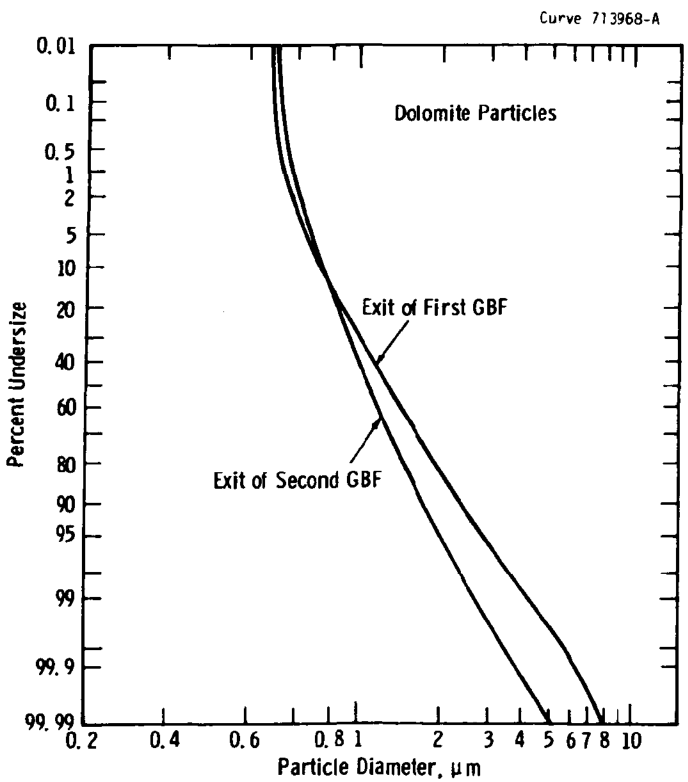

Figure 12 - Projected Outlet Size Distribution
Based on Rexnord Commercial Unit
(Dolomite Particles)

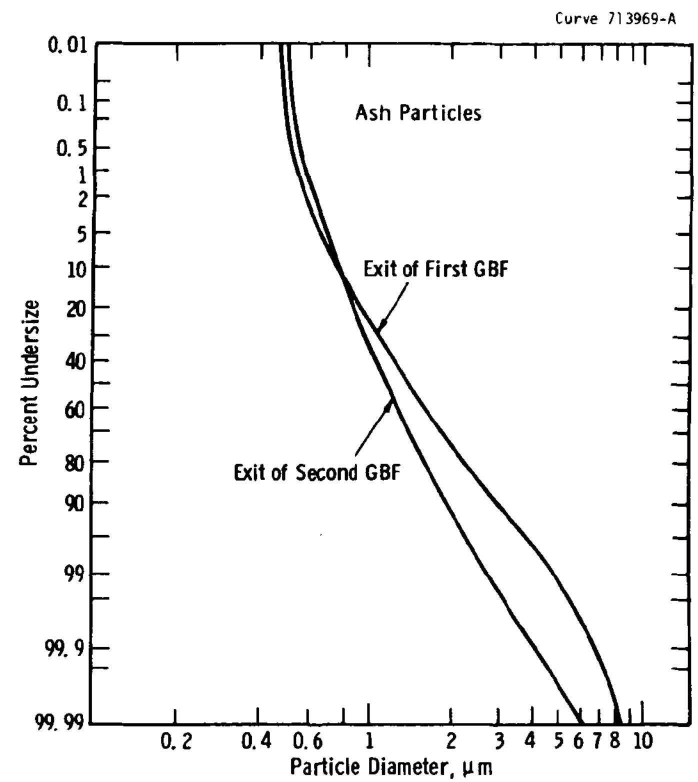

Figure 13 - Projected Outlet Based on
Rexnord Commercial Unit
(Ash Particles)

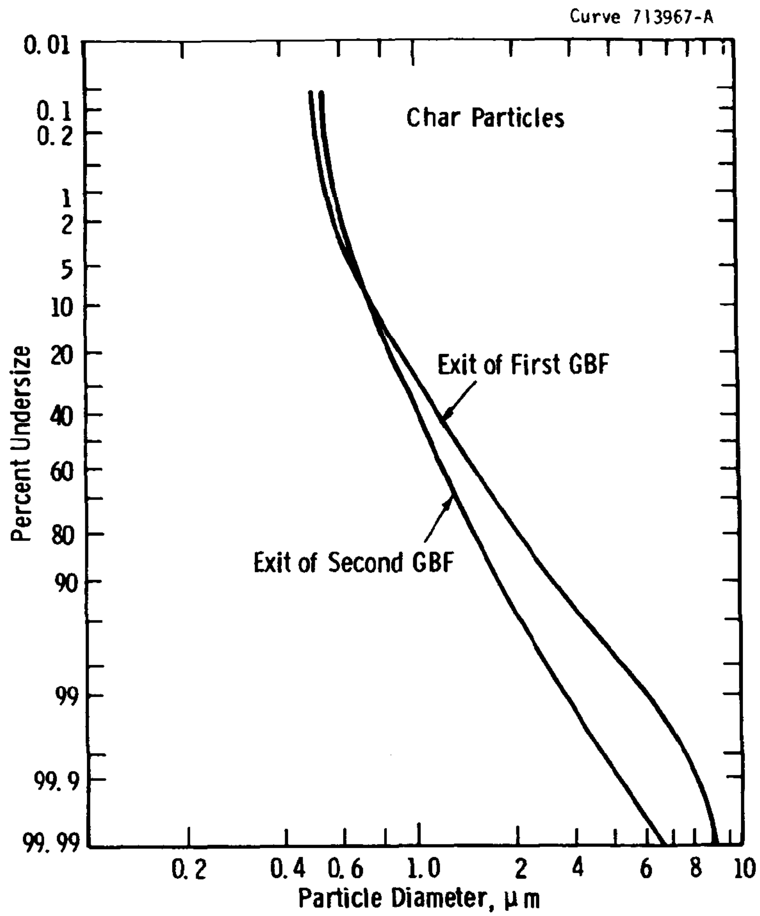

Figure 14 - Projected Outlet Based on Rexnord Commercial Unit

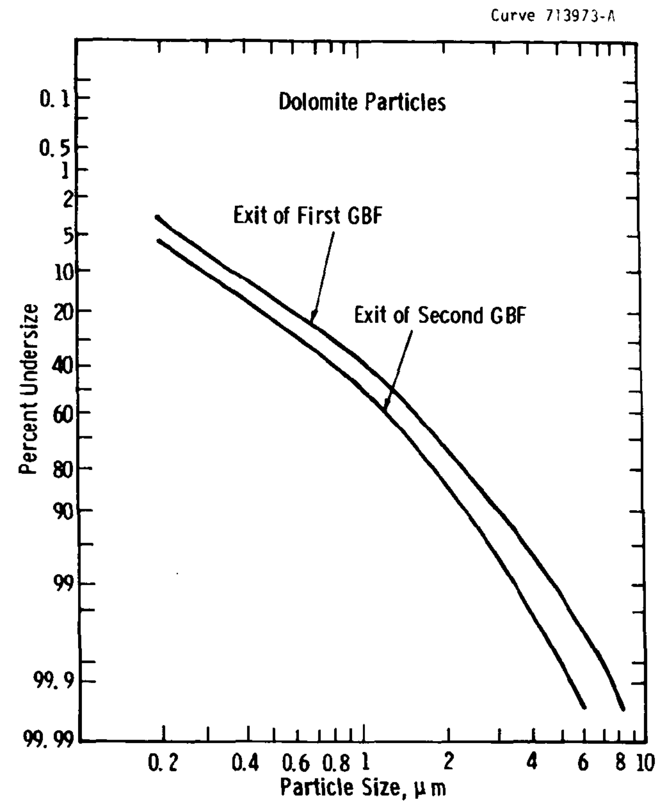

Figure 15 - Projected Granular-Bed Filter Outlet Size Distribution Based on Westinghouse Bench-Scale Experiments (Dolomite Particles)

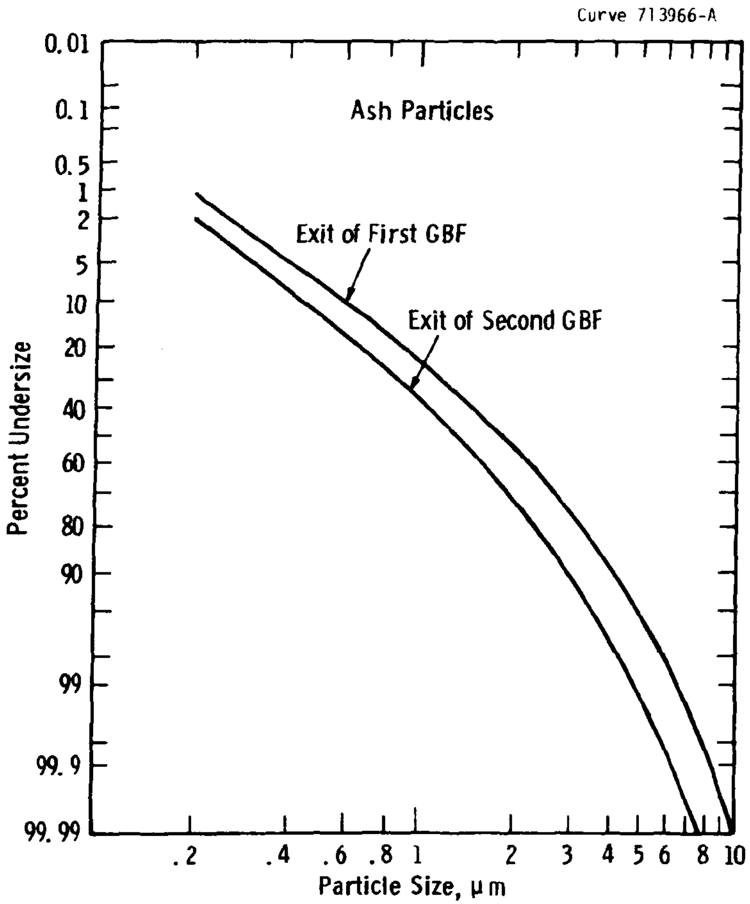

Figure 16 - Projected Granular-Bed Filter Outlet Based on Westinghouse Bench-Scale Experiments (Ash Partles)

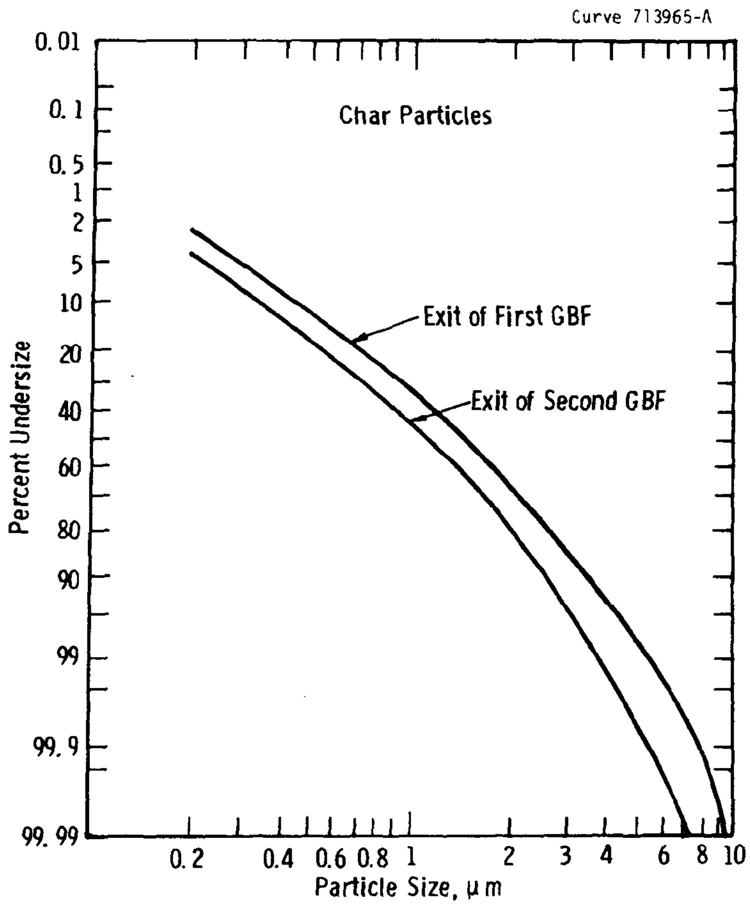

Figure 17 - Projected Granular-Bed Filter Outlet Based on Westinghouse Bench-Scale Experiments (Char Partcles)

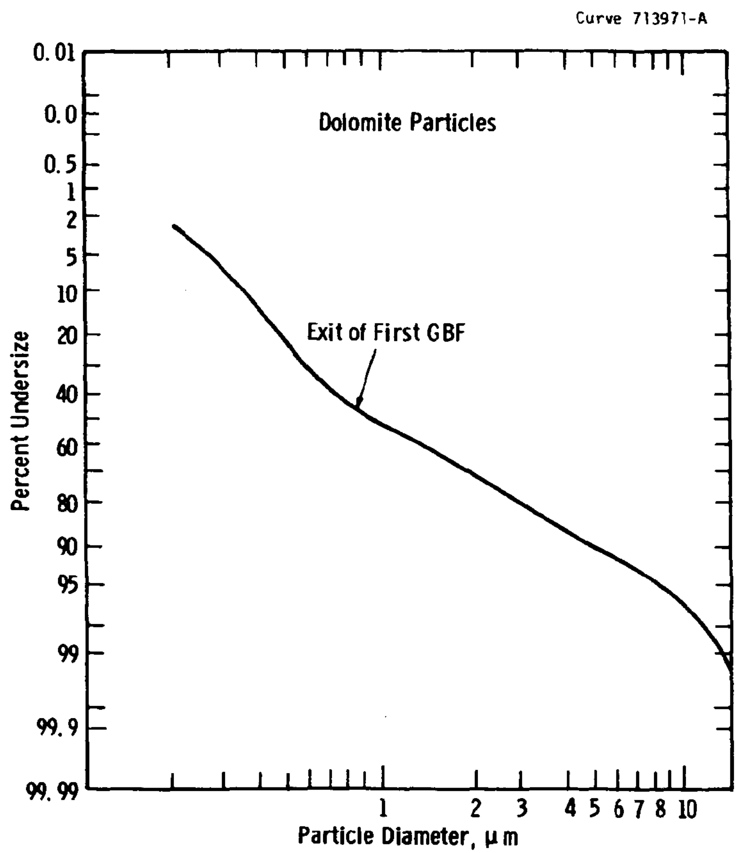

Figure 18 - Projected Granular-Bed Filter Outlet Size Distribution Based on Conventional Fabric-Filter Unit Performance (Dolomite Particles)

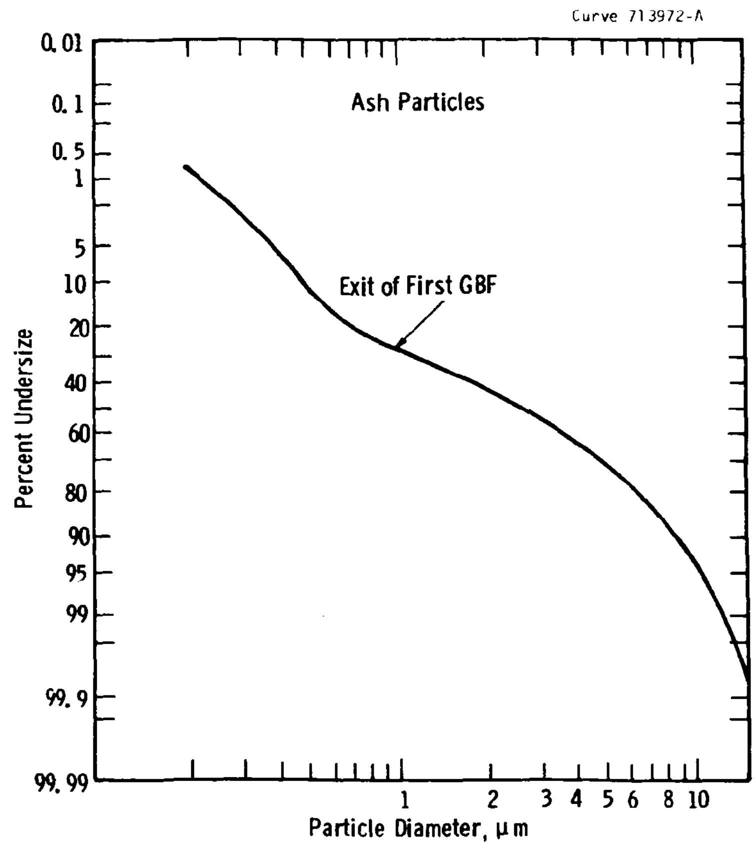

Figure 19 - Projected Granular-Bed Filter Outlet Based on Conventional Fabric Filter Unit Performance (Ash Particles)

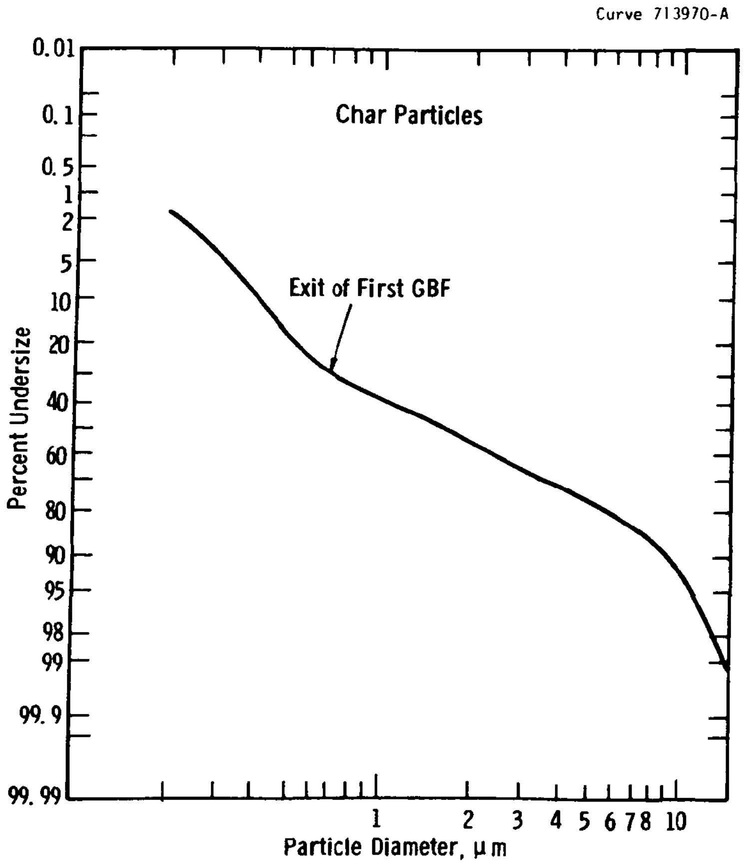

Figure 20 - Projected Granular-Bed Filter Outlet Based on Conventional Fabric-Filter Unit Performance (Char Particles)

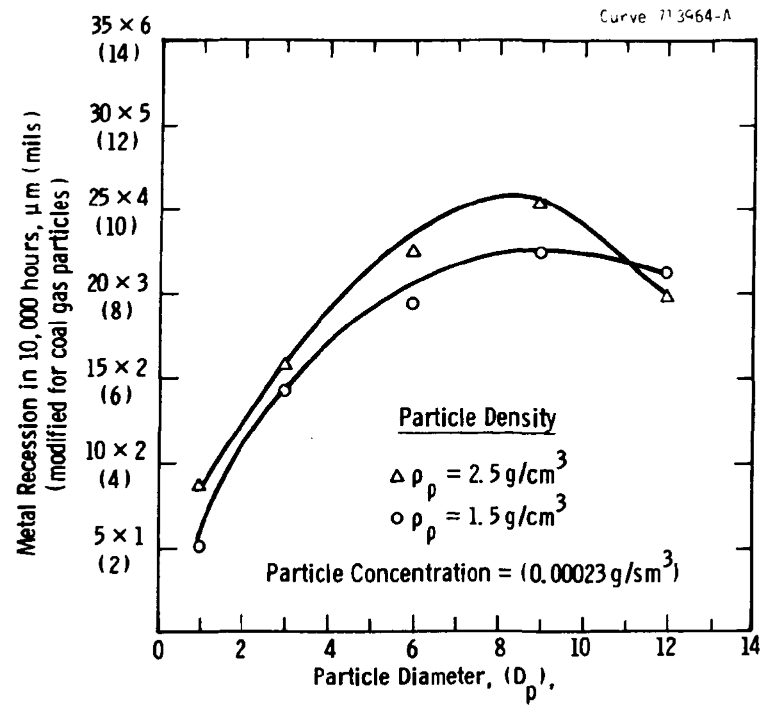

Figure 21 - Blade Leading Edge Erosion Rates[8]

1. Choose a step size (e.g., 1 μm).
2. Enter Figure 11 (or appropriate subsequent figure) and read off the percentage undersize at each side of the step.
3. Take the difference between these two values. This gives the percentage of the particulate size at the mean of the step.
4. Multiply this percentage by the total particulate concentration entering the turbine.
5. Enter Figure 21 (taken from Reference 8) and read from the curve the metal recession (for 10,000 hours of operation) at the rotor leading edge for the particular particle size of this integration step. (Note the density parameter. The erosion rate reduces slightly for lower density particles since they deviate less from the gas streamlines. Note also that the erosion of Figure 21 is for a given particle concentration.)
6. Multiply the erosion read from the curve by the ratio of the actual concentration (from Step 4) to the concentration for which Figure 21 has been calculated.
7. From Table 11, which traces the trajectory of a given size particle in a specified boundary layer thickness, read the velocity of the particle at impact with the blade surface.
8. Calculate the square of the ratio of the velocity at impact to the velocity entering the boundary layer.
9. As mentioned in Reference 8, the erosive effect of a particle is proportional to the square of the impact velocity. Therefore, multiply the recession rate calculated in Step 6 by the factor calculated in Step 8.
10. Repeat Steps 1 through 9 over the complete range of particle sizes.
11. Add the erosion rates of all the particle sizes.
12. Based on the estimate of the Westinghouse Gas Turbine Division that the maximum allowable amount of erosion of a turbine rotor blade would be 254 mm (0.100 in), multiply 10,000 hours by the

Table 11

THERMAL HISTORY OF PARTICLES ENTERING A 0.005 INCH THICK BOUNDARY LAYER WITH 10 DEG. INCIDENCE, AT R1-L.E.
GAS VEL. = 1200 FT/S, TO = 2060.DEG.R, TW = 1460.DEG.R, THER.EMISS. = 0.9

-.94115-26	.00000	.99154+00	.11980+04	.15990+04	.12000+04	-.21156+03	.15994+04	.10401+02	.11773+05	.49687+05
.11111-08	.32000+00	.94377+00	.11862+04	.15931+04	.12000+04	-.21013+03	.15994+04	.10393+02	.11741+05	.33002+04
.22222-06	.63997+00	.88793+00	.11716+04	.15359+04	.11998+04	-.20672+03	.15993+04	.10430+02	.11713+05	.15458+04
.33333-06	.95989+00	.83249+00	.11568+04	.15784+04	.11996+04	-.20732+03	.15992+04	.10523+02	.11696+05	.99565+03
.44444-06	.12797+01	.77738+00	.11411+04	.15705+04	.11992+04	-.20592+03	.15990+04	.10679+02	.11691+05	.72610+03
.55556-06	.15995+01	.72263+00	.11245+04	.15623+04	.11988+04	-.20453+03	.15988+04	.10903+02	.11699+05	.56570+03
.66667-06	.19191+01	.66827+00	.11071+04	.16635+04	.11932+04	-.20315+03	.15986+04	.11205+02	.11721+05	.45902+03

ratio of 254 mm (0.100 in) to the total recession calculated in Step 11. This gives the time in hours required to erode away 254 mm (0.100 in).

13. Repeat the above procedure for each of the particulate constituents -- i.e., dolomite, ash, and char.

RESULTS

The boundary layer near the leading edge of a rotor blade is very thin, of course, since it has had little distance in which to develop. The thickness is a function of the Reynolds number and can be estimated from simple flat plate theory. Because of the rapid acceleration of the main stream flow around the leading edge of a blade, the boundary layer growth tends to be retarded in this region compared to a flat plate. In Figure 144 of Reference 15, an experimental determination of boundary layer thickness is given for an airfoil at approximately the right Reynolds number and thickness chord ratio. The thickness shown in the figure near the leading edge is approximately 127 μm (0.005 in). Accordingly, calculations of the erosive life of the blading have been made for boundary layer thicknesses of 0.0, 127, 254, and 508 μm (0.0, 0.005, 0.010, and 0.020 in) to determine sensitivity to this critical parameter.

Another parameter of importance, which is not well defined at this point, is the erosiveness of the particulates. On the basis of evidence available from the literature, Reference 8 assumed that the erosiveness of the ash and dolomite particles was 1/25 of that of silcon carbide (SiC) particles.

Figures 22 through 24 show the results of the calculation for the three levels of performance of the granular bed filters. Blade life, in hours, is plotted against boundary layer thickness with parameters of erosiveness. The range of erosiveness chosen is from twice to half what was indicated in Reference 8.

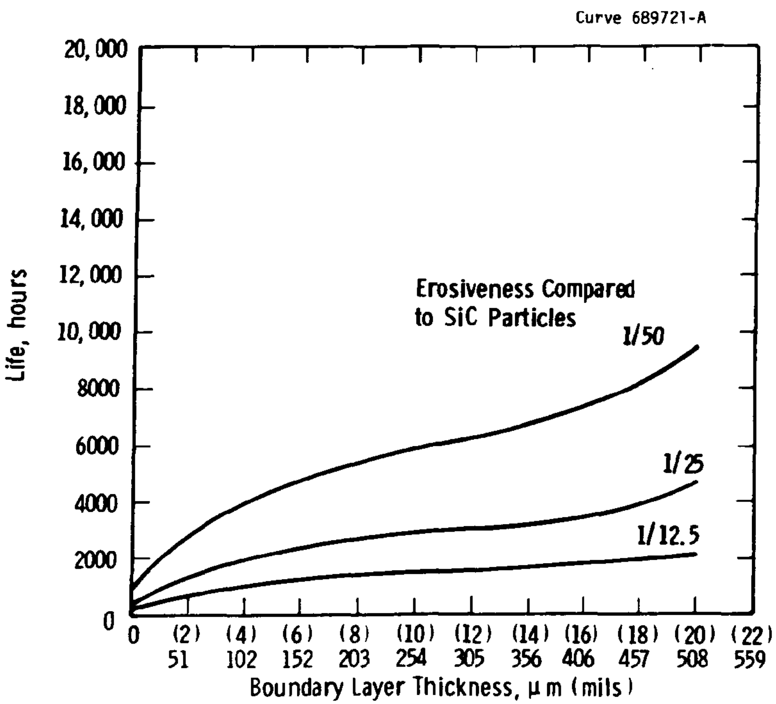

Figure 22 - Projected Turbine Life for a Particulate Removal System with Two Stages of Granular-Bed Filters (performance based on Rexnord, commercial unit)

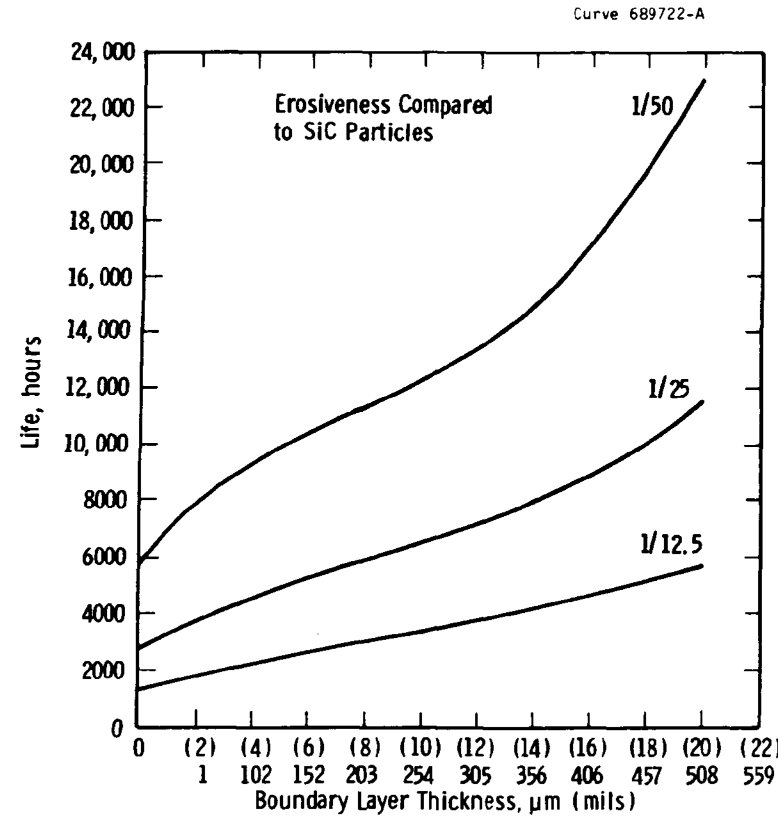

Figure 23 - Projected Turbine Life for a Particulate Removal System with One Stage of Granular-Bed Filters (performance based on Westinghouse bench-scale experiments)

Figure 22 shows the life of the turbine with two stages of granular bed filters, each with a performance equal to that of the Rexnord commercial unit. As the figure indicates, for a boundary layer thickness in the expected range of 0 - 127 μm (0 to 5 mils), the life of the turbine is very short -- no more than six months, even assuming an optimistic level of erosivity. One stage of Rexnord filter was found to give inadequate life.

Figure 23 shows the life of the turbine with one stage of granular bed filters with a performance equal to that of the Westinghouse bench-test unit. As the figure indicates, this level of performance increases the turbine life appreciably, although it is still much shorter than is usual for utility equipment.

Not shown on a figure is the life resulting from the use of two stages of granular-bed filters with the Westinghouse measured

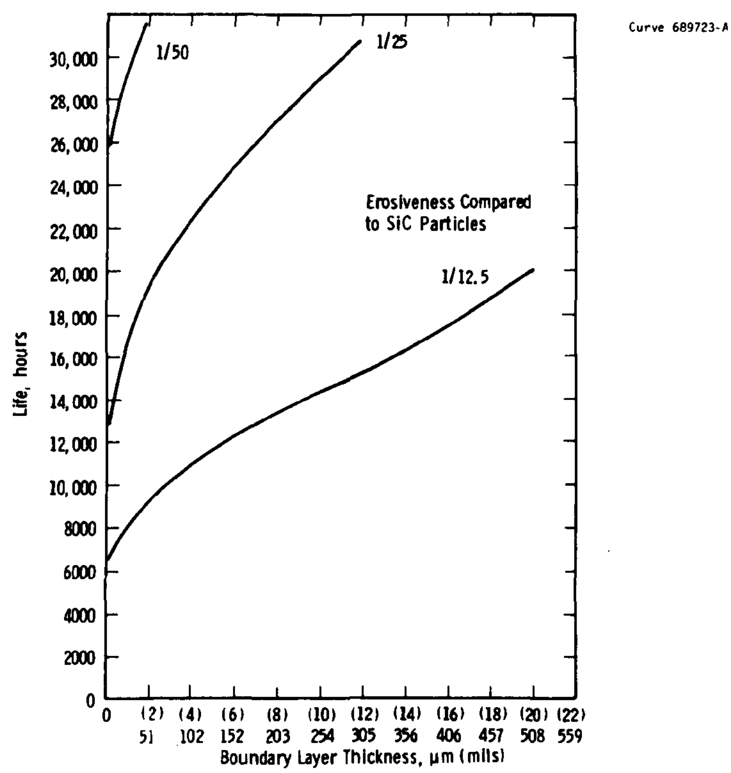

Figure 24 - Projected Turbine Life for a Particulate Removal System with One Stage of Granular-Bed Filters (performance based on conventional fabric filter efficiency)

performance. This is because the particulate concentration is so low that even if a zero boundary layer thickness is assumed, the turbine life is 19 years. In other words, the erosion rate is negligible.

Figure 24 shows the life of the turbine with one stage of granular-bed filters with a performance equal to that of a conventional low-temperature fabric filter. The turbine life here is appreciably increased over that of the previous two cases. The curves are relatively steep, especially so near the zero boundary layer thickness, but with the erosiveness that Reference 8 assumed, the turbine life is approximately two years. With two stages of granular-bed filters, however, the erosion rate is essentially zero since the particulate concentration is zero when calculated to six decimal places.

If we assume that replacing the turbine blading would be considered an operating and maintenance (O&M) expense, we can calculate it in the form of a cost of electricity (COE) in units of mills/kWh. Based on information from the Westinghouse Gas Turbine Division, the installed cost of a complete change of turbine blading is approximately \$2 million per turbine. (No charge is included for plant down time. We have assumed for purposes of this study that blade changes would be made during normal maintenance periods and not as a result of forced outages.) As shown in the ECAS report,[12] the power output of a pressurized fluidized-bed power plant with two W501 gas turbines is 679,000 kW. Thus, the equation for the cost of electricity in mills/kWh due to a blading change in terms of the time in hours between blading changes is as follows:

$$\mathrm{COE} = \frac{\$2 \times 10^6/(\text{turbine-change})\ (2\ \text{turbines})\ (1000\ \text{mills}/\$)}{\mathrm{CF}\ \frac{\text{kW output}}{\text{kW rating}}\ (679{,}000\ \text{kW rating})\ (A\ \text{hr/change})} \; .$$

Note that as the capacity factor (CF) goes down, the kW output goes down so the COE goes up. On the other hand, as the CF goes down, the coal input and resultant particulate loading goes down so the life, A, goes up. Since the life was calculated on the basis of full load, the sensible approach is to take CF = 1.0. Thus, the equation becomes:

$$\mathrm{COE} = \frac{5891}{A} \quad .$$

With this equation the lifetimes shown in Figures 22 through 24 can be converted to COE. This result is shown in Figures 25 and 26 for two of the three assumed granular bed filter performances. The calculation was not made for the Rexnord performance since the turbine lives would be too short to have practical application in a utility power plant. The COE is plotted versus the boundary layer thickness, with parameters of erosiveness for one stage of granular-bed filters. The Δ COE associated with a two-stage, granular-bed system has been calculated and is also shown in Figures 25 and 26.

Calculations have been made to estimate the cost of granular-bed filters for the gas flow conditions of this power plant. Each module handles a volume flow rate of 15.72 m^3/s (33,300 acfm), which results in a filter pressure vessel diameter of 7.6 m (25 ft), if we assume a design face velocity of 15.2 m (50 ft/min), a face area per filter element of 0.37 m^2 (4 ft^2), and a plan area per element of 0.26 m^2 (2.8 ft^2) (which includes the open flow area around each element). The cost of a 7.6 m (25 ft) diameter, granular-bed filter module for operation at 982°C (1800°F) and 1013 kPa (10 atm) has been estimated on the following basis:

Cost base	- mid-1975 $
Field labor	- 51% of direct installation costs
Professional services	- 10%
Escalation	- constant dollars
Interest during construction	- 10%
Construction time	- 5 years

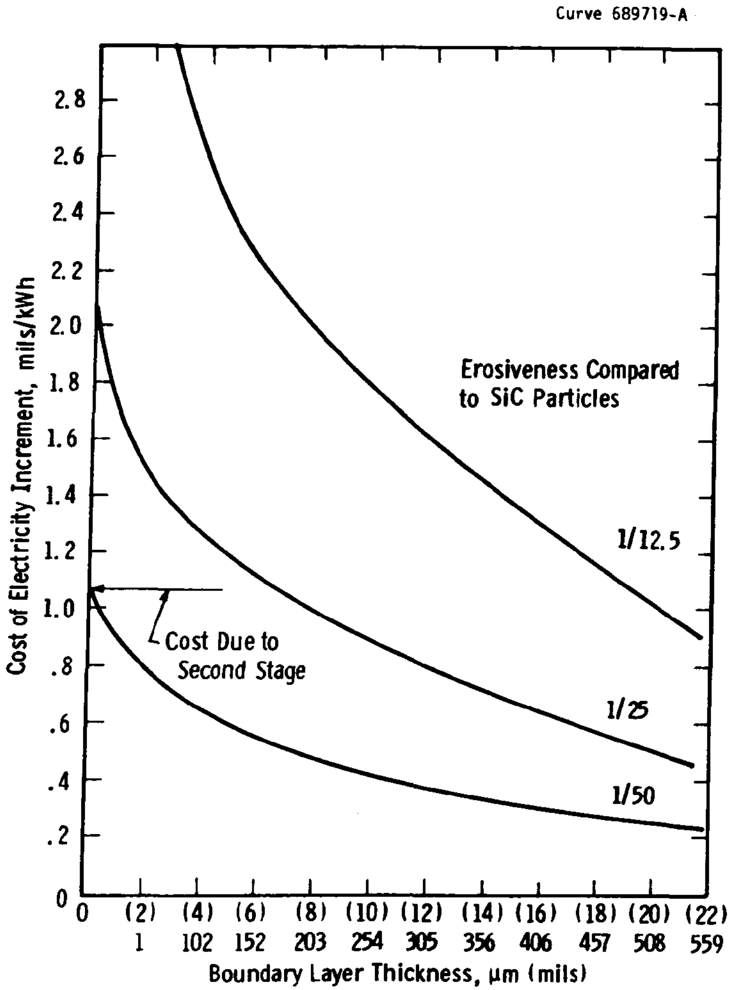

Figure 25 - Cost of Electricity Increments due to Turbine Blade Replacement Using Granular-Bed Filters (performance based on granular-bed efficiency)

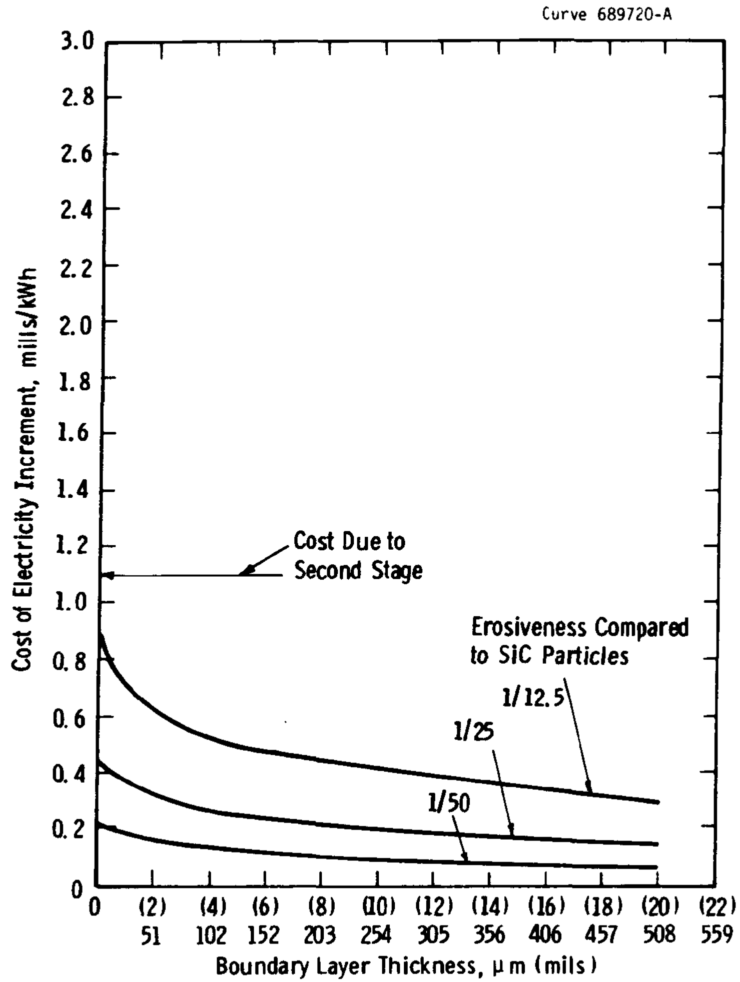

Figure 26 - Cost of Electricity Increments due to Turbine Blade Replacement Using Granular-Bed Filters (performance based on fabric-filter efficiency)

The resulting cost for a 7.6 m (25 ft) single-stage, granular-bed filter module is \$1,850,000.

The equation for the cost of electricity due to capital expense is

$$\text{COE} = \frac{(\text{yearly cost of capital, \$/\$-yr})\ (\text{capital cost, \$})\ (1000\ \text{mills/\$})}{(\text{capacity factor})\ (8760\ \text{hrs/yr})\ (\text{plant power, kW})}.$$

The incremental COE due to the incremental cost of two stages compared to one stage is calculated to be 1.08 mills/kWh.

CONCLUSIONS

1. This analysis provides a basis for evaluating the economic advantage of improving filter performance rather than replacing turbine blades.
2. More accurate experimental information on particulate erosiveness and staged-bed filter performance is required before a definitive comparison can be made.
3. Although the cost of staged filters is substantial, the difference between this cost and the cost of blade replacement in the case of least frequent blade replacement (Figure 25) is only about one half mill/kWh, which indicates that rather large capital costs can be tolerated for efficient filter systems.

RECOMMENDATIONS

1. Accelerate development of granular-bed filters and alternative filter concepts designed for gas turbine applications.
2. Continue experimental studies to investigate particulate erosion rates of gas turbine blading.

8. INDIRECT AIR-COOLED PRESSURIZED FLUIDIZED-BED COMBUSTION CONCEPT SYSTEMS EVALUATION

INTRODUCTION

A gas turbine cycle combustor using PFBC with indirect heating of pressurized air in immersed tubes is being investigated by Curtiss Wright under DOE funding.[16] We compared the performance and cost of energy of a combined-cycle plant using this configuration and those of a combined-cycle plant using an adiabatic fluidized-bed combustor. We considered two configurations of the partially indirectly heated system: one with a CBC and one without. In each case we selected the amount of excess air that would give an overall carbon loss equivalent to 1 percent of the energy in the coal.

Since environmental concerns are of primary importance to EPA, this study included an assessment of the effect of the partially indirectly heated concept on pollutant emissions (particulates, SO_x, NO_x, products of incomplete combustion, and solid wastes) compared to the adiabatic PFBC System.

This technoeconomic study was carried out in late 1976 and early 1977 and reflects the FBC technology and the economic situation that obtained in that time period. No attempt has been made to update the results.

BACKGROUND

The products of combustion from PFBC of coal are passed through a gas turbine expander. The cost of the particulate removal equipment required to clean the combustion products well enough to avoid problems with erosion of, corrosion of, and deposition on the expander parts is expected to be significant. This is particularly true of gas turbines with an adiabatic fluidized-bed combustor, where air equivalence ratios of approximately 3 prevail (200% excess air), since the cost of the particulate removal equipment is roughly proportionate to the volume of the combustion products and the ratio of combustion products to coal is high.

In the partially indirectly heated gas turbine combustor concept the volume of the combustion products that must be cleaned is reduced substantially by using the minimum amount of combustion air and heating the balance of the gas turbine airflow indirectly with tubes submerged in the fluidized bed. This indirectly heated air is then mixed with the combustion products after they have been cleaned, and this mixed stream constitutes the flow to the gas turbine expander. There is a trade-off, therefore, between the cost of the particulate removal equipment and the cost of the heat transfer surface required for indirectly heating the air that bypasses the combustor.

Description of Systems Evaluated

The coal-fired power systems that were evaluated and compared in this study are:

- Base Case - A combined-cycle system with an adiabatic fluidized-bed combustor and *in situ* desulfurization
- Alternative Case I - Partially indirect heating with a CBC
- Alternative Case II - Partially indirect heating without a CBC.

Performance calculations were made for each of these configurations with Ohio Pittsburgh No. 8 seam coal with 3 percent moisture.

The gas turbine design conditions were as follows:

Ambient air conditions - International Standards Organization (ISO)
Compressor airflow - 345 kg/s

Compressor pressure ratio - 10

Compressor isentropic efficiency - 0.853 (polytropic efficiency $\approx$ 0.89)

Expander isentropic efficiency - 0.927 (polytropic efficiency $\approx$ 0.90)

Temperature drop due to heat transfer between combustion products and combustion air - 8°C.

The fluidized-bed combustor design conditions common to all cases were as follows:

Primary bed temperature - 1010°C (value used in ECAS[17])

Coal size - <6.4 mm

Dolomite size - <4.8 mm

Ca/S atom ratio - 1.5

Superficial velocity - $\leqq$1.5 m/s

Maximum bed depth - 4.6 m

Pressure loss (including particulate removal equipment) - 7.5%.

The design conditions for the heat recovery steam generator (HRSG) were as follows:

Type - unfired

Pinch - 22.2°C

Offset - 2.8°C.

The Base Case was previously treated in Reference 18 in a somewhat different configuration. The configuration used in this study is shown in Figure 27. A Ducon cyclone separator was used for the first stage of particulate removal, a Ducon granular-bed filter for the second stage. Grade efficiency plots for these components are given in Appendix A. The fluidized-bed design conditions specific to the Base Case were as follows:

Excess air - 237 percent

Superficial velocity - 1.5 m/s

Bed depth - 2.0 m

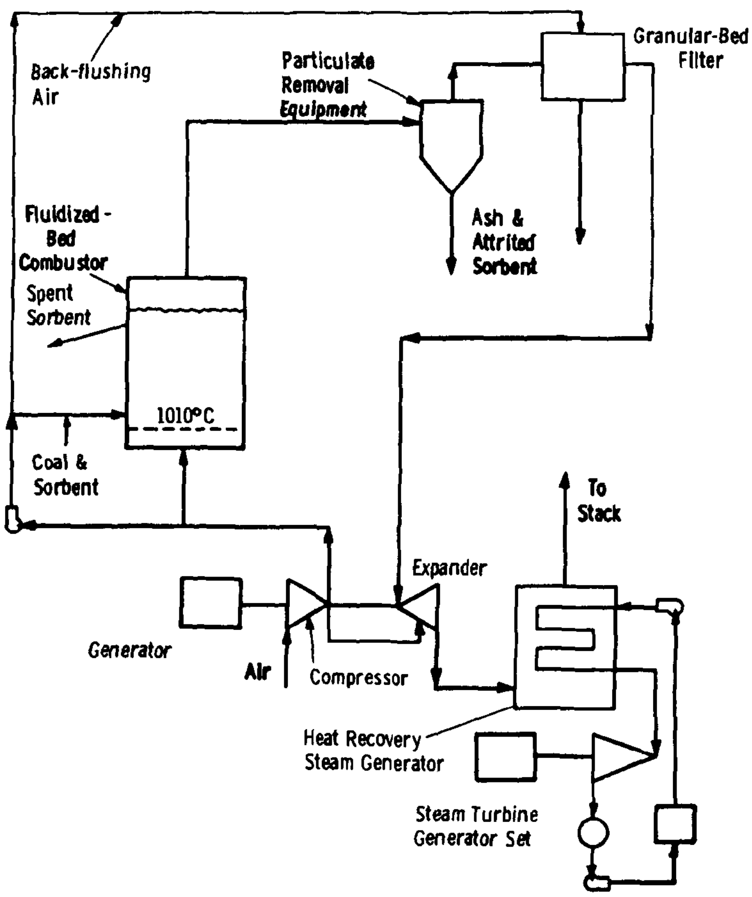

Figure 27 - Combined-Cycle System Utilizing Fluidized-Bed Combustion (Base Case)

Combustion losses

Incomplete combustion	-	0.4
Losses to atmosphere	-	0.8
Sensible heat in solids	-	1.25
Desulfurization reactions	-	0.1
Total		2.55 percent
Overall combustion efficiency		97.45 percent.

With allowances for thermal losses from the ductwork, the temperature of the gas at the gas turbine expander inlet was 996°C, and the gas turbine expander cooling airflow was 8.3 percent.

The Alternative Case I configuration is shown in Figure 28. A Ducon cyclone separator was used for the primary bed effluent, a Tan Jet cyclone separator for the CBC effluent, and a Ducon granular-bed filter for the final separator. Grade efficiency plots for these separators are given in Appendix A. The fluidized-bed design conditions specific to Alternative Case I are as follows:

Primary bed

- Excess air - 0 percent
- Bed depth - 4.6 m
- Superficial velocity - 1 m/s*
- Carbon losses - 10 percent of equivalent energy in coal

CBC bed

- Excess air - 0 percent
- Bed depth - 4.6 m
- Superficial velocity - 0.85 m/s
- Carbon losses - 10 percent of input
- Temperature - 1010°C

Combustion losses

Incomplete combustion	-	1.0 percent
Losses to atmosphere	-	0.8
Sensible heat in solids	-	1.25
Desulfurization reactions	-	0.10
Total		3.15

Overall combustion efficiency - 96.85 percent.

The effectiveness of the submerged heat exchangers in the primary and CBC beds for indirectly heating part of the gas turbine working fluid were assumed to be 85 percent. This assumption gives an air outlet temperature of 907°C from both the primary and CBC bed heat exchangers. After the combustion products and the indirectly heated air

*Superficial velocity reduced below 1.5 m/s nominal design value to satisfy limits on maximum bed depth.

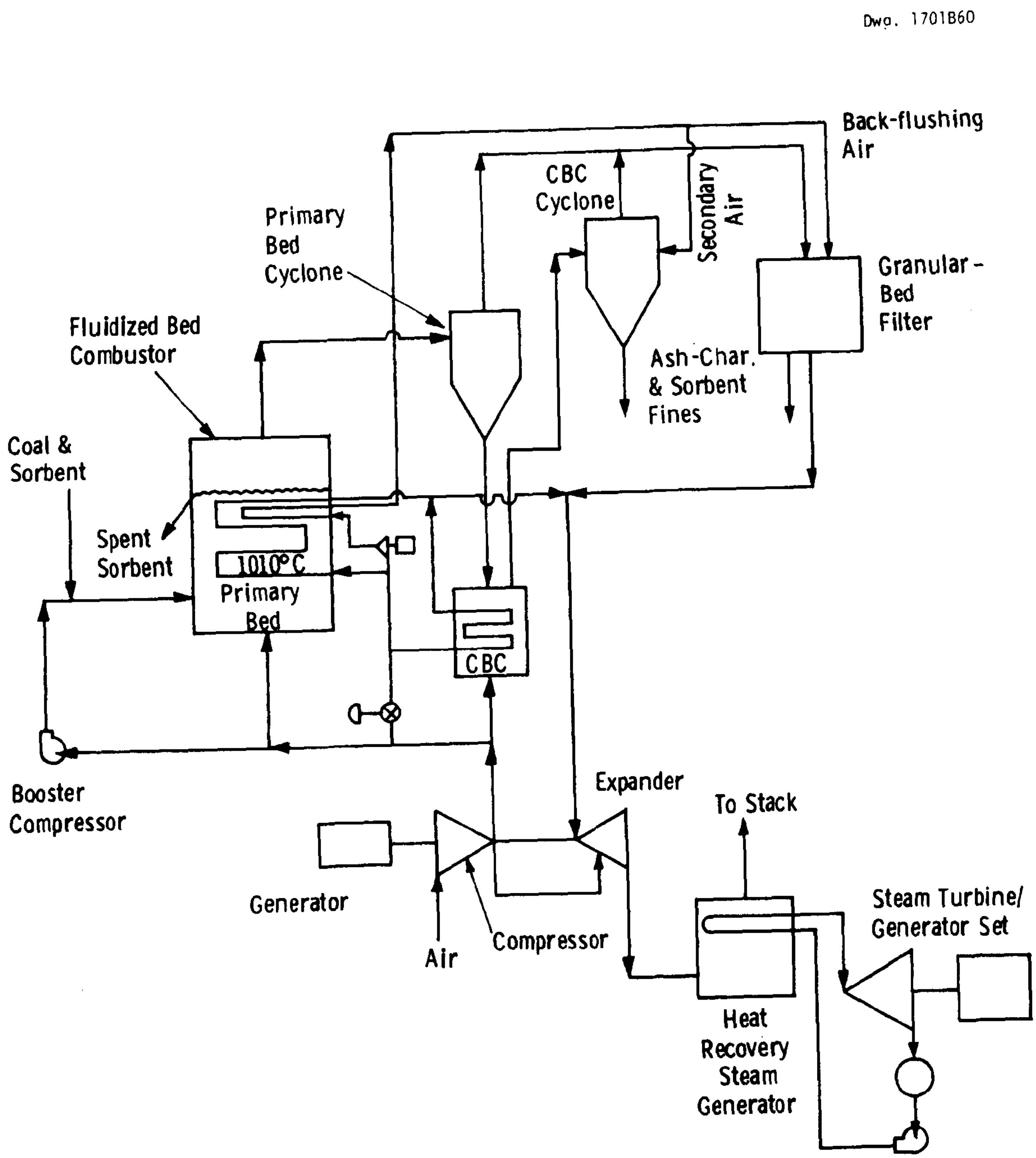

Figure 28 - Combined-Cycle System Utilizing Fluidized-Bed Combustion with Indirect Heating of Part of the Working Fluid (Alternative Case I - CBC)

have been mixed, and radiation losses and heat transferred between the hot products and the combustion air have been allowed for, the gas turbine expander inlet temperature is only 928°C.

The Alternative Case II configuration is shown in Figure 29. This configuration is obviously considerably less complicated than that with the CBC shown in Figure 28. A Ducon cyclone was used for the first-stage separator, a Ducon granular-bed filter for the second stage. Grade efficiency plots for these separators are given in Appendix A. The fluidized-bed design conditions specific to Alternative Case II were as follows:

Excess air - 60 percent

Superficial velocity - 1.5 m/s

Bed depth - 3.3 m

Combustion losses

Incomplete combustion	-	1.00 percent
Losses to atmosphere	-	0.80
Sensible heat in solids	-	1.25
Desulfurization reactions	-	0.10
Total		3.15

Overall combustion efficiency - 96.85 percent.

The effectiveness of the heat transfer surface submerged in the bed was again assumed to be 85 percent, giving an air temperature out of the heat exchanger of 907°C. After the combustion products have been mixed with the indirectly heated air, and allowances for losses to the atmosphere and heat transferred from the hot combustion products to the combustion air have been made, the temperature at the gas turbine expander inlet is 942°C.

Alternative Case I is probably not a practical configuration because of the likelihood of severe corrosion of the immersed air heater tubes in a bed with zero excess air. It does, however, represent one boundary of the design spectrum for partially indirectly heated systems (the other boundary being the Base Case).

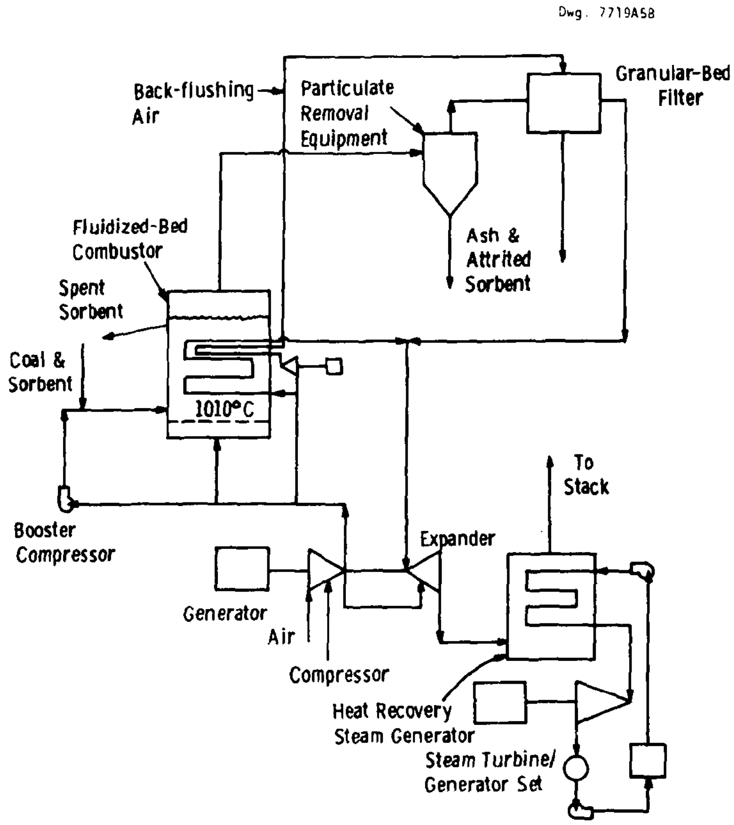

Figure 29 - Combined-Cycle System Utilizing Fluidized-Bed Combustion with Indirect Heating of Part of the Working Fluid (Alternative Case II - no CBC)

RESULTS OF STUDY

A summary of the plant performance for the cases studied is given in Table 12.

All three of these configurations have heat rates that are appreciably better than that for a conventional steam plant with FGD (i.e., 10,475 kJ/kWh[19]) with the admittedly optimistic bed temperature of 1010°C. If limitations on bed temperature per se and/or hot gas duct temperature require the bed temperature to be reduced significantly, the heat rates for all configurations will be increased uniformly. If limitations on the maximum metal temperature of the in-bed air tubes require the temperature of the indirectly heated air to be reduced significantly, the heat rates of the partially indirectly heated configurations will increase relative to that for the Base (adiabatic) Case.

Table 12

SUMMARY OF PLANT PERFORMANCES

Case	Gas Turbine Output, MW/G.T. Module	Steam Turbine Output, MW/G.T. Module	Total Electrical Output, MW/G.T. Module	Coal Feed Rate, ton/hr/ G.T. Module	Heat Rate, kJ/kWh
Base (Adiabatic)	73.8	34.1	107.9	36.2	9148
Alternative I	66.3	27.4	93.7	32.7	9587
Alternative II	67.8	28.7	96.5	33.4	9504

Plot plans of a single gas turbine module for the Base Case, Alternative Case I, and Alternative Case II are shown in Figure 30 through 32. Table 13 summarizes the plant design configurations for plants with a nominal capacity of 400 MW.

Estimates of the capital cost of nominally 400 MW plants for each of the three configurations were made on the basis of the following assumptions:

Cost base - 1st quarter of 1976

Construction time - 4 years

Indirect construction costs - 13.5 percent of the total direct costs*

Professional services - 10% of the direct plus indirect costs

Contingency - 10 percent

Escalation rate - 6-1/2 percent

Interest during construction - 10 percent

Expenditure rate - S curve supplied by NASA for use in the ECAS study[17]

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*Equivalent to ~50 percent of direct installation costs.

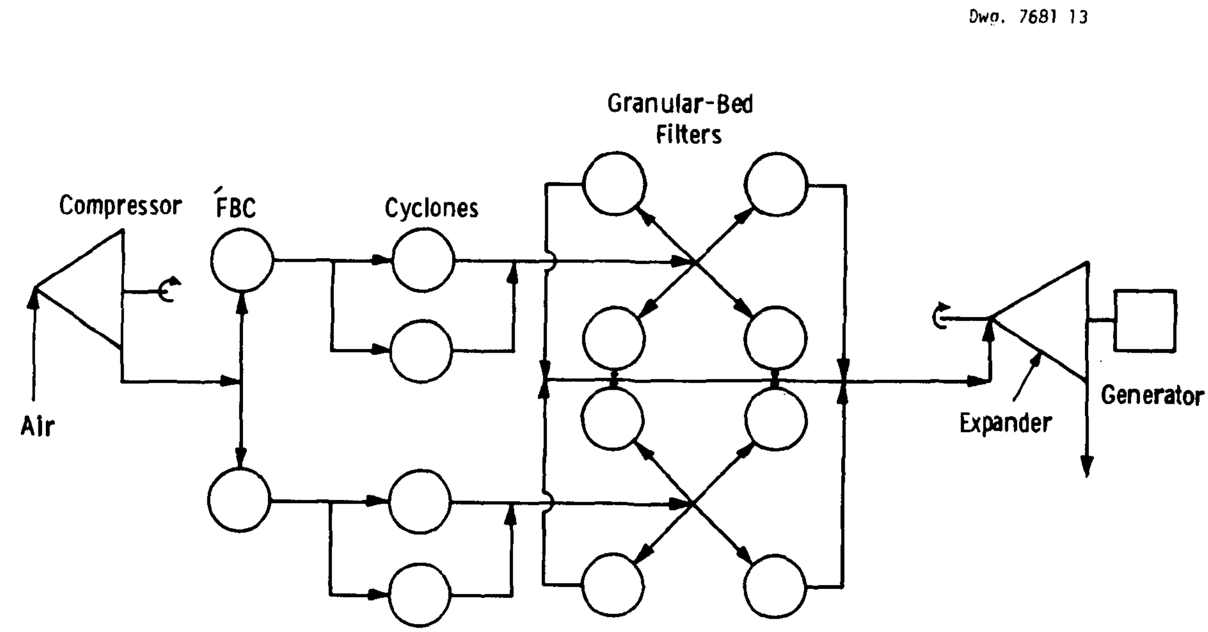

Figure 30 - Plot Plan of Single Gas Turbine Module for Base Case

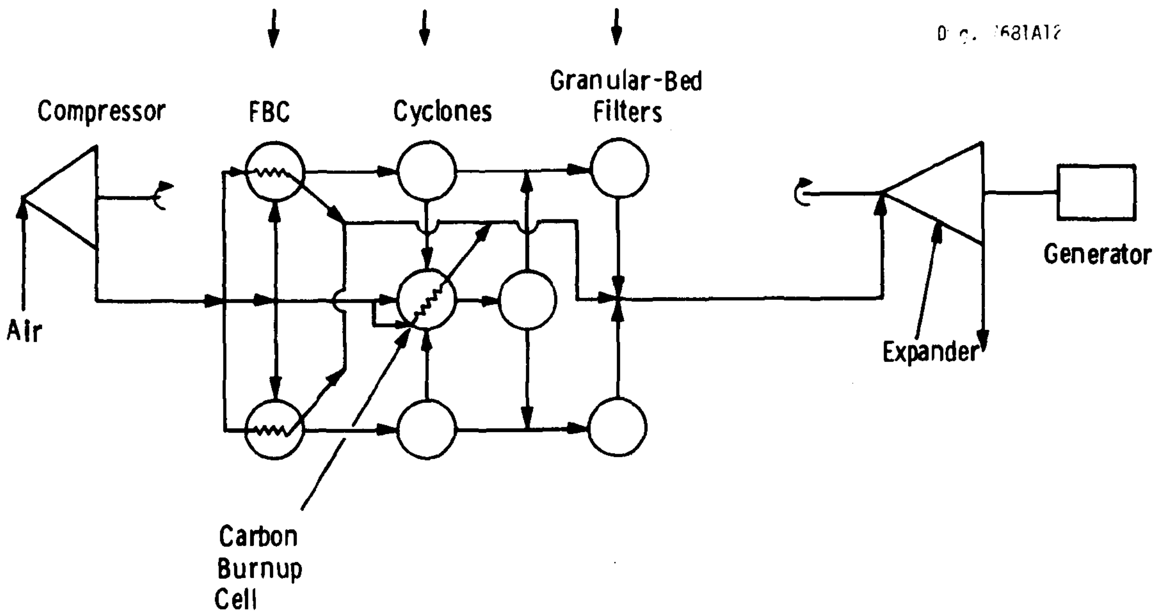

Figure 31 - Plot Plan of Single Gas Turbine Module for Alternative Case I

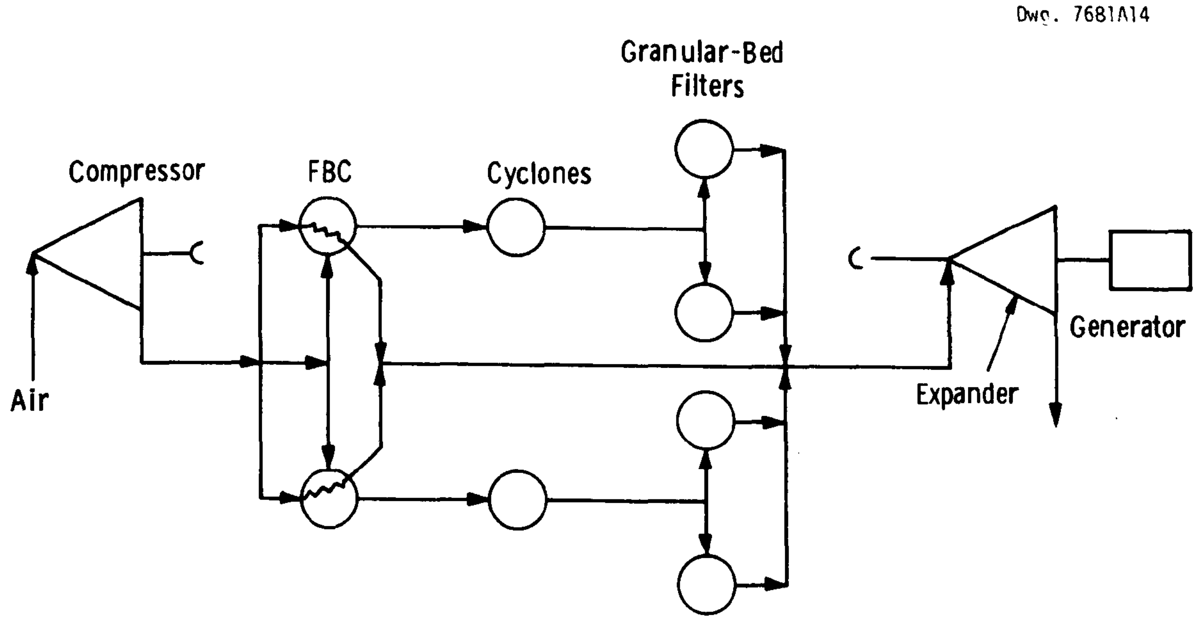

Figure 32 - Plot Plan of Single Gas Turbine Module for Alternative Case II

Cost estimates for equipment manufactured by Westinghouse, such as the gas turbines, heat recovery steam generators, and steam turbines, were obtained from cost correlations supplied by the pertinent Westinghouse divisions during ECAS.[17] Cost estimates for high-temperature particulate removal equipment were based on information obtained from equipment suppliers during ECAS. Cost estimates for FBC equipment, solids feeding equipment, and in-bed heat transfer surface were made using procedures that originated in the Evaluation of the Fluidized Bed Combustion Processes[20] and were used in ECAS.

Summaries of the capital cost estimates for the Base Case, Alternative Case I, and Alternative Case II are given in Tables 14 through 16. The only costs that vary significantly with system configuration are those for the FBC modules and the gas-cleaning equipment. While the variations in the cost of these two components are opposite (e.g., the

Table 13

SUMMARY OF PLANT DESIGN CONFIGURATIONS

Configuration	Case		
	Base	Alternative I	Alternative II
Capacity, MW	431.6	374.8	386.0
No. of Gas Turbines	4	4	4
Combustion Modules			
Number	8	8	8
Diameter, m	3.8	3.65	3.65
Height, m	25.6	39	41
Beds/module	4	4	5
Bed depth, m	1.98	4.57	3.35
CBC Modules			
Number	--	4	--
Diameter, m	--	3.65	--
Height, m	--	10.97	--
Beds/module	--	1	--
Bed depth, m	--	4.57	--
1st-Stage Separator Modules			
Volumetric flow, am^3/s	29.3	16.1	25.3
Number	16	8	8
Diameter, m	0.76	1.0	1.3
Height, m	3.0	4.0	5.25
CBC Separator Modules			
Volumetric flow, am^3/s	--	3.21	--
Number	--	4	--
Diameter, m	--	2.6	--
Height, m	--	3.05	--
2nd-Stage Separator Modules			
Volumetric flow, am^3/s	14.6	17.9	12.7
Number	32	8	16
Diameter, m	7.77	8.23	7.01
Height, m	8.84	9.3	8.23
No. of HRSG Modules	4	4	4
No. of Steam T-G Modules	1	1	1

Table 14

CAPITAL COST ESTIMATE FOR BASE CASE
(4 gas turbines)

Item	10^6 $
1.00 Land and Land Rights	4.600
2.00 Structures & Improvements (on-site waste disposal)	13.845
3.00 Heat Rejection System	3.605
4.00 Material Handling and Storage	16.445
5.00 Energy Conversion	
PFBC	2.381
Combustion air piping	0.827
Transport air subsystem	1.546
Gas cleaning	41.341
Refractory-lined pipe	1.608
Refractory- and metal-lined pipe	2.368
Gas turbine/generator	36.542
Steam turbine/generator	7.606
HRSG	6.366
Subtotal	100.585
6.00 Auxiliary Mechanical Equipment	5.037
7.00 Auxiliary Electrical Equipment	9.614
8.00 Station Transmission Equipment	2.403
Total direct costs	156.134
Indirect construction costs (13.5% of total direct costs)	21.078
Subtotal	177.212
Professional services (10% of direct and indirect costs)	17.721
Subtotal	194.933
Contingency (10% of above)	19.493
Subtotal	214.426
Escalation during construction (6 1/2-4) (15.8% of above)	33.879
Interest during construction (10-4) (21.4% of above)	45.887
Total capitalization	294.193

Table 15

CAPITAL COST ESTIMATE FOR ALTERNATIVE CASE I

Item		10^6 \$
1.00 Land and Land Rights		4.144
2.00 Structures and Improvements (on-site waste disposal)		12.473
3.00 Heat Rejection System		3.108
4.00 Material Handling and Storage		14.815
5.00 Energy Conversion		
PFBC		12.614
Combustion air piping		0.836
Transport air subsystem		1.452
Gas cleaning		13.408
Refractory-lined pipe		0.926
Refractory- and metal-lined pipe		3.032
Gas turbine/generator		36.540
Steam turbine/generator		6.400
HRSG		5.357
	Subtotal	80.565
6.00 Auxiliary Mechanical Equipment		4.342
7.00 Auxiliary Electrical Equipment		8.288
8.00 Station/Transmission Equipment		2.072
Total direct costs		129.807
10.0 Indirect construction costs (13.5% of total direct costs)		17.524
	Subtotal	147.331
11.0 Professional services (10% of direct and indirect costs)		14.733
	Subtotal	162.064
12.0 Contingency (10% of above)		16.206
	Subtotal	178.370
13.0 Escalation during construction (6 1/2-4) (15.8%)		28.183
14.0 Interest during construction (10-4) (21.4%)		38.171
Total capitalization		244.724

Table 16

CAPITAL COST ESTIMATE FOR ALTERNATIVE CASE I

Item	10^6 \$
1.00 Land and Land Rights	4.455
2.00 Structures and Improvements (on-site waste disposal)	13.408
3.00 Heat Rejection System	3.372
4.00 Material Handling and Storage	15.926
5.00 Energy Conversion	
PFBC	10.094
Combustion air piping	0.836
Transport air subsystem	1.517
Gas cleaning	18.583
Refractory-lined pipe	0.688
Refractory- and metal-lined pipe	3.200
Gas turbine/generator	36.540
Steam turbine/generator	6.400
HRSG	5.608
Subtotal	83.466
6.00 Auxiliary Mechanical Equipment	4.711
7.00 Auxiliary Electrical Equipment	8.992
8.00 Station/Transmission Equipment	2.248
Total direct costs	136.578
10.0 Indirect construction costs (13.5% of total direct costs)	18.438
Subtotal	155.016
11.0 Professional services (10% of direct and indirect costs)	15.501
Subtotal	170.517
12.0 Contingency (10% of above)	17.052
Subtotal	187.569
13.0 Escalation during construction (6 1/2-4) (15.8%)	29.636
14.0 Interest during construction (10-4) (21.4%)	40.140
Total capitalization	257.345

adiabatic system has the highest gas cleaning cost and the lowest combustion cost), they are not equal, and the specific costs of the the three configurations vary significantly, as shown in Table 17. The adiabatic system is estimated to have the highest specific capital cost and Alternative I the lowest.

Table 17

SPECIFIC COST COMPARISON

Case	Specific Cost-$/kW
Base	682
Alternative I	653
Alternative II	667

The cost of electricity was calculated for the three cases studied on the basis of the following assumptions:

Life of plant - 30 years
Annual charge - 18 percent
Capacity factor - 65 percent
Fuel cost (1st quarter of 1976) - $0.95/GJ
Fuel escalation rate - 5%/year
O&M (including dolomite) - 2.5 mills/kWh

Table 18 summarizes the energy costs for the three cases.

Table 18

COST OF ELECTRICITY SUMMARY, mills/kWh

Item	Base	Alternative I	Alternative II
Capital	21.6	20.6	21.0
Fuel	21.5	22.6	22.3
O&M	2.5	2.5	2.5
Total	45.6	45.7	45.7

This table shows that the variations in specific capital costs of the three systems are nearly balanced by the variations in fuel consumption and that the spread in cost of electricity among the three cases is about 0.2 percent. Since the uncertainties in estimating the cost of the cost-variable components (the fluidized-bed combustor modules and the gas cleaning equipment) are considered large as compared to the spread in cost of energy, no significance can be attributed to this spread in cost of energy.

The in-bed heat transfer surfaces used in this study were plain tubes placed in horizontal array. No attempt was made to evaluate the finned/finned-tubing concept that was proposed for the partially indirectly heated cycle by Curtiss Wright.

PARTICULATE CONTROL/GAS TURBINE EXPANDER EROSION CONSIDERATIONS

The requirements for particulate removal from the combustion products of PFBC for economical gas turbine expander life are predicted to be in excess of those for meeting emission limits.[21] For that reason this assessment of the merit of the partially indirectly heated concept emphasizes the control of gas turbine expander erosion and deposition rather than particulate emissions.

Erosion of and deposition on gas turbine expander parts are functions of the concentration, size distribution, and physical properties of the particles entrained in the working fluid. (They are also functions of the size and design of the gas turbine expander, but consideration of the latter aspect is outside the scope of this study.)

The particulate in the products of combustion from fluidized beds with in situ desulfurization has three components: ash from the coal, fines from the desulfurization sorbent, and unburned carbon.

The ash from FBC of coal consists of friable platelets that have an erosivity substantially less than the fused cenospheres generated in the combustion of pulverized coal. All of the ash in the coal is assumed to be entrained in the products of combustion, either as free ash or associated with unburned carbon. Actually, small quantities of ash are carried out with the coarse spent sorbent that is removed directly from the fluidized bed. The size distribution of the free ash is based on data given in Reference 22.

For this study we have assumed desulfurization using once-through dolomite. The dolomite feed is single screened so it contains a significant amount of fines that are elutriated from the bed almost immediately after the dolomite is injected into it. The amount of excess sorbent is small (50%) and the bed volume is large, so the residence time of the coarse dolomite is long (~10 hr). As a result, there is a significant amount of attrition and decrepitation of the coarse dolomite and subsequent elutriation of the sorbent fines generated thereby. The size distribution of the attrited and decrepitated sorbent fines was also based on data given in Reference 22.

The quantity of unburned carbon (char) entrained in the combustion products from a fluidized bed is primarily a function of bed temperature and the amount of excess air. Bed depth, feed particle size and distribution, and superficial velocity also are factors. Fluidized-bed combustion efficiency is based on data given in Reference 23. The size distribution of the char is assumed to be the same as that for the coal feed for those particles having diameters smaller than the particle whose terminal velocity is equal to the bed superficial velocity. The char composition is assumed to be that of devolatilized high-volatility bituminous coal.

The concentration and size distribution of particles in the discharge of the particulate removal equipment are based on separate calculations for each of the three constituents. This individual calculation is necessary because of the differences in the density of ash, spent sorbent, and char, which have a significant effect on the performance of centrifugal separation equipment.

Appendix B contains detailed information on the concentration and size distribution of ash, sorbent, and char particles at various stages in the particulate removal subsystems for Alternative Case I. Summaries of this information are given in Figures 32 through 34 for the Base Case, Alternative Case I, and Alternative Case II. Table 19 compares the particulate loadings at the gas turbine expander inlet for each of these cases.

Table 19

EXPANDER INLET PARTICLE LOADING

Loading, g/sm^3	Base	Alternative I	Alternative II
Sorbent	0.0050	0.0033	0.0038
Ash	0.0030	0.0031	0.0025
Char	0.00004	0.00073	0.00006
	0.00804	0.00713	0.00636

This table shows that the amount of char is less than 1 percent of the total particulate for the Base Case and for Alternative Case II and about 10 percent for Alternative Case I, which has the CBC. Since there is a good possibility that the fine carbon particles in the products of combustion will be oxidized after the bypass air is added, the amount of char shown for Alternative Case I is considered to be of no significance.

The concentrations of ash in the alternative cases are within about 15 percent of the value for the Base Case so the variance is insignificant.

The only variation considered to be significant is that in the sorbent concentration, with the values for the alternative cases being 25 to 35 percent lower than that for the Base Case. Since the sorbent is the most erosive constituent of the particulate, this difference might

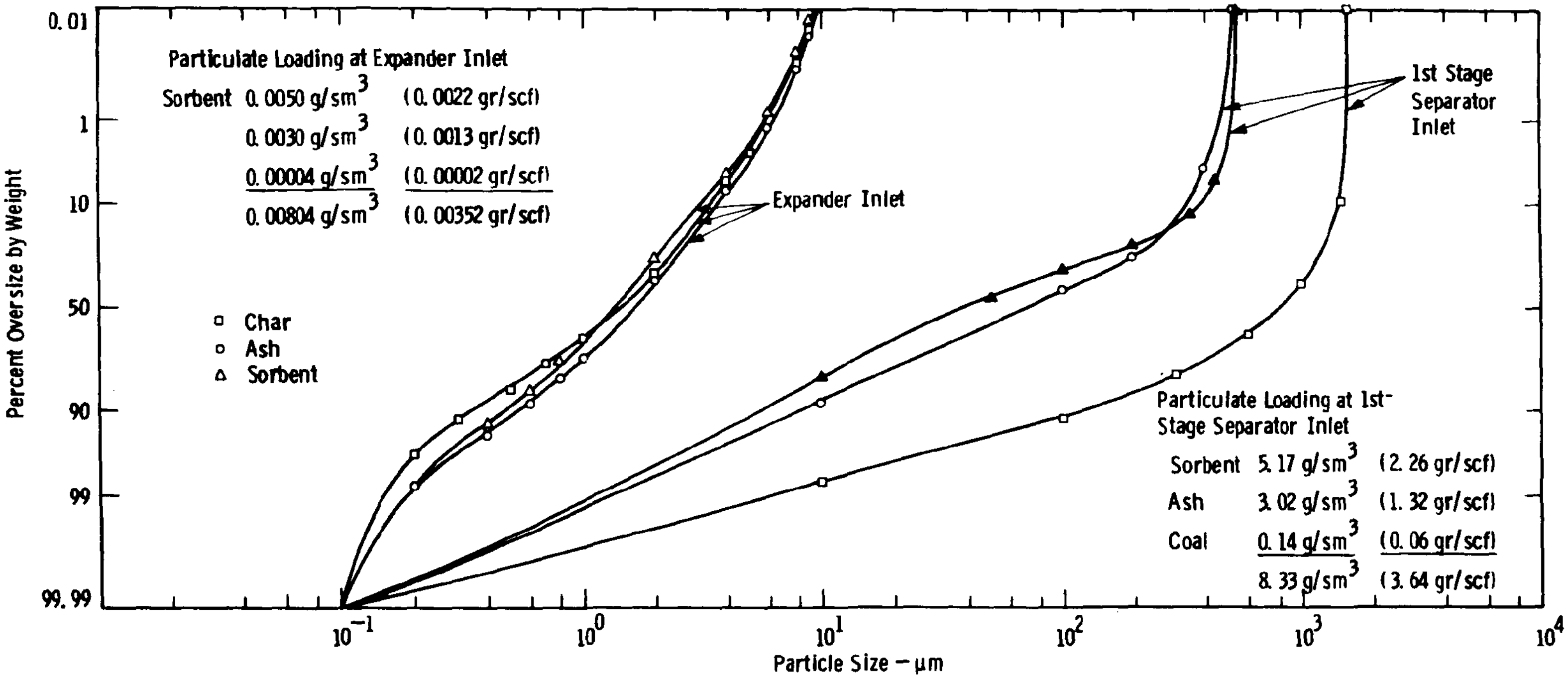

Figure 33 - Summary of Particulate Loading and Size Distribution for Base Case

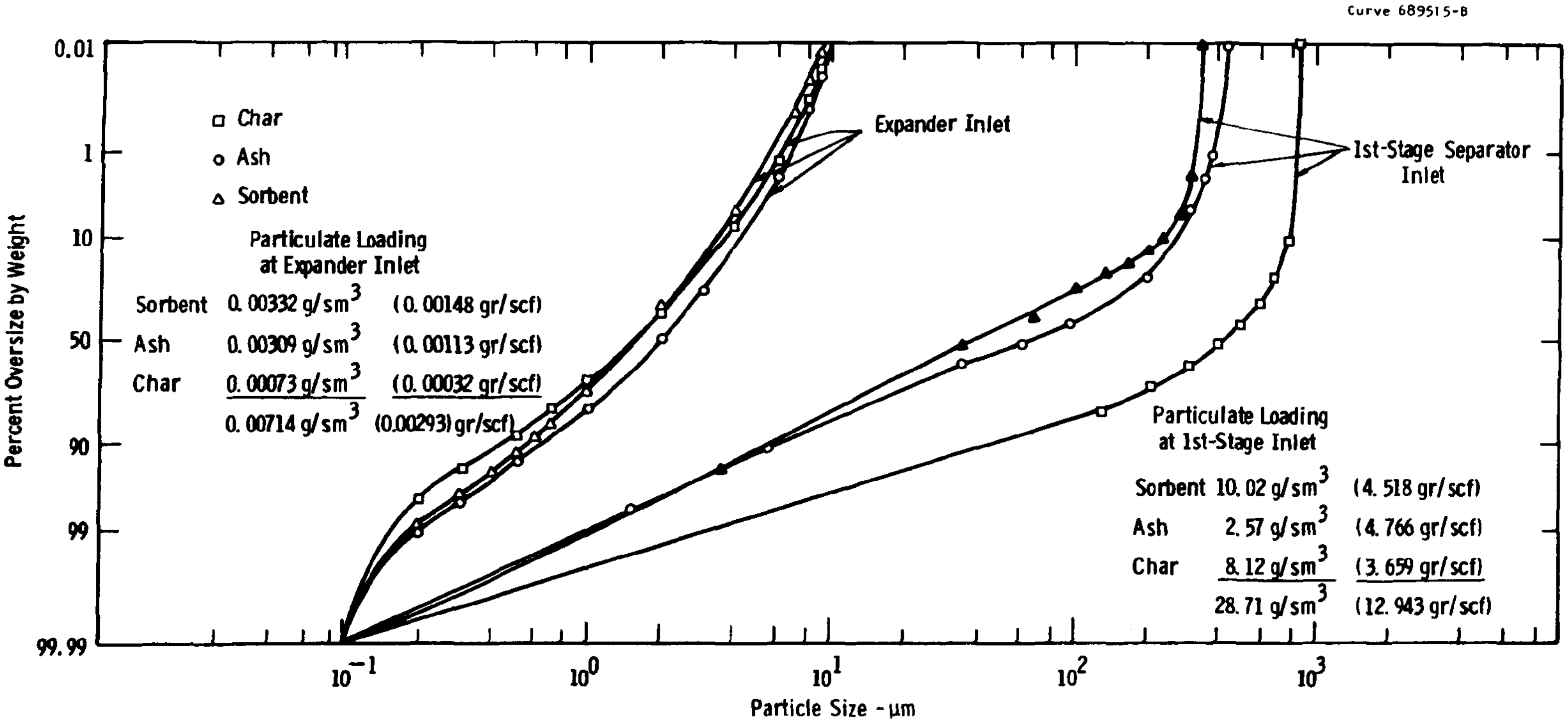

Figure 34 - Summary of Particulate Loading and Size Distribution for Alternative Case I

be expected to significantly affect the life of the gas turbine expander vanes and blades. Examination of the size distribution plots for the sorbent particles entering the expander inlet for the three cases (Figures 33 through 35) shows, however, that 100 percent of the particles are smaller than 10 μm and 50 percent are smaller than 2 μm. A recent survey[24] of turbine manufacturers indicated that 0.009 to 0.045 g/sm^3 of particles larger than 10 μm is tolerable. We conclude, therefore, that none of the cases would have an erosion problem if, in fact, a granular-bed filter or other device having the performance assumed becomes a commercial reality.

If the performance assumed for the granular-bed filter cannot be achieved in a practical separation device with the feed sorbent size distribution used here, an excessive amount of particle larger than 10 μm may be present at the expander inlet. Since a substantial fraction of the sorbent elutriated from the beds are fines present in the sorbent feed, the use of double-screened sorbent would be expected to reduce substantially the amount of sorbent in the combustion products going to the gas turbine expander. This suggests that there may be a trade-off between the cost of double-screened sorbent and the cost of replacing gas turbine expander parts. Consideration must be given, however, to the effect of the use of double-screened sorbent on the effectiveness of its desulfurization.

ENVIRONMENTAL CONSIDERATIONS

We have estimated the emission performance of the three configurations using available process models and emission data for PFBC. Using a Ca/S atom feed ratio of 1.5 for all three configurations, based on an average activity dolomite such as Tymochtee, we determined the sulfur removal efficiency. In addition, we estimated the NO_x, CO, and particulate emissions and projected the resulting solid waste product rate. These estimates are shown in Table 20.

Table 20 indicates that the three configurations would satisfy all of the current NSPS for coal-fired boilers: all SO_2 emission of

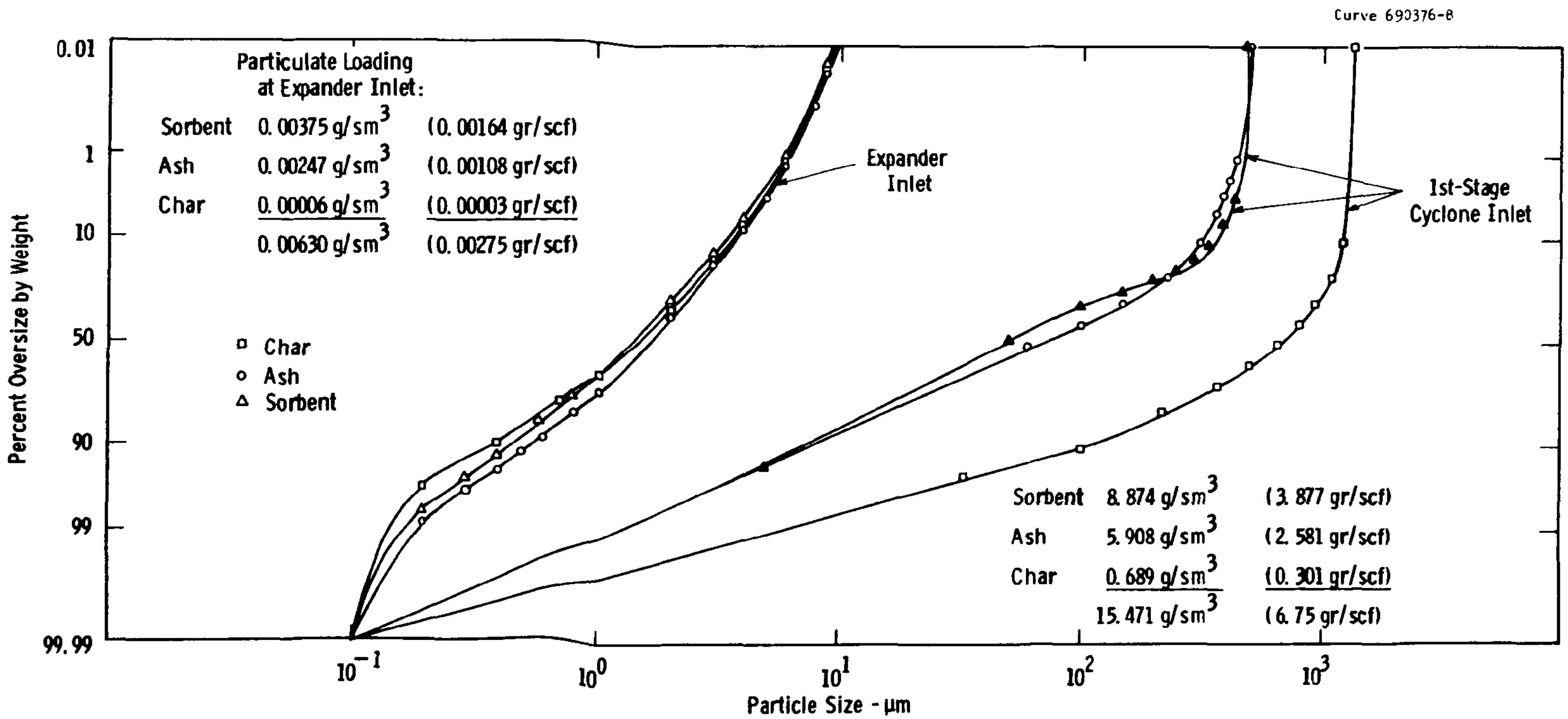

Figure 35 - Summary of Particulate Loading and Size Distribution for Alternative Case II

Table 20

ENVIRONMENTAL COMPARISON OF THE CONFIGURATIONS

Configuration	SO_2, ng/J (lb/MBtu)	NO_x, ng/J (lb/MBtu)	CO, ppm	Particulate, ng/J (lb/MBtu)	Solid Waste, kg/kg (lb/lb coal)
Base Case	284 (0.66)	215 (0.5)	50	7.3 (0.017)	0.38 (0.38)
Alternative Case I	116 (0.27)	86 (0.2)	300	7.7 (0.018)	0.38 (0.38)
Alternative Case II	116 (0.27)	150 (0.35)	100	6.5 (0.015)	0.38 (0.38)

516 ng/J (1.2 lb/MBtu), an NO_x emission of 301 ng/J (0.7 lb/MBtu), and a particulate emission of 43 ng/J (0.1 lb/MBtu). The base configuration would achieve 90 percent sulfur removal at the selected Ca/S feed ratio of 1.5, while the two options would achieve 96 percent sulfur removal because of significantly longer gas residence in the fluidized-bed combustor. Sulfur losses from the CBC have been accounted for in the estimate for Alternative Case I. The desulfurization efficiency for the Base Case could be increased at a modest cost, however, by increasing the bed depth.

The oxides of nitrogen would vary considerably among the configurations because of the variation in excess air levels. The low excess air level in Alternative Case I would result in a low NO_x emission but a relatively large CO emission. Unburned hydrocarbon emissions may also be significant in Alternative Case I, but little information is available to make such a projection.

Particulate emissions would be comparable for the three configurations and much less than the environmental limit if the particulate control system selected for turbine protection were to be used. The solid waste produced by the three configurations would be similar in rate of production and in physical/chemical properties. Small differences could exist in the particle size distributions of the waste materials, but the resulting environmental factors (e.g., leaching characteristics) would be expected to be very similar. The solid waste production could be reduced slightly in the alternative cases by operating at the smaller Ca/S feed ratios required to yield the 90 percent sulfur removal efficiency of the base configuration, or a Ca/S ratio of about 1.2 for Alternative Case I and 1.3 for Alternative Case II. The solid waste would thus be reduced to 0.34 and 0.35 kg/kg coal, respectively. This reduction in ratio could also result in a reduction in particulate emissions for the alternative cases.

The emissions from the three configurations would also be sensitive to the properties of the coal and sorbent selected for operation. The coal ash and sulfur content directly affect the fluidized-bed combustor control requirements, while dolomites vary significantly in sulfur removal activity, attrition resistance, and trace element content.

CONCLUSIONS

As a result of this evaluation of partially indirect heating of the working fluid for a gas turbine with a pressurized fluidized-bed combustor we conclude that:

- With the projected granular-bed filter performance, gas turbine expander erosion problems are not anticipated for either the base (adiabatic) configuration or the two partially indirectly heated configurations. If, however, the particulate removal performance projected herein cannot be attained on a commercial basis, both of the partially indirectly heated alternatives would have a potential for significantly larger expander life because of lower sorbent fines concentration in the gas entering the expander.

- With the replacement cost of gas turbine expander parts assumed to be uniform, the estimated costs of electricity for the three configurations considered are essentially equal. The variations in capital costs among the three cases would be balanced by variations in heat rate.
- Acceptable environmental performance is indicated for all three configurations. The partially indirectly heated alternatives indicate a potential for environmental performance significantly better than that of the Base Case.
- All three of the configurations considered have a potential for heat rates appreciably better than that of a conventional coal-fired steam plant with flue gas desulfurization.

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APPENDIX A

GRADE EFFICIENCIES FOR PARTICULATE REMOVAL EQUIPMENT

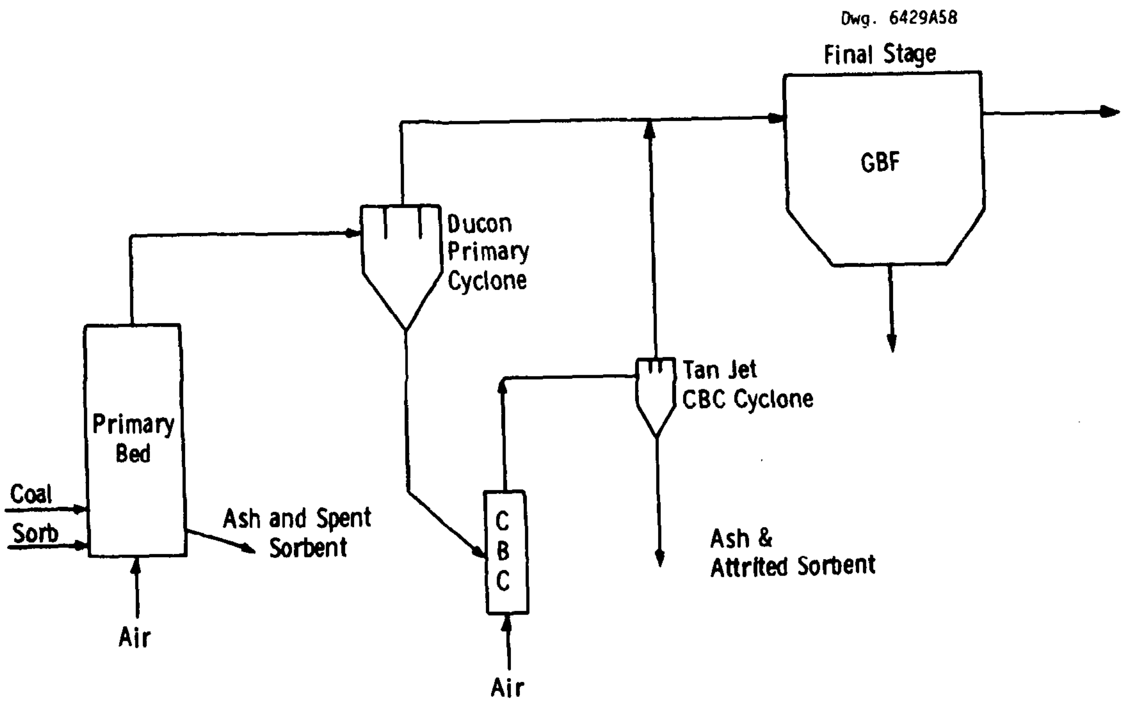

Figure A-1 - Schematic of Particle Control System

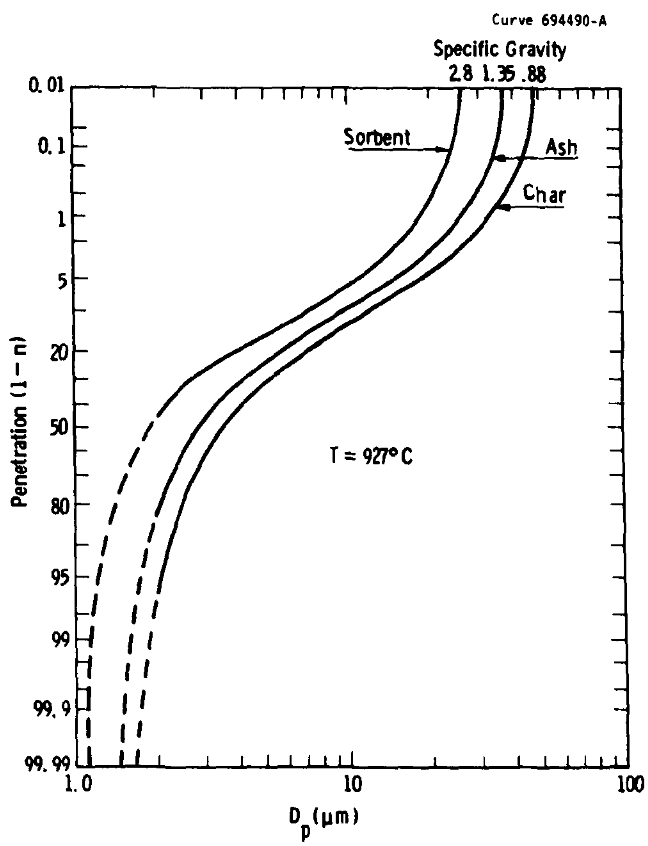

Figure A-2 - Grade Efficiency of Primary Cyclone Separator

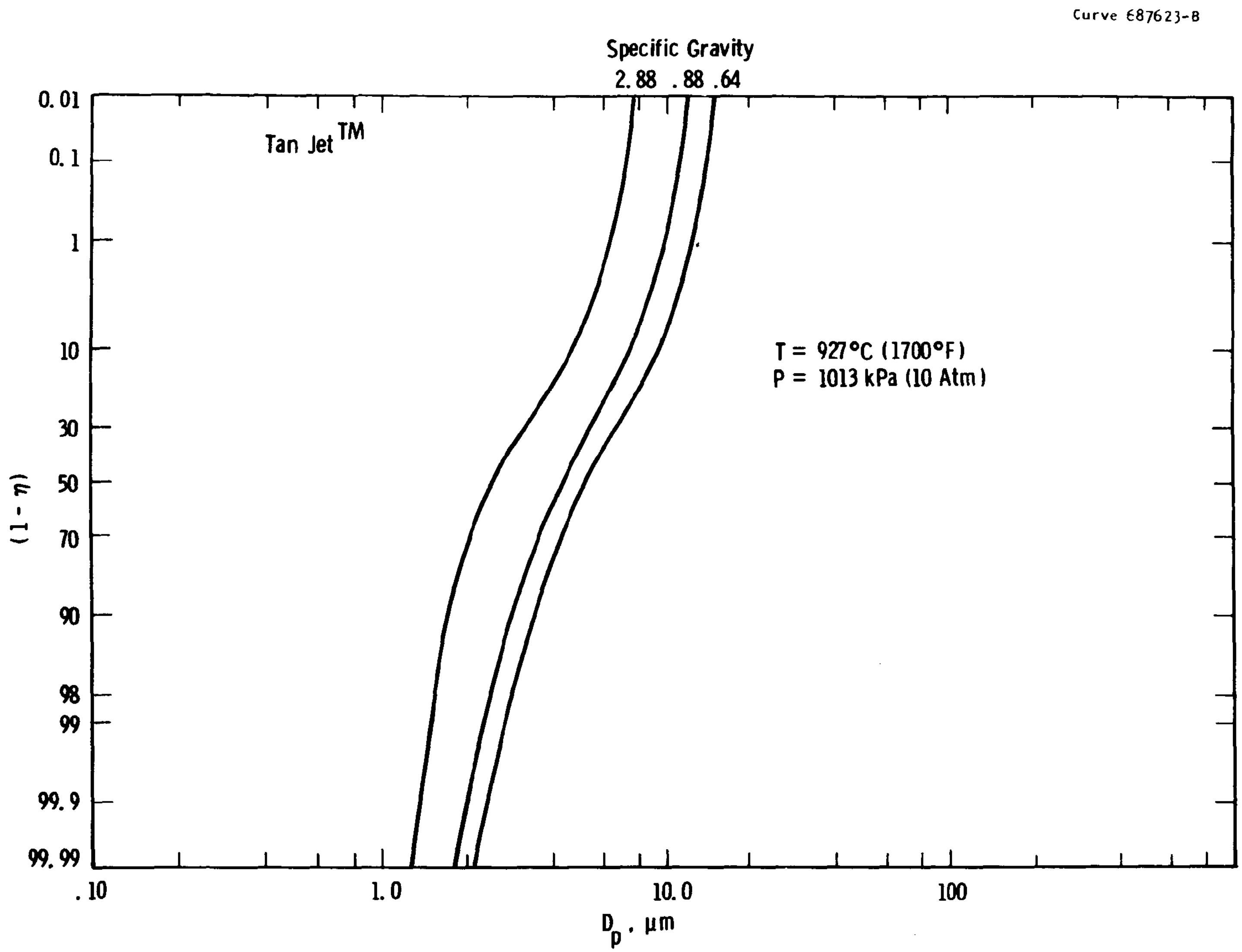

Figure A-3 - Grade Efficiency Curve for Tan Jet

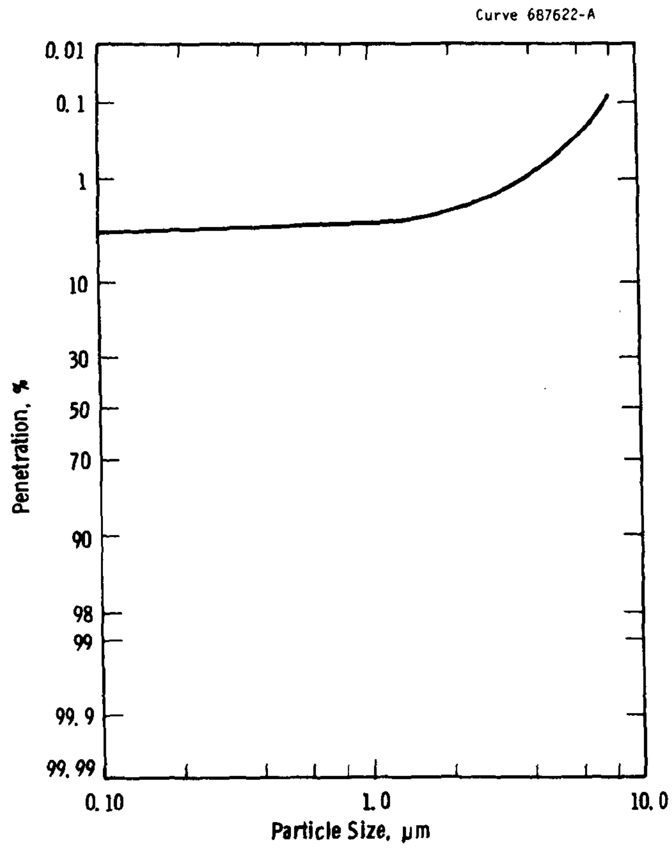

Figure A-4 - Grade Efficiency of Granular-Bed Filter

APPENDIX B

PARTICLE SIZE DISTRIBUTION AT PERTINENT STATIONS IN PARTICULATE REMOVAL SUBSYSTEM FOR ALTERNATIVE CASE I

Analyses of particle size distributions and concentration were made at each pertinent station in the particulate removal subsystem for each case studied. Plots of the size distribution at each station for Alternative Case I are included in this appendix. Figure B-1 gives the size distribution plots for sorbent particles, Figure B-2 gives those for ash particles, and Figure B-3 gives those for char particles.

The concentrations of sorbent, ash, and char particles at each of the pertinent stations are given in Table B-1.

Table B-1

Station	Solids Concentrations, g/sm^3			
	Sorbent	Ash	Char	Total
Primary Bed Exit	10.35	10.92	8.38	29.64
1st Stage Separator Exit	0.55	0.69	0.21	1.44
Carbon Burnup Cell Exit	9.52	8.86	1.08	19.46
CBC Separator Exit	1.72	1.31	0.021	3.04
GBF Inlet	0.67	0.76	0.20	1.63
GBF Exit	0.011	0.0085	0.0025	0.022
GT Expander Inlet	0.0034	0.0026	0.0007	0.0067

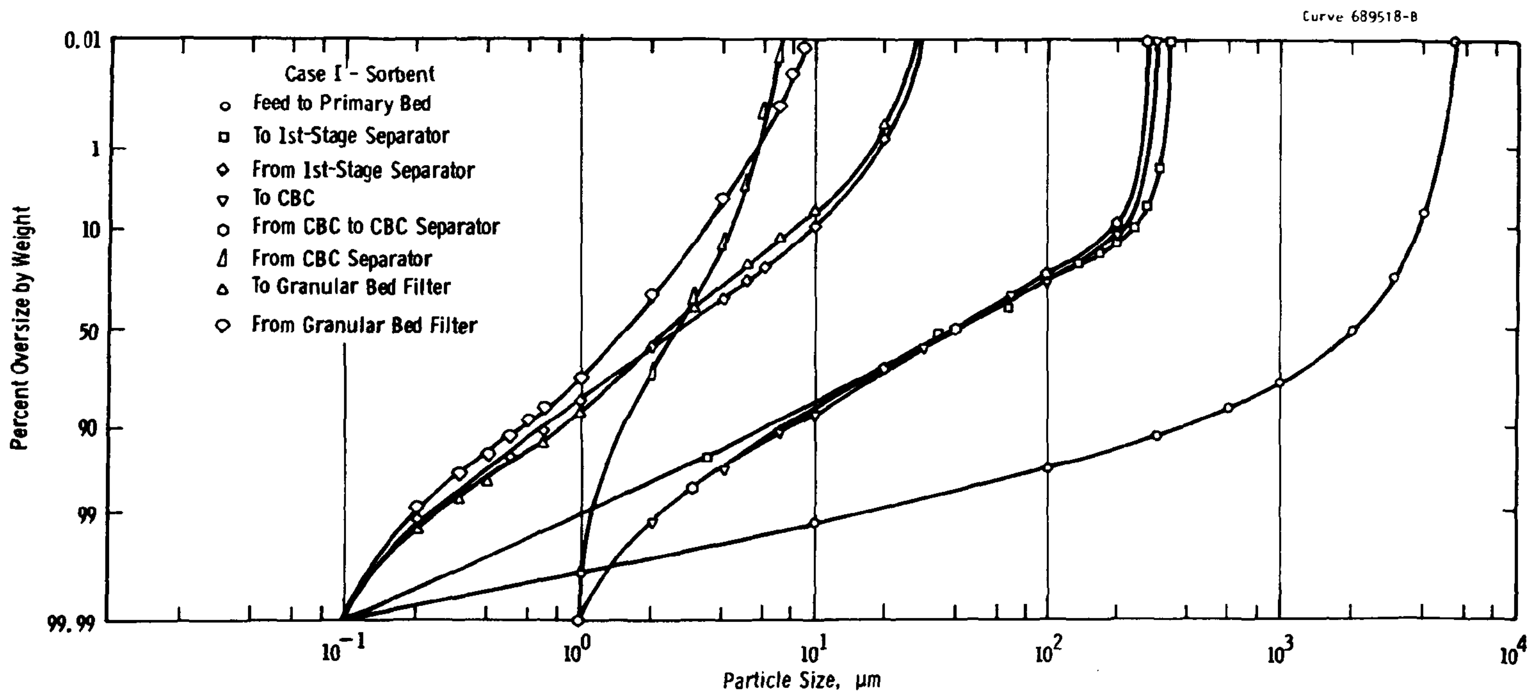

Figure B-1 - Size Distributions of Sorbent Particles for Alternative Case I

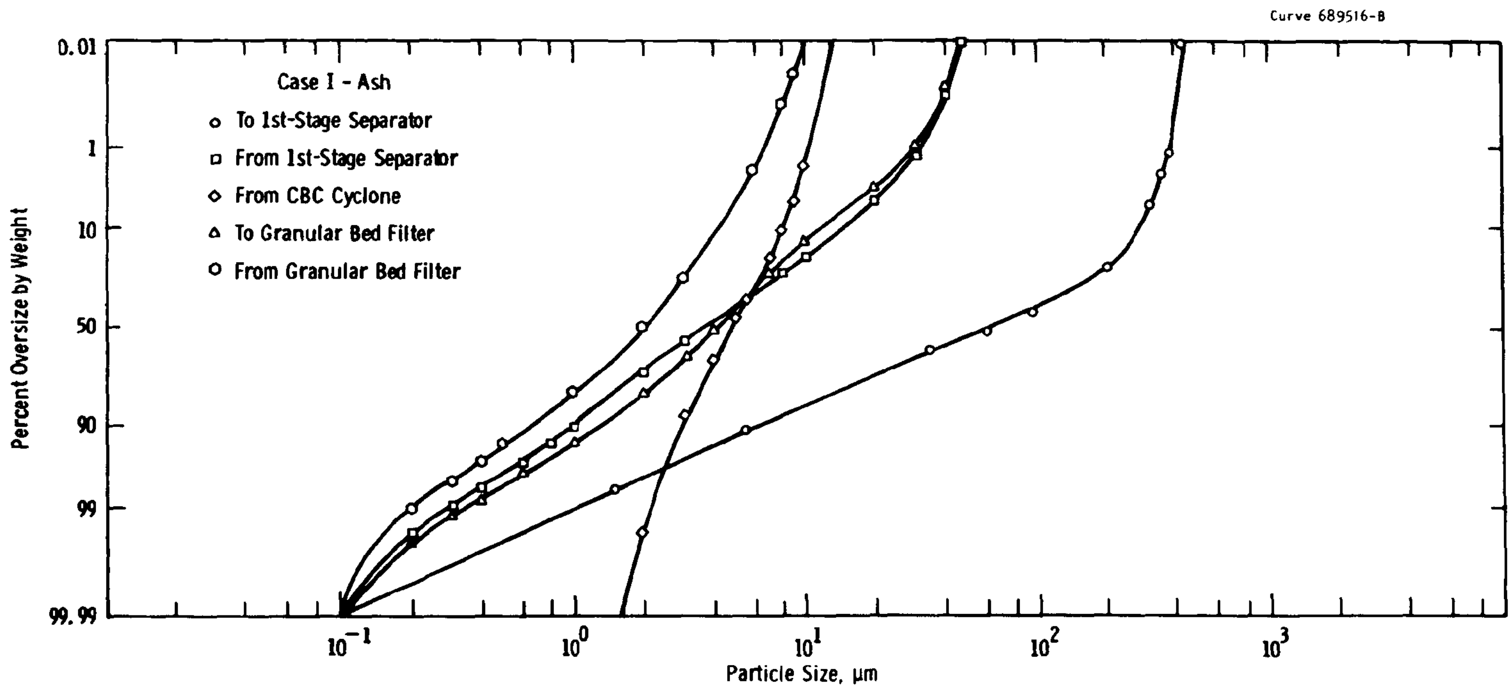

Figure B-2 - Size Distributions of Ash Particles for Alternative Case I

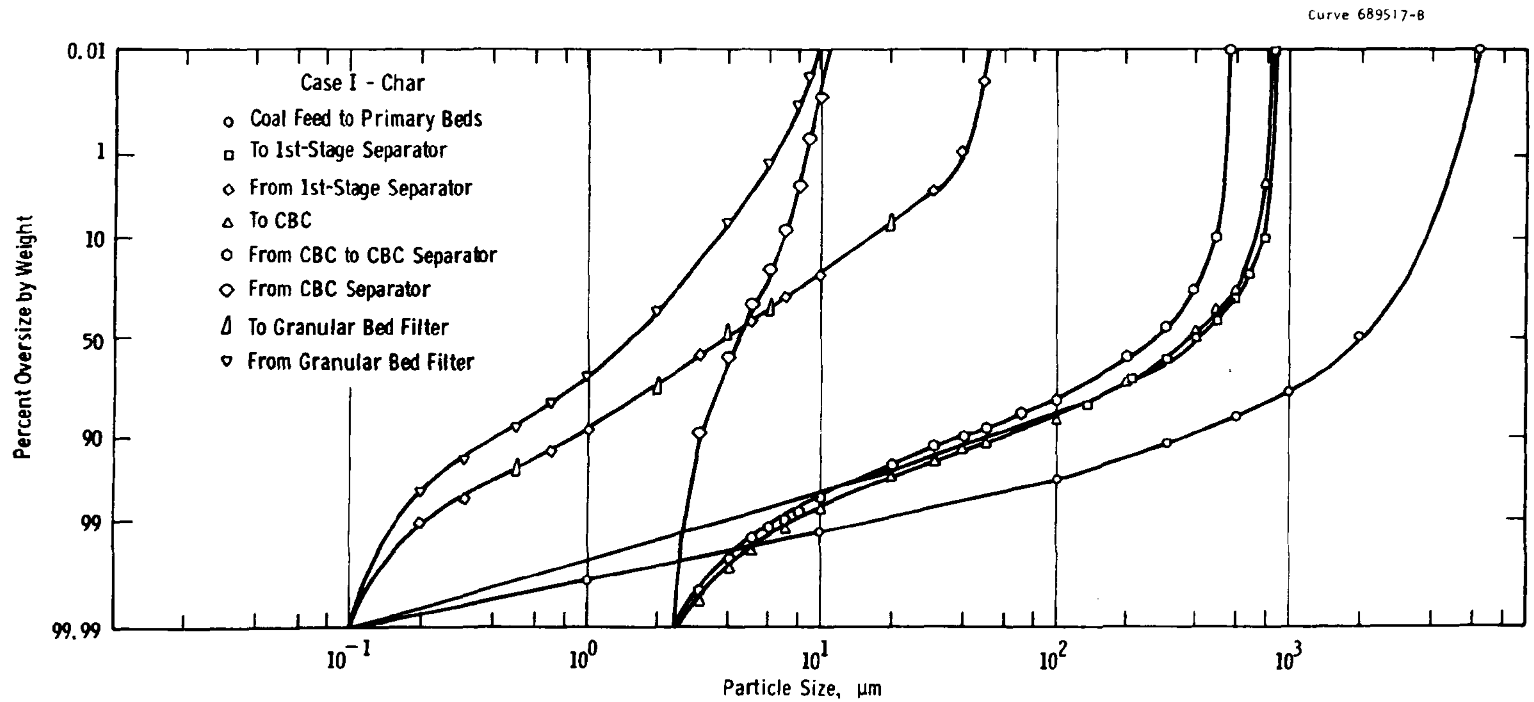

Figure B-3 - Size Distributions of Char Particles for Alternative Case I

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16. ABSTRACT The report gives results of engineering studies addressing several aspects of fluidized-bed combustion (FBC) system design and performance, as applied to coal. It reviews an evaluation of the impact of SO2 emission requirements on FBC system performance and cost. Stringent SO2 emission requirements can be satisfied economically if design and operating parameters are properly selected. An alternative SO2 control concept for pressurized FBC (PFBC), pressurized scrubbing of the products of combustion with water, is evaluated. The concept is not economically competitive because of reduced plant efficiency and the need for recuperative heating. A potential reduction in solid waste is realized with the concept, but the SO2 control efficiency may be limited. An evaluation of PFBC, examining the technical and economic trade-offs between the level of particulate control achieved and the frequency of gas-turbine blade replacement, is described. The evaluation incorporates models of PFBC particulate carry-over, particulate control device efficiency, and turbine erosion. Also, an indirect air-cooled PFBC concept is compared with other PFBC concepts. The indirect air-cooled concept provides significant particulate control advantages over the adiabatic combustor PFBC concept, resulting in about 4% lower plant efficiency and 1% higher cost of electricity.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Dust	Pollution Control	13B	11G
Combustion	Gas Turbines	Stationary Sources	21B	13G
Fluidized Bed Processing		Particulate	13H,07A	
Coal			21D	
Sulfur Oxides			07B	
Scrubbers			13I	

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