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DR# 0050-7

DOE/MC/16049-1558. Vol. I
(DE84003090)

HOT-GAS CLEANUP SYSTEM MODEL DEVELOPMENT

Volume I

Final Report

**By
K. Ushimaru
A. Bennett
P. J. Bekowies**

November 1982

Work Performed Under Contract No.: DE-AC21-81MC16049

**For
U. S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia**

**By
Flow Industries, Inc.
Research and Technology Division
Kent, Washington**

**Technical Information Center
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Foreword

This study of Hot-Gas Cleanup System Model Development was conducted under Contract No. DE-AC21-81MC16049 for Morgantown Energy Technology Center of the United States Department of Energy. The direction and valuable comments from the past and present project officers, Dr. L. E. Graham and Ms. R. A. Bajura, respectively, have been greatly appreciated.

The authors are grateful for the cooperation and dedicated effort of the individuals who have been actively involved in this study of Hot-Gas Cleanup System Model Development. The authors are particularly indebted to Dr. G. D. Case of Resource Technologies Group, Inc., and Dr. H. Huynh and Ms. M. L. Johnson of Flow Industries, Inc., for their technical assistance.

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Nomenclature

a	=	particle radius, m
A	=	cross-sectional area, m^2
C_s	=	Cunningham slip correction factor, dimensionless
d_c	=	collector fiber/granule diameter, m
d_p	=	particle diameter, m
d_{pa}	=	aerodynamic particle size, m
d_w	=	wire diameter, m
E	=	electric field strength, V/m
g	=	acceleration due to gravity, 9.8 m/s^2
k	=	Boltzmann's constant, $1.38 \times 10^{-23} \text{ J/molecule} - ^\circ\text{K}$
K_C	=	electrostatic force parameter, dimensionless
K_D	=	diffusion parameter, dimensionless
K_I	=	direct interception parameter, dimensionless
K_{II}	=	inertial impaction parameter, dimensionless
K_S	=	gravitational settling parameter, dimensionless
l	=	characteristic dimension, m
L	=	length of wire electrode, m
N_{Kn}	=	Knudsen number, dimensionless
N_{Pe}	=	Peclet number, dimensionless
N_{Re}	=	Reynolds number, dimensionless
N_{St}	=	Stokes' number, dimensionless
P	=	pressure, N/m^2
Q	=	volumetric gas flow rate, m^3/s
r_o	=	radius of wire electrode, m
t	=	time, s
U_f	=	face velocity, m/s
U_g	=	superficial gas velocity, m/s
U_s	=	gravitational settling velocity, m/s
U_t	=	tangential velocity of gas, m/s
V_o	=	applied voltage, V
w	=	migration velocity, m/s
z	=	bed depth, m
ΔP	=	system pressure drop, N/m^2

Nomenclature (Cont.)

ϵ	=	bed porosity, fraction
ϵ_0	=	initial bed porosity, fraction
η	=	collection efficiency, dimensionless
κ	=	permittivity constant of space, $C^2/N\text{-}m^2$
λ	=	mean free path, m
μ	=	gas viscosity, $N\text{-}s/m^2$
ρ_c	=	collector density, kg/m^3
ρ_g	=	gas density, kg/m^3
ρ_p	=	particle density, kg/m^3
σ	=	inlet particle concentration, kg/m^3
τ	=	relaxation time, s

1. Executive Summary

Pressurized, fluidized-bed combustion (PFBC) for electric power generation provides a direct combustion process for coal and low-grade fuels with the potential for improved thermal conversion efficiency, reduced costs and acceptable environmental impacts. For the successful operation of PFBC combined-cycle power plants, particulate removal from the combustion off-gases at operating temperatures and pressures is critical. The gas cleanup system must be capable of reducing the particulate loading in the combustion off-gases to levels compatible with gas turbine operating conditions and environmental standards.

The purpose of this research effort is to provide computationally efficient mathematical models and associated computer codes to describe the performance of advanced high-temperature, high-pressure (HTHP) gas cleanup devices, designed to protect gas turbines in a PFBC combined-cycle power generation plant. The computer programs for predicting the performance of such devices are compatible in terms of input/output parameters and subroutine requirements with the large-scale process simulation program ASPEN (Advanced System Process Engineering), which is currently in use by Morgantown Energy Technology Center (METC). The resultant set of computer programs will thus allow a plant designer to conduct overall system simulations of a PFBC combined-cycle power generation plant.

Significant Contributions

Most of this research effort was spent (1) gathering existing information, (2) evaluating and reporting the information, and (3) transcribing the information into a uniform set of ASPEN compatible computer codes. This in itself was a significant accomplishment. Additional significant contributions include:

- (1) The development of a uniform software framework system applicable for any cleanup device.
- (2) An independent evaluation of common HTHP cleanup devices.
- (3) The development of a rough wall pipe pressure drop model for the Research Cottrell (RC) electrostatic precipitator (ESP).

- (4) The development of an improved computer collection efficiency model for the RC ESP.
- (5) Improvements to the Combustion Power Company (CPC) granular bed computer model to include the critical state of the fluidized bed, i.e., whether blow-out is imminent.
- (6) The first effort made toward development of a uniform costing algorithm for HTHP cleanup devices.

The Report

This two-volume report summarizes the state of the art in performance modeling of advanced HTHP gas cleanup devices. Volume I contains the culmination of the research effort carried over the past 12 months and is a summary of research achievements. Volume II is the user's manual for the computer programs developed under the present research project. In this volume, Section 2 presents background information on PFBC concepts, a description of the role of the advanced gas cleanup systems, and a list of advanced gas cleanup systems that are currently in development under DOE sponsorship. Section 3 describes the methodology for the software architecture that forms the basis of the well-disciplined and structured computer programs developed under the present project. Section 4 reviews the fundamental theories that are important in analyzing the cleanup performance of HTHP gas filters. Section 5 discusses the effect of alkali agents in HTHP gas cleanup. Section 6 evaluates the advanced HTHP gas cleanup models based on their mathematical integrity, availability of supporting data, and the likelihood of commercialization. As a result of the evaluation procedure detailed in Section 6, five performance models were chosen to be incorporated into the overall system simulation code, ASPEN. These five models (the electrocyclone, ceramic bag filter, moving granular bed filter, electrostatic granular bed filter, and electrostatic precipitator) are described in Section 7. The method of cost projection for these five models is discussed in Section 8. The supporting data and validation of the computer codes are presented in Section 9, and finally the conclusions and recommendations for the HTHP gas cleanup system model development are given in Section 10.

2. Introduction

Fluidized-bed combustion potentially offers several advantages over conventional methods of firing coal, including improved thermal efficiency, reduced costs and acceptable environmental impact. The technology involves the combustion of coal in a bed of limestone or dolomite and inert ash. The bed is maintained in a fluidized state by the passage of combustion air through a gas distributor located at the base of the bed. Coal is introduced to the bed and is mixed with the bed material. It then burns, releasing energy which heats the bed material. The limestone or dolomite in the bed reacts with the sulfur dioxide formed during the combustion of the coal to form a solid sulfate material that can be disposed of as a dry solid waste along with the coal ash. Energy can be extracted from the bed either indirectly, by immersed heat exchange surfaces that heat a working fluid such as steam or air, or directly, by heating excess air passed through the bed. Variations in the technology also include operation at atmospheric or elevated pressures. Both open- and closed-cycle turbine systems have been proposed for PFBC.

Studies of the integration of fluidized-bed combustors into large power systems have concluded that the most cost-effective power systems would use a combined-cycle plant (open-cycle gas turbine and steam turbine) in which the fluidized-bed combustion system operates at high pressures. Using a sulfur absorbing bed material, SO_x emission standards can be met and NO_x emissions are inherently low. Calculated efficiencies of up to 41% have been reported (Bertrand, 1976).

There are three PFBC systems now being actively investigated. The first and simplest involves a bed with no internal heat exchange surface; all the heat is removed by direct heating of excess air. The mixture of hot air and combustion gases, after particulate removal, is expanded through a gas turbine for energy recovery. Since no heat is removed from the fluidized bed, it is called an adiabatic burner. The second concept involves indirect heating of

air passed through heat exchange tubes located in the fluidized bed. The heated air is mixed with combustion gases and expanded through a turbine. The third variation involves raising steam by in-bed heat exchanger tubes and combining both a steam turbine and a gas turbine to generate power.

For particulate cleanup, the adiabatic concept is the most difficult to deal with. About 300% excess air is mixed with the combustion gases and this entire stream must be cleaned before exhausting into the turbine. The other two concepts require about 15-35% excess air, since heat is removed indirectly from the bed by air or steam heat exchange tubes. Therefore, the adiabatic concept requires the treatment of a much larger volume of gas, increasing the size and cost of the cleanup system compared to the other two concepts.

For the two concepts that include bed cooling, particulate cleanup requirements are less severe for the air-cooled system than for the water-cooled system. In the air-cooled system, the combustion exhaust gases are mixed with the clean, indirectly heated air before exhausting into the turbine. This dilution effect reduces the concentration of particulates going into the gas turbine by about one-third.

The key issue that must be solved in further developments in this technology is reliable and efficient removal of particulates and other fouling matters from the combustion products at high temperatures and pressures. Removal of ash particles and alkali compounds from the products-of-combustion stream that flows through the gas expansion turbine is necessary to prevent turbine erosion and corrosion and must be done at high temperatures and pressures to preserve the thermal efficiency of the cycle. Such hot-gas cleanup devices may, when developed, also be applicable to the cleanup of fuel gases from gasification of coal, leading to improved efficiency of gasification combined-cycle systems. Following is a list, not necessarily complete, of HTHP gas cleanup systems that are being developed under DOE sponsorship.

COMPANY	CONCEPT
General Electric Company	Electrostatic Granular Bed Filter
	Electrocyclone
Westinghouse	Ceramic Cross-Flow Filter
	Fixed Granular Bed Filter
State University of New York at Buffalo, NY	Acoustic Agglomerator
Air Pollution Technology, Inc.	Dry Plate Scrubber
Acurex	Ceramic Bag Filter
Research Cottrell	High-Temperature/Pressure Electrostatic Precipitator
Combustion Power Company	Moving Granular Bed Filter
Mechanical Technology, Inc.	Cyclocentrifuge

Mathematical modeling activities to describe and understand the detailed process dynamics and particulate removal mechanisms are being conducted by several of the developers listed above. In addition, a considerable literature base exists in detailed process modeling for various particle removal concepts.

The purpose of this work is to provide computationally efficient mathematical models and associated computer codes to describe the performance of advanced HTHP gas cleanup devices designed to protect gas turbines in a PFBC combined-cycle power generation plant. All of the devices listed above were included in the device and model evaluation task of this work except the cyclocentrifuge by Mechanical Technology, Inc.

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3. Software Engineering

Computationally efficient mathematical models and computer codes that describe the performance of advanced HTHP gas cleanup devices have been developed during this effort. The objectives were to develop a set of codes that accurately represent available models, that are efficient to use, and that are easy to read and maintain. The design of our software followed the life-cycle stages of (1) system analysis, (2) system specifications, (3) algorithm design, (4) detailed design, (5) coding, (6) testing, (7) maintenance, and (8) operation. All of the codes developed as part of this effort have been coded in ANSI FORTRAN X3.9-1966 standard.

The deliverable codes interface with the ASPEN simulation code developed for the Department of Energy by the Massachusetts Institute of Technology (MIT). Our codes meet or exceed the design standards set by MIT. A concise description of the ASPEN code is given in Section 3.1. The general algorithm, the coding framework, and the naming conventions for the physical and cost models are discussed in Sections 3.2 and 3.3.

3.1 The ASPEN Simulation Code

In 1976 the Department of Energy contracted with MIT to develop an advanced computer code that included the capability to simulate gas, liquid, and solid flow systems in chemical, coal, and nuclear processing plants. The result of that \$6 million effort was the ASPEN code delivered to the Morgantown facility in the fall of 1981.

The ASPEN code consists of more than 375,000 lines of FORTRAN statements requiring more than 27 million bytes of disc space to store the source and associated data banks. The code has been very well received by government and industrial users. In addition, an ASPEN Users Group meets concurrently with the American Institute of Chemical Engineers (AIChE). The code is now distributed for the U.S. Government by the National Energy Software Center located at the Argonne National Laboratory and by commercial vendors, two of which are composed of members from the original MIT development team.

Basically, the ASPEN code is composed of a keyword translator and FORTRAN code generator. The user creates a computer input file containing a set of keywords and phrases structured in a manner familiar to process engineers.

The input file is then submitted as a batch job to the input translator, which parses, verifies, and validates the command phrases. The code generator creates a set of FORTRAN programs and data arrays representing the process to be simulated. The data arrays are constructed from a vast data file that contains all the pertinent information required. After that, the FORTRAN programs are submitted automatically to the computer operating system to be compiled, linked, and executed, and the simulation results are returned as a computer file that can be examined with a CRT or spooled out as a typical computer printout.

New models can be added to the ASPEN system in two ways. The formal way is to design models that are included in the ASPEN library and called directly from ASPEN. The advantage of this method is that all the proper keywords can be implemented and available to the user. However, implementation of this method requires a great deal of knowledge and experience about ASPEN. A more convenient means is to develop a user subroutine interface for the new model. The advantage of the user interface is that the developer requires only a limited knowledge of ASPEN. The disadvantage of this method is that keywords cannot be used and input data needed by the user must be inserted directly in the ASPEN code as data arrays that are rigidly structured. Because of our lack of experience with ASPEN, we chose the user model interface method for the development of the HTHP cleanup codes.

In an actual HTHP gas cleanup device, there is a single hot-gas stream entering the device and two output streams leaving. The input stream contains the hot gas and particulates. The two output streams are the cleaned gas stream and the ash stream.

To simulate the process in ASPEN, the user creates a single input data stream by combining two or more data substreams that serve as the input for the computer gas cleanup model. The first is normally a conventional gas substream composed of components that can be obtained directly from the ASPEN library. The second is normally a conventional particulate substream developed in the same way as the conventional gas substream. The third substream is normally a particulate substream of nonconventional chemicals defined entirely by the user. The substream data are linked together in a single one-dimensional array as the primary input data stream for the cleanup models.

3.2 HTHP Cleanup Models

Basically, the input data stream for the gas cleanup models consists of a one-dimensional ASPEN array that contains the description of the physical hot-gas stream. The data stream must be identified by sections, and then individual parameters, such as temperature, pressure, etc., must be identified for the cleanup model. A set of pointers (subscripts) are created to accomplish this task for the cleanup models.

The output data streams from the cleanup model consist of two one-dimensional data arrays, which consist of the cleaned input data stream and the ash removal data stream.

The results of the cleanup simulation are written to the ASPEN HISTORY file, which contains all the critical summary data.

Since the models are designed to be used as ASPEN user models, the user data entered into the model procedure through the ASPEN REAL array must be verified by the error trapping procedure before it is passed on to the model.

Also obtained from ASPEN are the necessary controls such as the initialization and report writing flags.

Although each hot-gas cleanup device is unique in configuration and means of accomplishing particulate removal, all devices are designed to do the same thing, i.e., to remove particles from a stream of hot coal gas. This device functional commonality implies a possible computer modeling commonality. The basic design premise for the cleanup models was that a common algorithm existed for all models. To take maximum advantage of this opportunity, more initial design effort was required; but that effort was more than repaid during the final implementation stage and will be especially rewarding during the maintenance part of the code's life-cycle. The major life-cycle cost for these codes will be the time spent in changing and revalidating the codes as the models are improved due to changes in device technology. Since it is most likely that different people will be responsible for maintaining the codes over their life-cycle, it is imperative that they be easy to understand and modify.

The basic design tools available to the software engineer are pseudocodes, modularity, and structured program constructs. Pseudocoding is a means of flow charting by using English-like statements that can be converted directly into structured code constructs. Modularity implies that tasks are isolated to specific sections of code and are kept as short as possible, normally no longer than two pages of code. The structured constructs are (1) the simple sequence statement, (2) the IF ... THEN ... ELSE block of code, (3) looping using the DO WHILE and DO UNTIL coding, and (4) the CASE construct. One of the proofs of computer science is that the proper use of these four constructs can be used to solve any problem. The proper use of these constructs implies that the code reads top-down, i.e., it begins at the top and flows down the page to the end.

The following algorithm was found to serve well as the framework for all of the gas cleanup devices.

```
Begin procedure;
  if this is an initialization pass, then;
    set pointers;
    get user input from REAL array;
  end if;

  define simulation and/or report pass;

  if this is a simulation pass, then;

    calculate input flow and pressure drop;
    determine auxiliary power requirement;

    if data correct, then;
      apply filter model;
    end if;

  end if;

  transfer results to output stream;

  if this is a report pass, then;
    write to the HISTORY file;
  end if;

  return to ASPEN;

end of procedure.
```


The actual FORTRAN programs for the algorithm consist almost entirely of IF and subroutine CALL statements and are liberally sprinkled with enough comment statements to ensure self-documentation. Each task listed in the above algorithm is isolated to a separate subroutine. Two global procedures to set the pointers and transfer the results to the output stream are common to all cleanup models. All other tasks in the algorithm are unique to each cleanup device model.

Because of the large number of routines required for this effort, a naming convention consistent with the requirements of ASPEN and METC was established for the codes. All user interface modules begin with the character string USR followed by a three-letter code identifying each cleanup device model. For example, the name of the General Electric electrocyclone model subroutine is USRGCY. In agreement with METC, the name of all subordinate modules called by the user interface routine (USR---) begins with the letter Q with the global routines mentioned above beginning with the letters QQ.

A list of all cleanup device routines and their FORTRAN names is found in Volume II along with the actual codes.

3.3 Cost Models

The cost models are the simplest of the codes developed. Unlike the multiple-module physical cleanup models, the cost codes are single modules, i.e., there are no subordinate routines. Since they are simple codes, they were written directly as ASPEN routines, using the same interface between the cost model and the ASPEN code as that used by the developers of the original ASPEN code.

The design philosophy for the cost models was the same as that used for the physical cleanup models, i.e., use a single basic algorithm for all devices. However, there is a significant difference between the output generated by the cost models and by the physical models. In the physical models, only one printout is supplied. In the cost models, the user is provided a choice of three printouts: a complete detailed report or a choice of two levels of summary costs. This option is also available in other ASPEN cost models developed by MIT.

The algorithm for the cost model for the GE electrocyclone is listed below.

Begin procedure;

calculate single module direct equipment cost (base 1981);

cost of carbon steel;

cost of 310 stainless steel;

cost of refractory material;

cost of electric insulators;

cost of power supply;

==

total direct equipment cost;

adjust total for miscellaneous equipment;

adjust equipment total for cost variance since 1981;

calculate single module installed cost;

indirect cost (% of total cost)

debt service on 90% of total cost

==

total indirect and debt service;

single module installed cost;

calculate effective number of modules required;

calculate total installed cost for all modules;

normalize installed cost to plant capacity (\$/MW-hr);

calculate single module operating cost;

auxiliary power cost (\$/hr);

maintenance cost (\$/hr);

ash removal cost (\$/hr);

==

total operating cost per module (\$/hr);

calculate operating cost for all modules;

normalize operating cost to plant capacity (\$/MW-hr);

calculate total projected cleanup cost (\$/MW-hr);

determine ASPEN report to be printed;

if ASPEN level 4, then;

print a summary report;

else, if ASPEN level 5, then;

print report showing all totals;

else, if ASPEN level 6, then;

print a complete report;

end if;

return to ASPEN;

end of procedure.

The only differences between the cost models for the various cleanup systems are in the device components required. For example, the ceramic bag does not require a set of electric insulators but it does require a back-flush system. The differences are in the items required, not in the form of the costing algorithm.

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4. Analytical Performance Models

When analyzing and evaluating the performance of particulate control devices, it is important to understand the physical mechanisms by which the particles are removed from the primary gas stream. Particularly in HTHP operations, where the performance of existing models remains unproven, a firm understanding of the temperature and pressure effects is essential. If a performance evaluation is based on sound theoretical analysis, the results will provide future guidelines for the design and scale-up of existing laboratory models.

In this section, the primary particulate collection mechanisms are described. The variables involved in these physical mechanisms are grouped mathematically to aid in conducting a parametric study of temperature and pressure effects.

4.1 Fundamental Theories

In actual computation, the degree of particle collection is expressed in terms of the so-called collection efficiency, η . The collection efficiency is affected by the following physical processes:

- (1) Inertial impaction
- (2) Diffusiophoresis
- (3) Gravitational settling
- (4) Direct interception
- (5) Electrical migration

These physical processes have been identified as the key particle collection mechanisms in gas cleanup devices. The general approach in the development of a mathematical model for particle collection has been to assign the parameters associated with the above processes (Shackleton and Kennedy, 1977; Parker et al., 1977). These parameters indicate the degree of importance of each mechanism in the overall particle collection process. In other words, if the inertial impaction parameter is found to be large as compared to the others for a given device, one can conclude that the fundamental collection mechanism for that device is inertial impaction. One can then develop an analytical model for collection efficiency based on the inertial impaction parameter.

The fundamental theories associated with the development of analytical performance models are discussed below. These theories will be used to define the key parameters for particle collection in the following section.

Mean Free Path and Knudsen Number

The average distance traveled by a gas molecule before colliding with other molecules is an important parameter in analyzing a certain class of flow problems. This distance is called the mean free path and is a function of molecular density.

Momentum transfer in a gas is due to molecular collisions. If the mean free path is large as compared to the aerosol particle diameter, the momentum transfer takes place in a diffused manner over relatively large distances from the particle. Consequently, the moving particle has little effect on the velocity distribution of the neighboring molecules.

If the mean free path is much smaller than a typical particle dimension, molecules rebounding from the particle may be expected to collide with other molecules near the particle. These collisions influence the molecular velocity distribution near the particle. In particular, the velocity distribution of molecules that are about to strike the particle will be different from the distribution observed when the mean free path is relatively large. Consequently, the mathematical expressions used to compute the force exerted by the fluid on the particle are different in the two cases.

The relative magnitudes of the mean free path, λ , and a characteristic particle dimension, a , are significant in classifying the flow problems. The ratio of the mean free path to the particle radius is the Knudsen number, defined as

$$N_{Kn} = \frac{\lambda}{a} \quad . \quad (4.1)$$

Flows where the Knudsen number is large are termed free molecule flows and are treated by using methods of statistical mechanics. Flows where the Knudsen number is much less than unity have been successfully analyzed using continuum models.

The region where the Knudsen number is near unity is termed the transition regime. As may be expected, analysis is most difficult in this regime and

empirical drag relations are often used. For small particles ($\leq 1 \mu\text{m}$), a continuum slip flow approximation, often referred to as the Cunningham slip correction, has been employed to yield first-order corrections. The Cunningham slip correction factor is calculated as

$$C_s = 1 + N_{Kn} [1.257 + 0.40 \exp (-1.1/N_{Kn})] \quad (4.2)$$

Equation (4.2) is an empirical expression based on Millikan's oil drop experiments conducted at room temperature and reduced pressures. At extreme conditions, the Cunningham slip correction factor remains unproven.

Stokes' Law

A basic approach in analyzing the particle transport in a continuum flow involves a solid sphere moving through a continuous viscous gas. For general flow problems in a continuum regime ($N_{Kn} \ll 1$), we define the Reynolds number, N_{Re} , as

$$N_{Re} = \frac{\ell U_g \rho_g}{\mu_g} \quad (4.3)$$

where U_g is the superficial gas velocity, ρ_g is the gas density, μ_g is the gas viscosity, and ℓ is a characteristic dimension, such as the particle diameter, cyclone diameter, etc., depending on the application. The Reynolds number can be interpreted as the ratio of inertial force to viscous force.

The quantity of great practical interest is the drag force exerted by the fluid on the body. We consider the case of a small spherical particle moving in a continuous viscous gas. For a $10 \mu\text{m}$ particle in air at 300°K with the relative velocity of 1.0 m/s , we obtain

$$N_{Re} = 6.7 \times 10^{-2}$$

Thus, we conclude that the inertial effects are negligible. We then have a simplified Navier-Stokes equation (Batchelor, 1967) from which the drag about the sphere is computed as

$$F_D = 3 \pi d_p \mu_g U_g \quad (4.4)$$

where F_D is the drag force parallel to U_g . This is commonly known as Stokes' law for the resistance to a moving sphere.

Tchen's Model

The equation of motion of a particle in a fluid flow has been successively derived and improved by Basset (1888), Boussinesq (1903), and Oseen (1927). The model is that of a slowly translating sphere in a fluid at rest. Tchen (1947) extended this idea to the case in which a spherical particle is translated in a body of fluid with a variable velocity field. Tchen's equation can be written as

$$\begin{aligned}
 & \overset{A}{\frac{\pi}{6} d_p^3 \rho_p \frac{dU_p}{dt}} = \overset{B}{3 \pi \mu_g d_p (U_g - U_p)} + \overset{C}{\frac{\pi}{6} d_p^3 \rho_g \frac{dU_g}{dt}} \\
 & + \overset{D}{\left[\frac{1}{2} \frac{\pi}{6} d_p^3 \rho_g \frac{dU_g}{dt} - \frac{dU_p}{dt} \right]} + \overset{E}{\left[\frac{3}{2} d_p^2 (\pi \rho_g \mu_g)^{1/2} \int_{t_0}^t dt' \frac{\frac{dU_g}{dt'} - \frac{dU_p}{dt'}}{(\tau - \tau')^{1/2}} \right]} \\
 & + F_e
 \end{aligned} \tag{4.5}$$

where t_0 is the starting time, U is the velocity, ρ is the density, d is the diameter, and F_e is the external force acting on the particle; the index p refers to the particle quantity, and the index g refers to the gas flow quantity. Each term is described as follows:

- A: The force required to accelerate the particle.
- B: The viscous Stokes' drag.
- C: The pressure gradient caused by the acceleration of the fluid surrounding the particle.
- D: The force required to accelerate the apparent mass of the particle relative to the ambient fluid.
- E: The "Basset term," which is the effect of the deviation in flow pattern from steady state due to the fluctuating component of the velocity field.

F: The external force such as gravity, magnetic force, buoyancy, and electrostatic force.

This is, therefore, the equation derived from the first principle. The assumptions in Tchen's model are as follows:

- (1) Flow is described by continuum theory, i.e., $N_{Kn} \ll 1$.
- (2) There are no particle-to-particle interactions.
- (3) The Reynolds number based on the relative velocity of the fluid and the particle is small and thus the particle motion relative to the ambient fluid follows Stokes' law, i.e., $N_{Re} \ll 1$.

Stokes' Number

In hot-gas cleanup devices, the density of the solid particle is many times greater than the density of the ambient air, i.e.,

$$\frac{\rho_g}{\rho_p} \ll 1 \quad . \quad (4.6)$$

Then Equation (4.5) can be greatly simplified to give a first-order ordinary differential equation of the form

$$\frac{dU_p}{dt} = \frac{18 \mu_g}{d_p^2 \rho_p} (U_g - U_p) + F_e \quad , \quad (4.7)$$

where terms C, D, and E in Equation (4.5) have been neglected by the assumption made in Equation (4.6).

Equation (4.7) has the solution of the form

$$U_p = \exp(-t/\tau) U_o + \left(U_g - \frac{F_e}{m_p} \right) \left[1 - \exp(-t/\tau) \right] \quad , \quad (4.8)$$

where τ is the characteristic response time, m_p is the mass of the particle, and U_o is the initial velocity of the particle. The response time is an indication of the particle's ability to respond to the changing fluid velocity and is defined as

$$\tau \equiv \frac{C_s \rho_p d_p^2}{18 \mu_g} \quad , \quad (4.9)$$

where we have included the Cunningham slip correction factor, C_s , to include the first-order corrections to the continuum approach. Since the Cunningham slip correction factor is a function of the mean free path of the gas and the particle dimension, it becomes important at high temperatures (large λ) and for small particles ($\leq 1 \mu\text{m}$).

We now define Stokes' number as

$$N_{St} = \frac{U_g \tau}{D}, \quad (4.10)$$

where D is the characteristic dimension of the domain in which the particle travels. The Stokes' number is an indication of how closely the particle follows the mean streamline of the fluid flow. At small Stokes' numbers, the particle and fluid streamlines correspond closely. At large Stokes' numbers, the particle continues to move essentially unaffected until any possible impaction.

4.2 Parameters of Particle Collection

The fundamental theories discussed above may be used to develop the key parameters for particle collection efficiency analysis. The parameters that play an important role in describing the collection mechanisms are identified as

- (1) Inertial impaction parameter
- (2) Diffusion parameter
- (3) Gravitational settling parameter
- (4) Interception parameter
- (5) Electrical migration parameter

In this section, these parameters are analyzed and the effects of temperature and pressure are examined.

Inertial Impaction Parameter

One of the most important mechanisms for the collection of particles larger than $1 \mu\text{m}$ in diameter is inertial impaction. Inertial impaction takes advantage of the inertial driving force of the particle (i.e., large

Stokes' numbers) and impinges the particle on a target. The relative effect of inertial impaction for a variety of particles and flow conditions may be characterized by the inertial impaction parameter, defined as

$$K_{II} = \frac{C_s \rho_p d_p^2 U_g}{9 \mu_g d_c} \quad , \quad (4.11)$$

where d_c is the collector diameter.

It can be easily seen in the parameter K_{II} that the effects of temperature and pressure are reflected in the ratio C_s/μ_g . Figure 4-1 (Parker et al., 1977) illustrates the change in the inertial impaction parameter at various HTHP conditions for different particle sizes.

For large particles, a large Stokes' number implies that inertial impaction increases. However, for large particles for which the Cunningham slip correction factor is near unity, the Stokes' number depends mainly on the gas viscosity for a given fluid velocity. Thus, inertial impaction decreases for high temperatures as μ_g (for air) increases with temperature. For small particles for which the slip correction factor becomes important, the Stokes' number increases with temperature due to the increasing mean free path, but high pressure tends to nullify this effect.

Diffusion Parameter

Small particles under constant molecular bombardment can exhibit significant Brownian motion. The rate of diffusion is characterized by the diffusion parameter, K_D , which is a measure of mean-flow convection to molecular thermal diffusion:

$$K_D = \frac{C_s k T}{3 \pi \mu_g d_p U_g d_c} \quad , \quad (4.12)$$

where k is Boltzmann's constant and T is the absolute temperature. This particular grouping of the variables bares a special importance in fluid mechanics. The inverse of K_D is often referred to as the Peclet number, N_{Pe} . Smaller particles undergo higher rates of diffusion. High temperature increases the molecular thermal diffusion for all particle sizes. High pressure decreases the thermal diffusion because of its effect on the mean free path.

Figure 4-1. Effect of Temperature and Pressure on C_s/μ_g Ratio

Gravitational Settling Parameter

Using Stokes' law with a steady settling velocity, we can define the gravitational settling parameter as

$$K_S = \frac{C_s d_p^2 (\rho_p - \rho_g) g}{18 \mu_g U_g} , \quad (4.13)$$

where g is the acceleration due to gravity. In general, even at relatively high pressures (≈ 50 atm), the gas density is much smaller than the particle density and may be neglected in K_S . We thus have

$$K_S = \frac{C_s \rho_p d_p^2 g}{18 \mu_g U_g} . \quad (4.14)$$

Thus, the effects of temperature and pressure are reflected in the ratio C_s/μ_g , which is illustrated in Figure 4-1.

Interception Parameter

The classical approach in analyzing the particle-to-particle interaction is to assume that a particle will stick to the collector within its radius from the fiber/granule element. We may then characterize the interception mechanism by the direct interception parameter defined as

$$K_I = \frac{d_p}{d_c} , \quad (4.15)$$

where d_c is the collector diameter. The interception parameter is not a function of temperature and pressure.

Electrical Migration Parameter

The migration of a particle in an electric field is proportional to the field strength, the charge on the particle, and Stokes' law. We characterize this mechanism by the Coulombic force parameter K_C defined as

$$K_C = \frac{C_s q_p E}{3 \pi \mu_g d_p U_g} , \quad (4.16)$$

where q_p is the particle charge and E is the electric field strength. The effects of temperature and pressure on the gas flow are given from the ratio C_s/μ_g . The particle charge and electric field strength, however, are complex functions of temperature and pressure and cannot be generalized.

These parameters form the basis of describing the particle collection mechanisms. The results are summarized in Table 4-I.

4.3 Key Mechanisms for Pressure Loss

In PFBC hot-gas cleanup, the particulate control equipment is used to treat the working fluid in the power-generating cycle. To design, troubleshoot, or evaluate such a system, it is important to understand the basic mechanisms for the pressure loss of the working fluid.

The degree of complexity of the gas treatment system can be quite varied. In some cases, relatively small ducts are used, but the ducts may have a large number of fittings, branches, and various obstructions such as a bed of granules or a series of bag filters.

Moreover, the fluid resistance of these systems by analytical calculation is difficult, and it is often estimated empirically. In practice, the pressure loss of a working fluid is mainly caused by the following physical mechanisms:

- (1) Expansion loss, ΔP_E
- (2) Friction loss, ΔP_F
- (3) Swirling loss, ΔP_S
- (4) Blockage loss, ΔP_R
- (5) Dust cake resistivity loss, ΔP_D

In this section, these physical mechanisms are discussed.

Expansion Loss

When a working fluid is fed into a large chamber, such as in an electrostatic precipitator, it experiences some loss of energy. Because of the adverse pressure gradient, it is difficult to avoid flow separation and the consequent energy loss and increase in turbulence. In this expansion process, the pressure loss is related as

$$\Delta P_E = (1 - \eta_D) \left[\frac{1}{2} \rho_g U_g^2 (1 - A_1^2/A_2^2) \right] , \quad (4.17)$$

Table 4-I
Effects of Temperature and Pressure
on Collection Parameters

	<u>High T</u>	<u>High P</u>	<u>Small d_p</u>	<u>Small d_c</u>
K_{II}	↓	↓	↓	↑
K_D	↓	↓	↑	↑
K_S	↓	↓	↓	—
K_L	—	—	↓	↑
K_C	Complex Function of T and P			—

where A_1/A_2 is the channel-to-chamber area ratio and η_D is the diffuser efficiency.

Friction Loss

When a fluid travels along a wall, a boundary layer forms. The nature of the boundary layer can be used to account for the existence of drag. Because of the presence of the boundary layer, the effective shape of the wall modifies the flow outside the boundary layer. This changes the streamline of the fluid flow, which in turn gives rise to a modified distribution of pressure. When integrated over the whole surface, the pressure forces will give a resultant force in the stream direction which is known as form drag. Furthermore, the velocity gradient, $\partial u/\partial y$, at the boundary is large. Thus, there is viscous stress on the surface of the body and, when integrated over the whole surface, this tangential stress (as distinct from the normal stress which is pressure) gives a force in the stream direction which is known as skin friction.

These two elements make up for the so-called friction loss to the total head of the working fluid. The friction loss depends on the surface roughness, the shape of the wall, the viscosity of the gas, and the flow velocity. Laminar flow, in general, exhibits lower drag than a turbulent flow over a given wall section.

Swirling Loss

When a working fluid is introduced to a circular chamber with a high tangential velocity, such as in a cyclone, the fluid swirls as it travels along the axial direction. In this flow situation, the streamline of the fluid is modified by the wall. The fluid loses its energy in the process and, thus, a loss in stagnation pressure occurs. This is known as swirling loss.

Swirling loss depends on the tangential velocity, the wall curvature, the turbulence level, and the viscosity. This effect also applies when a flow must negotiate the corners of pipes, ducts, and conduits.

Blockage Loss

Blockage in a duct or a chamber causes flow separation and a sudden increase in the form drag because of the drastic change in the streamline shape of the flow. The flow must negotiate the blockage with a considerable loss in total energy. This is known as blockage loss.

Dust Cake Resistivity Loss

The total pressure drop across a fixed granular bed or a ceramic filter consists of the pressure drop due to the media and due to the accumulated dust cake.

Being consistent with conventional filtration theory, the pressure drop across filter media is proportional to the face velocity and filter thickness (or bed depth). The pressure drop across the dust cake is similarly proportional to the face velocity and the dust thickness. For a constant dust concentration, the cake thickness increases linearly with time until possible blowback takes place. Thus, the total pressure drop is given as

$$\Delta P_D = C_1 U_f + C_2 U_f^2 \sigma (t - t_o) , \quad (4.18)$$

where C_1 and C_2 are empirical constants, σ is the dust concentration, t is time, U_f is the face velocity, $U_f \sigma t$ is the rate at which the dust cake thickness increases, and t_o is the time the bags were last cleaned. Dust cake resistivity loss is therefore a function of U_f , U_f^2 , time, and the cleanup cycle period. The eventual blockage predicted by Equation (4.18) can be prevented by periodically cleaning the bags by backflushing air through them and dislodging the dust cake into a dust collection hopper.

These parameters form the basis of describing the pressure loss of the system. When designing a particulate control system with minimum power wastage, these parameters must be considered carefully.

4.4 Generic Device Technology

Certain physical and chemical factors are common within each of the device types, and the obtainable efficiency as a function of particle size is different for each separator system. Thus, to evaluate the various gas

purification systems, the interaction and interdependence of the basic physical processes must be identified.

The particulate collection mechanisms discussed in the previous section may be used to characterize various HTHP purification systems and to categorize them into five generic technologies as shown in Table 4-II.

Without further discussion of the separation effects of each single apparatus, it can be stated that the existing analytical models can be evaluated by identifying (1) the appropriate physics of particulate collection and (2) the correct assessment of the temperature and pressure effects for each generic device technology as discussed before.

Table 4-II
Five Generic Device Technologies

<u>Generic Device Technology</u>	<u>Examples of Specific Device</u>	<u>Predominant Collection Mechanism</u>
1. Inertial separators	Cyclone, cyclocentrifuge.	Inertial impaction.
2. Surface filters	Ceramic cross-flow filter, ceramic bag filter.	Inertial impaction, interception, diffusiophoresis.
3. Granular bed filters	Granular bed filter, dry plate scrubber.	Inertial impaction, direct interception, diffusiophoresis, sedimentation.
4. Electrostatic precipitators	Electrostatic precipitator.	Coulombic force.
5. Hybrid devices	Electrocyclone, electro- static granular bed, electrostatic filter, etc.	Conventional collec- tion mechanisms augmented by Coulombic force.
	Magnetically stabilized granular bed filter.	Conventional collec- tion mechanisms.*
	Acoustic agglomerator.	Particle agglomera- tion induced by acoustic pressure waves.**

* Experiments performed by EXXON Research and Engineering Company indicate that the collection efficiency is independent of the magnetic property of the particulate (Golan and Matulevicius, 1981). Thus, only the conventional collection mechanisms are of importance.

**Acoustic agglomeration technology is not a collection mechanism by itself. It is a gas conditioner by which the particulate-laden gas stream is treated to reduce the fine particle population.

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5. Alkali Effects in Hot-Gas Cleanup

Available data and information on the chemical properties of fly ash and alkali species were collected, and thermochemical equilibrium and kinetics data bases relevant to these systems were surveyed. The principal findings are summarized below.

Thermochemical data for complex aluminosilicate minerals and glasses similar in composition to PFBC fly ash have been compiled by the National Bureau of Standards (Schumm, 1981) and the U.S. Geological Survey (Robie et al., 1978). Their reports, along with a major revision of the JANAF Tables, became available in November 1981. These data bases all share common formats.

The purpose of the first-quarter alkali studies was to determine whether a modeling scheme for alkali release, its absorption (gettering), and its corrosive behavior in a PFBC gas stream should be included in the ASPEN code. This question was approached through a survey of published literature and available recent information from vendors, both under DOE Program Research and Development Announcement (PRDA) Contracts. Results were analyzed to resolve possible conflicts in the information and to identify a common thread within the existing data that could be used to model these three phenomena.

Kinetics data on alkali release and absorption (gettering) in a PFBC context have been compiled by a number of R&D groups, and the results have been quite consistent. In Westinghouse studies, the kinetics of alkali vaporization from dolomite ($d[\]/dt$), when recalculated in Arrhenius form, give the temperature function (Mulik, 1981)

$$\begin{aligned} \ln \left(\frac{d[\text{Na}]/dt}{[\text{Na}]_i} \right) &= 9.75 - 16158/T \text{ (s}^{-1}\text{)} \\ \ln \left(\frac{d[\text{K}]/dt}{[\text{K}]_i} \right) &= 11.26 - 19345/T \text{ (s}^{-1}\text{)} \end{aligned} \tag{5.1}$$

where $[\text{Na}]_i$ and $[\text{K}]_i$ are the respective initial concentrations in the dolomite, and T is the absolute temperature.

Alkali absorption kinetics have been measured for a variety of aluminosilicate getters, ash fractions, and sorbents for other hot-gas cleanup

systems such as granular bed filters. In general, sodium vapor absorption by getters is kinetically fast,

$$-\frac{d[\text{Na}]_{\text{gas}}}{dt} \approx (37 \text{ s}^{-1}) [\text{Na}]_{\text{gas}}, \text{ at } 1073^\circ\text{K} (1472^\circ\text{F}) \quad (5.2)$$

so that sorbents appear to be capacity limited, not rate limited. Only in the Westinghouse study, however, were Na and K both monitored in the same experiments. The volatilization of K appeared to coincide with and accompany the absorption of Na species from the hot gas. Thus, alkali removal in hot-gas particulate cleanup systems such as granular bed filters is not significant.

Measurements of alkali levels in PFBC environments all show concentrations (approximately 1 ppm) that are 1 to 3 orders of magnitude below equilibrium vapor pressures for NaCl(g) or NaOH(g) but are considerably higher than PRDA turbine maximum specifications.

Several investigators who we visited have suggested that the PRDA alkali specifications might be unrealistically stringent. One study by Combustion Power Company (CPC) of IN738 corrosion rates in PFBC using wood as fuel in a bed of aluminosilicate getters gives a temperature function which, recalculated in Arrhenius form, is (Cooper and Moresco, 1981)

$$\ln \left[\frac{d}{dt} (\text{penetration depth}) \right] = -13.2 - 12980/T \text{ (m/s)} \quad (5.3)$$

CPC observations that corrosion depends solely on temperature suggest that the kinetics are saturated with respect to alkali loading. The simplest kinetic description (for modeling purposes) of this phenomenon is a Michaelis-Menton mechanism in which

If this mechanism is kinetically accurate, the rate equation for corrosion becomes

$$\frac{1}{V} = \frac{1}{V_{\text{max}}} + \frac{(K_m)}{V_{\text{max}}} \frac{1}{[\text{Na/K}_x(\text{gas})]} \quad (5.5)$$

where V is the corrosion rate; V_{max} is the maximum corrosion rate at saturation; $[\text{Na/K}_x(\text{gas})]$ is the gas stream alkali loading; $K_x(\text{gas})$ is the molar

fraction of potassium in salt compound; and K_m is the Michaelis constant on which the turbine specifications should be based (equal to the alkali loading at which the corrosion rate is one-half V_{max}). For coal-based PFBC tests on IN738, V_{max} can be determined using Equation (5.3), and CPC data.

Table 5-I summarizes the principal findings obtained from each vendor. In evaluating these findings, we proceeded to investigate critically both the actual measurements obtained and the assumptions inherent in their studies; Table 5-II summarizes these. Our purpose here was to see whether alternative explanations would be consistent with the data. In evaluating this information, note that the following measurement parameters are not consistent among one another:

- o Particle-bound alkali
- o Water-extractible or acid-extractible alkali
- o Vapor phase alkali
- o Dissociable alkali
- o Weight changes
- o Alkali entrainment

All of these measurements are meaningful, but they have (or at least they can have) different meanings.

Conceptually, alkali removal and gettering can be visualized as a reversible two-step physical-chemical process: (1) condensation and desorption of alkali on particle surfaces and (2) dissolution of the alkali surface deposit into the particle interior. In a PFBC hot-gas cleanup environment, it may be necessary to consider kinetics as well as thermodynamics in modeling the alkali behavior.

The results obtained by General Electric (GE) were methodologically limited to process (1) above. By contrast, experiments carried out at Argonne National Laboratory (ANL), Air Pollution Technology, Inc. (APT), and Combustion Power Company (CPC), as well as Coal Utilization Research Laboratories (CURL) in Leatherhead, England, all treated only step (2) above. Historically, workers at Westinghouse demonstrated the reverse process of step (1) and have since examined the chemistry and physics of gettering. Given this variety of

Table 5-I
Primary Vendor Findings

- o GE--Alkali desorption occurs solely by Anderson adsorption, with diffusional kinetics. (Intermediates between Langmuir BET adsorption conditions.) (Boericke et al., 1981a)
- o Westinghouse--Alkali release occurs primarily from dolomite. (Mulik, 1981)
 - Aluminosilicate getters are a mixed blessing.
 - Gettering is capacity limited, not rate limited.
 - Alkali absorption in a granular bed filter is modeled as an irreversible process.
- o CPC--Alkali removal is unimportant in granular bed filters. (Cooper and Moresco, 1981)
- o APT--Some getters are effective for alkali removal, particularly diatomaceous earth. (Calvert et al., 1981)
- o ANL--Alkali removal by bauxite is chemical-reaction limited and is efficient. (Lee and Johnson, 1981)
- o CURL--Alkali removal by granular bed filters is extremely efficient. Alkalies are not a problem in PFBC. (Pillai and Wood, 1980)

Table 5-II
Quantities Actually Measured by
Various Alkali Behavior Investigators

- o GE--Alkali deposits on solids by thermogravimetric analysis (TGA). Sodium only or potassium only.
- o Westinghouse--dissociable alkalis "gas phase, direct combustion probe" in dolomite evaporation studies. Simultaneous sodium and potassium determination in gettering experiments for solid phases. TGA weight gain for getters, exclusive of gas phase composition.
- o CPC--Particle-bound alkali only, corrosion rates for IN738 coupons.
- o APT--Captured alkali by direct measurement on solids; coupled with initial alkali loading during entrainment.
- o ANL--Captured alkali by direct measurement on solids; coupled with alkali loading upon initial entrainment.
- o CURL--Particulate alkalis only; except, when GE-SRI probe was used, those alkalis rapidly condensable in an ice bath were also measured.

approaches, it is hardly surprising that the findings are quite different. The best clues to the kinetics of alkali release appear to come from the Westinghouse work. Recalculation of their data shows

- o Alkali release from dolomite is kinetically a first-order process, and it can be modeled as such.
- o Recalculation of the extent and rate of alkali vaporization into Arrhenius form gives a picture of the process consistent with that obtained from thermodynamics, and it can be modeled as such.

In these studies, the system would be described kinetically according to the Michaelis-Menton equation familiar to biochemists, and direct measurements of sodium gettering kinetics (ANL, APT) show first-order rate contents on the order of 37 reciprocal seconds. Getters are therefore solubility capacity limited, not rate limited. GE and Westinghouse reached the same conclusions. Only Westinghouse, however, measured sodium and potassium together in the same experiment. Westinghouse found potassium desorption during sodium gettering. This argues against the simple physical absorption-desorption process. The ratio of sodium to potassium measured in the vapor stream for these experiments was quite variable, ranging from $\text{Na/K} = 1$ to $\text{Na/K} = 15$.

When the alkali absorption-desorption is considered as a two-step reversible process, many of the disagreements seem to vanish.

Corrosion tests in a PFBC environment reported by Combustion Power Company showed relatively low corrosion rates [Equation (5.3)].

Other alloys (less resistant) give corrosion rates up to 10^4 times faster (Bornstein et al., 1968). CPC rates are not a function of sodium. The reason for this is not known at this time, although one might speculate that the corrosion process here also proceeds according to the Michaelis-Menton equation. Primarily, numbers indicate that this is the case.

We then conclude that:

- (1) Westinghouse is the only company that measured both Na absorption and K desorption. Na-K displacements on a getter granule may thus create an additional alkali problem in PFBC environments.

- (2) The presence of alkalis in PFBC environments can be modeled, at least as far as their production from dolomite is concerned.
- (3) The removal of alkalis by hot-gas cleanup systems can be modeled because the forms of the equations (Michaelis-Menton) are available. The required parameters, however, are not all available.
- (4) It remains an open question whether hot-gas cleanup devices contemplated for inclusion as ASPEN modules can do anything significant for alkali control.

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6. Model Evaluation

Based on the literature search and actual facility visitations, much information on the advanced gas cleanup systems was obtained and examined during the last half of 1981 for the purpose of making a selection of an initial set of gas cleanup devices for development as cleanup system simulation models. A detailed set of evaluation criteria was established to aid in that selection. In this section, we present the general approach used in model evaluation, the evaluation criteria, and the device-by-device analysis. Of the eleven devices listed in Section 2, only the cycloentrifuge from Mechanical Technology, Inc. (MTI) was not evaluated. MTI declined to participate in the evaluation effort.

Experimental Data

To evaluate the performance of hot-gas cleanup devices, we examined the available test data. The factors of importance are the experimental conditions, the measurement techniques used, and the particulate sources.

Devices have been tested in three broad categories of experimental conditions: (1) low temperature, low pressure; (2) high temperature, low pressure; and (3) high temperature, high pressure. In addition to temperature and pressure, several other basic conditions are important: for model verification, the internal or superficial velocity is essential; and for determining the collection capabilities of the device, particle loading and particle size distributions are needed.

Measurements of temperature and pressure are easily accomplished with standard instrumentation. Superficial velocity is readily calculated from the velocity of the inlet or outlet gas stream, which is measured by standard means. The measurement of particle loadings and distributions, however, is much more complex. The change of particle loading across the cleanup system is the means for calculating device efficiency. The change of particle distribution is used to calculate the fractional efficiency, that is, the efficiency as a function of particle size. The fractional efficiency is particularly important in comparing devices. The submicron particulate is almost universally the most difficult to collect. The ability to determine how well a device works in the 0.1 to 1 μm range is a valuable gauge of device quality.

Four different methods for measuring particulates have been used in device testing. These are (1) an absolute filter, (2) a cascade impactor, (3) an opacity meter, and (4) a Coulter counter. The particle loading can be measured with an absolute filter sample in the inlet and compared with an absolute filter sample in the outlet for a direct measure of the mass efficiency of a device. For the ceramic bag and ceramic cross-flow filters, in fact, this is generally what has been done. More information can be obtained, however, by using an impactor. The impactor fractionates the collected particulates into several size ranges from which a distribution can be determined. By measuring the size distribution before and after a device, the fractional efficiency can be evaluated. This is a great help in both diagnostics and model verification. Optical methods, such as the opacity meter, have the advantage of being nonintrusive, but they are strictly qualitative as they are highly responsive to changes in the particle size distribution. Generally, optical methods are good for diagnostics and laboratory procedures but are not applicable when quantitative results are needed. The Coulter counter is an alternative method to the impactor where a collected sample is analyzed in vitro. The Coulter counter operates by resuspending an ash sample in a dilute electrolytic solution and measuring particles electronically as they pass through an aperture. Micron- and submicron-size particles which are important to the operation of cleanup devices are below the practical use range and generally not accurately represented by the counter. There is often considerable disagreement between Coulter counter results and results from an in situ method such as the impactor. For quality determination in our evaluation, the highest marks were given to impactor data.

The use of a specific particulate source used for testing a hot-gas device will influence or prejudice the results. The stickiness of the ash will affect device behavior. The efficiencies measured for the device will vary tremendously with the amount of fine particulate in the inlet distribution. Resuspended ash, for example, will give higher efficiencies than an actual PFBC source with a larger fraction in the submicron range. (Most resuspended ash has at most 1 to 2% of its mass in the submicron range.) The effect of this distribution size change on efficiency can be profound. For example, the Acurex tests at Westinghouse with resuspended ash achieved efficiencies of

99.5 to 99.8%, while tests on the slipstream of the Exxon miniplant under similar temperature and pressure conditions ranged from 96 to 99.5%. Other tests on resuspended ash have gone as high as 99.98% efficiency (Shackleton et al., 1981).

Because tests with resuspended ash are somewhat easier and less expensive to perform, much testing has been done with that source. However, care should be taken to note the source in comparing results with data taken on an actual PFBC source.

To simulate the stickiness of the hot ash in cold-flow testing, Burgess pigment #10 was used (Boericke et al., 1981b). This pigment consists of fine, sticky paint particles. Other materials such as limestone (finely ground) were also used. In laboratory tests, a monodisperse aerosol such as polystyrene latex spheres and methylene-blue was used to measure the efficiency at a specific particle size, e.g., 0.5 μm or 1.0 μm (Brown, 1981).

To assess the quality of the test data generated by a particular device, several factors were examined. These include the duration of tests, the repeatability of data, and the presence or lack of performance degradation with time. Data were ranked with respect to experimental conditions, measurement techniques, and the kind of particulate source. The highest ratings were accorded high-temperature and pressure testing on an actual PFBC with measurement of collection efficiency by impactor.

Cost Information

Cost forecasting of an engineering product under development is a complex problem. Methods vary from industry to industry and are affected by unforeseen elements in today's fast-paced technology development and volatile economy. The outcome is therefore subject to many interrelating variables. Of all the companies contacted in 1981 under the present contract, only Research Cottrell and General Electric had done extensive research in this area. CPC has developed some cost data, based on their experience in the development and sales of conventional granular bed filters and electroscrubbers, and Westinghouse has obtained a cost data base by subcontracting the cost analysis to Burns and Roe economic analysts.

Research Cottrell published a preprint summarizing some projected costs for various combinations of HTHP cyclones, granular bed filters, and electrostatic precipitators in a 600-megawatt power plant using eight PFBC boilers and four gas turbines. Although many of the capital cost elements were listed in the text of their report (shop fabrication, field fabrication, interconnecting duct work, etc.) together with labor rate assumptions, the actual numbers employed in their studies were not available from the text itself. The output from their study was very brief and lacked detailed information.

The General Electric cost estimates for their electrostatic granular bed filter went into considerably more detail. Very detailed cost estimates were spelled out on a system-by-system basis in terms of materials, labor, and the addition of a cyclone to the cleanup train. Materials were broken down according to the filter media handling system, the ash disposal system, the granular bed filter itself, and the vent support system. Labor estimates included the erection of the vessel bag house, the compressor, and the screw feeder. Altogether, their cost estimates provided us with some 19 pages of detailed costing information for their device.

Cost information for the electrocyclone, the magnetically stabilized granular bed filter, the cross-flow filter, the dry plate scrubber, the acoustic agglomerator, and the others we evaluated has been very difficult to come by. Costing estimates for these devices were generated by estimating the amount of materials and the approximate construction costs for standard items (such as pressure vessels) to obtain a rough idea of the capital costs for these units. Other factors influencing costs, such as system pressure drop, particle removal efficiency, and auxiliary power requirements, were somewhat easier to estimate.

Evaluation Criteria

To aid in the task of final model selection, a list of evaluation criteria was established and is shown in Table 6-I. The selection criteria were based on a grading system with a scale of 0 to 100 points and divided into three major categories: performance, cost, and potential for future commercialization.

Table 6-I
Model Selection Criteria (100 points)

Criterion I.	<u>Performance</u>	<u>60 points</u>
	A. Basic Model	<u>35 points</u>
	1. Availability	20
	2. Quality	15
	B. Supporting Data	<u>25 points</u>
	1. Availability	15
	2. Quality	10
Criterion II.	<u>Cost Projection</u>	<u>25 points</u>
	A. Availability	<u>15 points</u>
	B. Quality	<u>10 points</u>
Criterion III.	<u>Commercialization</u>	<u>15 points</u>
	A. State of Development	<u>9 points</u>
	1. Process tested	6
	2. Laboratory device	3
	B. Marketability	<u>6 points</u>
	1. Low operating cost	3
	2. Low capital investment	3

The fundamental basis for accepting any model was the capability of the model to provide computationally efficient mathematical methods most compatible with ASPEN. Thus, the 100-point grading system was skewed heavily toward the state of the analytical models, with 60 points for technical performance and 25 points for cost projection. The remaining 15 points were allocated for grading the potential commercialization of the device.

The points for the technical performance of a candidate model were divided between the availability and quality of the basic model and its supporting data. An edge was given to devices that had models well based in first principles as described in Section 4. The ideal model would be based on sound physical principles with sufficient supporting data to validate the model.

In evaluating the performance models, emphasis was placed on the following characteristics: (1) the analytical models were given higher scores than the empirical models, since the empirical models were limited by the specific experimental data and were not, in general, accurate for different geometries and different scale-up and operation conditions; (2) both analytical models and empirical models were examined for their assessment on the temperature and pressure effects as discussed in Section 4; and (3) device models which include the particle collection efficiency and pressure drop computations were given higher scores than either the collection efficiency or the pressure drop model alone.

In evaluation of the supporting data, scores were given incrementally on the amount and level of testing achieved as well as the perceived quality of the results. Emphasis was placed on the closeness of the approach to PFBC conditions. The highest scores were achieved by long-duration test programs with consistently high efficiency, measured as the fractional efficiency by inlet/outlet impactor results run at HTHP conditions on an operating PFBC source. Points were deducted for high-temperature only runs, for inconsistent data, for lack of fractional efficiency measurements, and for using resuspended fly ash as a source. Cold-flow only data were rated only as high as 50% of a full score on the same set of quality criteria, such as run duration, consistent results, impactor fractionalization, etc.

Cost projection models were rare, since most of the devices considered for HTHP gas cleanup are relatively new. Devices that have a good item-by-item data base scored well as technical models.

Regardless of how well the technical and cost models worked, the device itself may not have the potential for commercialization; thus, this category was added to address this concern. The commercialization factor was not weighted as heavily because, through the maturation process of technology, installation and operating costs tend to drop. The proposed grading system addresses the concern of the marketplace, but is not driven by it.

The following section describes our detailed examination of individual devices and gives a summary of the numerical evaluation of each device.

6.1 Electrocyclone by General Electric

The cyclone has been in use for particle control in various industries for nearly a century. It is perhaps the best known and least expensive device for particle collection. Consequently, the theory and practice of cyclone operation has been extensively documented.

Historically, cyclones were characterized by a cut size which was computed by equating the centrifugal force on a particle to the drag on the particle due to the radial gas velocity (Lapple, 1951; Stairmand, 1951). This theory is easily understood and widely employed.

A recent study by Leith and Licht (1972) proposed an improved model which included the inherently turbulent nature of flows inside cyclones. Consequently, their model incorporates empirical turbulent back-mixing of the suspended particles and provides reasonable agreement with data. However, the model assumes uniform mixing of the particulate-laden gas and ignores the interchange of particles between the upflowing and downflowing sections of the cyclone.

The three-region cyclone model developed by GE is a valuable advance in the analytical representation of cyclones (General Electric, 1981). The model includes separate representations for (1) the entrance region, (2) the downflow (or annular) regions, and (3) the upflow (or core) region. The model computes the inertial impaction due to the swirling flow according to Stokes' law. The effect of temperature is represented by change in viscosity. The effect of pressure is not represented. The model provides for variations in the particle concentration within the cyclone. Within each region, turbulence is assumed to promote mixing of the suspended particles. Incorporation of this mixing concept into the three-region model allows a more sophisticated analytical representation for the collection efficiency of the cyclone.

The three-region model has been extended to include the effects of space charge and applied electric field to model the electrocyclone concept. This effort was carried out under a related contract from the DOE Division of Environmental and Safety Engineering (Contract DE-AC02-79EV10143).

GE also developed the so-called five-region model to accommodate an air shield feature in their original design. However, due to poor efficiencies obtained by this design, the air shield has been removed from their concept, and the five-region model has also been dropped.

In summary, the three-region model has the following features:

- o The efficiency model is available and well documented.
- o The model does not compute the pressure drop.
- o The temperature effect is represented by change in viscosity.
- o The model does not include the analytical representation of pressure effects.

To this point, cold-flow data has been generated with a 46-cm (18-inch) cyclone that measured up well against theory. The electrocyclone was re-designed to a lower efficiency, higher throughput version. Fractional efficiency data have been taken showing a large effect with an applied electric field for low-velocity cases versus a relatively small effect at high velocities. There appears to be an unexpected falling off of efficiency, however, for larger particles (5 to 10 μm range). Hot-flow tests are planned with a 30-cm (12-inch) model at PFBC conditions.

Although neither the cost forecast model nor the cost data base were available from GE, some cost projections were achieved for this device. As manufacturing and operation costs for conventional cyclones are well developed, additional costs for the pressure vessel, duct work, and electrodes were added to supplement the preliminary cost data base.

The evaluation criteria scores for the GE electrocyclone are given in Table 6-II.

Table 6-II
Evaluation Results for Electrocyclone by General Electric

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	15
	2. Quality	6
	B. Supporting Data	
	1. Availability	5
	2. Quality	5
Criterion II.	<u>Cost Projection</u>	
	A. Availability	8
	B. Quality	6
Criterion III.	<u>Commercialization</u>	
	A. State of Development	4
	(process tested)	
	B. Marketability	
	1. Low operating cost	3
	2. Low capital investment	3
	<u>Total Score</u>	55

6.2 Ceramic Bag Filter by Acurex

The conventional approach in modeling bag filters has been to assume that the collection efficiency is constant and is very high over the range of particle sizes. Acurex is no exception in this approach; their model assumes the collection efficiency to be 99.95% or better for all particle sizes.

The pressure drop is empirically described as follows:

$$\Delta P = \zeta_d \mu_g \sigma U_f^2 t + P_o, \quad (6.1)$$

where

ΔP	=	total pressure drop across filter, in. H ₂ O
ζ_d	=	empirical dust cake resistivity, nondimensional
μ_g	=	gas viscosity, centipoise
σ	=	particle loading, gr/ACF
U_f	=	face velocity, air-to-cloth ratio, ft/min
t	=	cleaning interval, min
P_o	=	baseline pressure drop, in. H ₂ O

Thus, both collection efficiency and pressure drop models are empirical and subject to interpretation of experimental data.

Testing of ceramic bags at Acurex is well advanced. Bags have been tested for chemical reactivity at the CURL facility (Leatherhead, England). Durability testing has been conducted in the laboratory. Individual bags were HTHP tested at the Exxon miniplant, and a five-bag system was tested for 100 hours at the Westinghouse laboratory facility at PFBC conditions on resuspended ash. Problems with bag blow-through and bag cleaning were encountered but appear to have been resolved. Almost no fractional efficiency measurements have been made, and the efficiency data have been derived from a final total-flow filter.

Because of the experience the Acurex group has accumulated over the past few years in the development of ceramic bag filters in subpilot- to pilot-scale ventures, there exists a well-developed cost data base for this device. This data base includes not only the item-by-item price list, but also engineering and operation man-hour estimates. Such a data base strengthens the cost analysis algorithm already found in ASPEN.

The evaluation criteria scores for the ceramic bag filter are given in Table 6-III.

Table 6-III
Evaluation Results for Ceramic Bag Filter by Acurex

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	10
	2. Quality	5
	B. Supporting Data	
	1. Availability	12
	2. Quality	8
Criterion II.	<u>Cost Projection</u>	
	A. Availability	10
	B. Quality	8
Criterion III.	<u>Commercialization</u>	
	A. State of Development	7
	(process tested)	
	B. Marketability	
	1. Low operating cost	2
	2. Low capital investment	2
	<u>Total Score</u>	64

6.3 Ceramic Cross-Flow Filter by Westinghouse

The case for this device is somewhat unusual in that practically no basic model existed for predicting the performance of the device. The model assumed the ceramic cross-flow filter to be similar to a bag filter which assumed a very high particle collection efficiency. This assumption was unproven. However, the empirical pressure drop model had the potential to become available in the future as more experimental data are analyzed. Very little cost modeling input was available because the ceramic cross-flow filter was still at the bench-testing stage. What information existed was limited to the connection between pressure drop and operating cost, which provided a reasonable size estimate of the pressure vessel and required compressors. Because of the early development stage, the quality of any hardware cost projections was regarded as marginal. Finally, criterion III (commercialization) reflects the early state of development. Our evaluation about its marketability was predicted on the assumption that the low pressure drop and projected small device size would eventually result in a low operating cost and a reasonably low capital investment.

The cross-flow filters had been tested at ambient conditions and at high-temperature, ambient-pressure conditions. Initial plugging problems were solved by reducing the length of the dirty side channel from 15.2 to 7.6 cm (6 to 3 inches) per cube of filter monolith.

Impactor data had not been published and efficiency numbers were given only in the form of estimates. The evaluation criteria scores are given in Table 6-IV.

6.4 Moving Granular Bed Filter by Combustion Power Company

CPC approached the model development in three ways:

- (1) One-dimensional empirical process model (Guillory, 1980).
- (2) One-dimensional analytical process model (Wigton 1978, Goren 1979).
- (3) Three-dimensional axisymmetric analysis, which is currently under development.

Table 6-IV
Evaluation Results for Ceramic Cross-Flow Filter by Westinghouse

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	5
	2. Quality	5
	B. Supporting Data	
	1. Availability	10
	2. Quality	5
Criterion II.	<u>Cost Projection</u>	
	A. Availability	4
	B. Quality	5
Criterion III.	<u>Commercialization</u>	
	A. State of Development	5
	(process tested)	
	B. Marketability	
	1. Low operating cost	2
	2. Low capital investment	2
	<u>Total Score</u>	43

Guillory had correlated the collection efficiency and pressure drop in his empirical model by the following simple relation on the basis of linear regression analysis of his test data:

$$\eta = 1 - A_o \left(\frac{\Delta P}{U_g} - \frac{\sigma}{\dot{m}} \right)^{A_1} \quad (6.2)$$

where ΔP = pressure drop across the granular bed
 U_g = superficial velocity
 σ = inlet dust concentration
 $\dot{m}$ = media flow rate
 A_o, A_1 = empirical constants

Wigton and Goren had expressed the collection efficiency in the analytical model as an addition of four terms representing the collection by impaction, interception, diffusion, and sedimentation mechanisms, viz,

$$\eta = f(K_{II}, K_I, K_D, K_S) \quad (6.3)$$

where η is the fractional efficiency due to a single grain. To obtain the total (system) efficiency, we integrate this equation over the cross-sectional area and bed depth. From these equations, CPC could size the granular bed according to a desired collection efficiency. This is an extremely useful design tool.

The three-dimensional axisymmetric collection simulator was under development. This model included the effect of granule flow, particle-to-particle capture, and many other additional features of fundamental collection mechanisms.

Substantial amounts of testing had been done on the CPC granular bed filter at atmospheric conditions. Development tests had culminated in a 1000-hour duration test during which efficiencies ranged between 98 and 99%. Measurements were made with an Anderson Impactor raised to an intermediate temperature of 700 to 775°K (800 to 930°F). CPC used the APT high-temperature impactors. The consistency of the data over the duration of testing (1000 hr) made these high-quality results.

With their successful marketing of electroscrubbers and their experience over the past several years in subpilot-scale development of granular bed filters, CPC had accumulated a good cost data base. However, the development had been directed toward AFBC gas cleanup, and thus the quality of the data base as applied to PFBC conditions remained unproven.

The evaluation criteria scores for CPC's moving granular bed filter are given in Table 6-V.

6.5 Fixed Granular Bed Filter by Westinghouse

The basic models are available and the quality of these models is at least consistent with good engineering practice. The connection of these models to first principles was somewhat tenuous and the Westinghouse computer code was lacking in documentation. However, some supporting data were given for pressure drop, particularly with respect to the eductor design options.

The pressure drop model was consistent with good engineering practice. The model is empirical and can be described as

$$\Delta P_{\text{sec}} = \Delta P_o + \Delta P_{B,B} + \Delta P_D, \quad (6.4)$$

where ΔP_{sec} = section pressure drop
 ΔP_o = baseline section pressure drop, including the pressure drop due to flow contraction to the tubular gas exit and the pressure drop across the distributor plate (This is the baseline pressure drop without the granules for clean gas operation.)
 $\Delta P_{B,B}$ = pressure drop due to blowback flushing (time dependent)
 ΔP_D = pressure drop due to dust cake (time dependent)

Table 6-V
Evaluation Results for Moving Granular Bed Filter by
Combustion Power Company

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	17
	2. Quality	8 only AFBC tests
	B. Supporting Data	
	1. Availability	12
	2. Quality	7
Criterion II.	<u>Cost Projection</u>	
	A. Availability	10
	B. Quality	7
Criterion III.	<u>Commercialization</u>	
	A. State of Development	7
	(process tested)	
	B. Marketability	
	1. Low operating cost	2
	2. Low capital investment	2
	<u>Total Score</u>	72

However, at the time of the evaluation the pressure drop data base for this model had been obtained under cold flow conditions. Based on the general experimental approach, we expect the quality of that data to be reasonably high. Some cost data were available from Burns and Roe, consultants on the Westinghouse contract, though that information was not by itself suitable for inclusion into a model.

With regard to its potential for commercialization, the device had been tested on the sub-pilot scale, where it had run into its share of problems. Westinghouse reported that many of the initial problems had been overcome through design changes. We were not convinced that the device could be operated at a reasonably low cost, or that the capital investment required for the materials handling would be low.

Prior test data on bed operation at laboratory cold-flow conditions showed good potential with 99+% efficiencies with limestone ($<10\text{ }\mu\text{m}$) and fly-ash tests. Initial tests with the Ducon/Exxon designs proved problematical; screen plugging prevented useful data acquisition. With a revised Exxon design, collection efficiencies ranged between 31 and 97%. Rapid ΔP buildup, loss of bed material, and decreased performance with time were noted. These results indicate unresolved problems with the bed design, the cleaning cycle, and matching the flows in the two different elements. A redesigned system was currently ready for testing at the Waltz Mill hot-gas cleanup test facility.

The evaluation criteria scores for the fixed granular bed filter are given in Table 6-VI.

6.6 Electrostatic Granular Bed Filter by General Electric

During the evaluation period, an extensive modeling effort was underway at General Electric for the electrostatic granular bed filter. The effort included electrostatic charging, particle-particle interaction, granular flow, and simulated jetting effects to account for the presence of the nearest neighbors.

Table 6-VI
Evaluation Results of Fixed Granular Bed Filter
by Westinghouse

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	15
	2. Quality	7
	B. Supporting Data	
	1. Availability	10
	2. Quality	5
Criterion II.	<u>Cost Projection</u>	
	A. Availability	10
	B. Quality	7
Criterion III.	<u>Commercialization</u>	
	A. State of Development	6
	(process tested)	
	B. Marketability	
	1. Low operating cost	2
	2. Low capital investment	2
	<u>Total Score</u>	64

At that time it was believed that the model might be too complex and overly large for implementation on ASPEN. At the time of the evaluation, it required \$600/run of computing. It was impractical to consider the code for ASPEN.

A much simpler code was available from Grace et al. (1980) at Combustion Power Company. Electrostatic granular beds had already been developed for use under conventional operating conditions by CPC and MIT. Additional details and historical applications of CPC's electroscrubber filter were available from Parquet (1982). Electropacked (EP) and electrofluidized (EF) beds under investigation at MIT had been reported (see Johnson and Melcher, 1975; Zahedi and Melcher, 1976; Alexander and Melcher, 1977; Melcher and Presser, 1981; and Melcher and Rhoads, 1981).

The penetration of dust through the electrostatic granular bed is expressed as

$$\eta_{\text{tot}} = 1 - (1-\eta)_{\text{elec}} (1-\eta)_{\text{mech}}, \quad (6.5)$$

where η_{tot} = total system efficiency
 η_{mech} = collection efficiency for the baseline granular bed without the electrostatic enhancement
 η_{elec} = collection efficiency due to electrostatic augmentation

Grace et al. (1980) give the following expression for η_{elec} :

$$\eta_{\text{elec}} = 1 - \exp \left[\frac{-2(1-\epsilon) z E_0 q_p C_s}{\pi \mu_g d_c d_p U_g} \right], \quad (6.6)$$

where ϵ = bed porosity fraction
 z = bed thickness
 E_0 = average electric field strength
 q_p = particle charge

This approach could be conveniently incorporated into the granular bed model discussed in Section 7.3 for implementation in the ASPEN code. It was recommended that the mechanical efficiency be modeled by the method described

in Section 7.3 and, in addition, an electrostatic module be installed in the ASPEN model which could be activated when the electrostatic option is exercised.

Considerable effort had been made in developing an extensive cost data base in terms of materials and labor for manufacture and installation. GE's cost estimate provided some 19 pages of detailed document, which undoubtedly strengthens the data base stored in ASPEN.

The evaluation criteria scores for the electrostatic granular bed filter are given in Table 6-VII.

6.7 Dry Plate Scrubber by Air Pollution Technology, Inc.

The APT dry plate scrubber was unique in that the collector granule bed is thin (≈ 2 cm) and the high-velocity particle-laden jets locally fluidize the bed. This local fluidization is termed the "spouting jet" and helps to minimize the pressure drop. The spouting jets cause particles to be separated from the mainstream by inertial impaction and direct interception, but diffusion and sedimentation are not significant collection mechanisms because of the thin bed depth and short transient time (i.e., transient time = bed depth/jet velocity). APT claimed that the existing granular bed models would not adequately describe the spouting jet design. Thus, mathematical models were not available for the new design.

APT had a substantial amount of operational data at cold-flow conditions. Bed efficiency on a per stage basis had been measured between 20 and 80%, depending on configuration. The highest efficiencies were measured with charged particulates entering the bed in addition to an applied field across the bed. The lowest efficiencies were obtained with the nonelectrostatic configuration (no field, particles uncharged). APT planned to group several uncharged bed stages with a final electrostatic stage to achieve the required efficiencies. Alkali removal might be attempted by having at least one stage of getter material. Measurements of efficiency were made by impactor and appeared to be of very good quality throughout. Hot-flow work had been done for alkali removal and a laboratory AFBC had been constructed for high-temperature collection testing. Hot results would be available soon.

Table 6-VII
Evaluation Results for Electrostatic Granular Bed Filter
by General Electric

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	9
	2. Quality	8
	B. Supporting Data	
	1. Availability	7
	2. Quality	7
Criterion II.	<u>Cost Projection</u>	
	A. Availability	12
	B. Quality	8
Criterion III.	<u>Commercialization</u>	
	A. State of Development	4
	(process tested)	
	B. Marketability	
	1. Low operating cost	2
	2. Low capital investment	2
	<u>Total Score</u>	59

Although the device was still in laboratory-scale development, a preliminary item list for a subpilot AFBC study had been considered. The list would include the cost of the pressure vessel, granules, pipe work, and the perforated or slotted plates. However, the list was still premature and the quality was expected to be low.

The evaluation criteria scores for the dry plate scrubber are given in Table 6-VIII.

6.8 Magnetically Stabilized Granular Bed Filter by Exxon

Performance models, which related the magnetic field strength to bed stability and collection efficiency, were not available. However, it had been observed experimentally that the magnetic property of the ash material had no effect on collection efficiency. Thus, only the conventional collection mechanisms (i.e., inertial impaction, diffusion, interception, and sedimentation) were important.

Collection testing had been primarily at cold-flow conditions, with hot testing of the magnetic materials only. The data showed considerable improvement on resuspended fly-ash collection with the application of a 105 oersted field, but a continuing degradation with time was noted. Tests were of approximately 1-hour duration. A reduction of the pressure drop and an increase in the blowout velocity of the bed were noted. Efficiencies in the 98 to 99% range were claimed but had not been documented over a long duration with a well-established particle distribution. The facility for PFBC condition testing had been built and the results of the initial HTHP shakedown testing would be available soon.

Considerable time and effort had been spent on the study of the magnetic properties of the bed material. Exxon held the significant part of information concerning the manufacture of the bed material as being proprietary, and thus cost projection was unavailable.

The evaluation criteria scores for the magnetically stabilized granular bed filter are given in Table 6-IX.

Table 6-VIII
Evaluation Results for Dry Plate Scrubber
by Air Pollution Technology, Inc.

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	7
	2. Quality	7
	B. Supporting Data	
	1. Availability	8
	2. Quality	8
Criterion II.	<u>Cost Projection</u>	
	A. Availability	4
	B. Quality	4
Criterion III.	<u>Commercialization</u>	
	A. State of Development (process tested)	6
	B. Marketability	
	1. Low operating cost	2
	2. Low capital investment	2
	<u>Total Score</u>	48

Table 6-IX
Evaluation Results for Magnetically Stabilized
Granular Bed Filter by Exxon

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	0
	2. Quality	0
	B. Supporting Data	
	1. Availability	7
	2. Quality	7
Criterion II.	<u>Cost Projection</u>	
	A. Availability	4
	B. Quality	4
Criterion III.	<u>Commercialization</u>	
	A. State of Development	4
	(process tested)	
	B. Marketability	
	1. Low operating cost	2
	2. Low capital investment	1
	<u>Total Score</u>	29

6.9 Electrostatic Precipitator by Research Cottrell

In industrial applications at conventional temperatures and pressures, the use of ESPs had been popular in fine-particulate control ($d_p < 10 \mu\text{m}$). These devices exhibited high collection efficiencies and low pressure drops under conventional operating conditions. However, there was a fundamental problem associated with ESP design and operation under HTHP conditions, namely that of maintaining a stable corona discharge over a wide range of operating variables, such as pressure, temperature, electrode geometry, gas composition, and electrode polarity in the presence of fly ash and alkali vapor.

The approach RC has taken to analytically investigate the performance of HTHP ESPs was twofold:

- (1) Use the standard one-dimensional ESP model (the modified Deutsch model) to compute the collection efficiency; give the stable corona, particle charging rate, and the electric field strength.
- (2) Empirically determine the range of stable corona and the charge rate. These empirical data are, in turn, used in the Deutsch equation to perform the η calculation.

Thus, much of RC's work has been directed towards application of the Deutsch model with substantial empirical understanding of the corona stability problem under HTHP conditions.

The modified Deutsch equation may be written as

$$\eta = 1 - \exp \left[\left(\kappa E_o E_p / 3 \mu_g \right) (A/Q) (\alpha C_s d_p) \right]^2, \quad (6.7)$$

where	η	= collection efficiency, fraction
	κ	= permittivity constant of space, $\text{C}^2/\text{N-m}^2$
	E_o	= charging field strength, V/m
	E_p	= collection field strength, V/m
	μ_g	= gas viscosity, N-s/m^2
	A	= cross-sectional area of the device, m^2
	Q	= volumetric flow rate, m^3/s
	α	= Cochet equation for field charging, nondimensional
	C_s	= Cunningham slip correction factor, nondimensional
	d_p	= particle diameter, m

Here, the terms in the exponent can be grouped to give the Coulombic force parameter, K_C . The Cochet equation for the field charging was written as

$$\alpha = (1 + N_{Kn})^2 + 2/(1 + N_{Kn}) , \quad (6.8)$$

where N_{Kn} is the Knudsen number. The corona starting voltage is approximated by Peek's formula as

$$\frac{E_c}{\delta'} = C_1 + \frac{C_2}{(r_0 \delta')^{1/2}} , \quad (6.9)$$

where E_c = corona starting field strength, V/cm
 δ' = relative density conventionally taken with respect to 1 atm (1.0×10^5 N/m²) and at 298° K, nondimensional
 C_1, C_2 = empirical constants
 r_0 = wire radius, cm

The behavior of corona stability, however, had been found to depend on the electrode polarity. Thomas and Wong (1958) and Feldman et al. (1977) reported that the positive corona characteristics depend chiefly on gas density, whereas negative corona characteristics are a function of the gas temperature as well as gas density and electrode geometry. Thus, the analysis of ESP operation in HTHP conditions must take into account many design variables, particularly at temperatures over 1000°K (1500°F) where the effects of photoionization are expected to be important.

Furthermore, the pressure drop assumption for the conventional ESP operation, i.e., $\Delta P < 2$ in. H₂O, remained unproven at HTHP conditions. Unfortunately, no analytical models were available to estimate the pressure loss of the system.

Some corona data at HTHP conditions had been generated, but no collection testing had been done. A substantial pool of prior data existed, however, for operation and collection at similar conditions, which appeared to substantiate RC's expectation of a 90 to 98% collection efficiency. In their new design, they claim to have solved the mechanical thermal stress problems and insulator conductance problems previously encountered.

As one of the leading manufacturers of conventional electrostatic precipitators, Research Cottrell had accumulated an extensive cost data base including manufacturing cost, engineering and operating costs, and estimated years of service. Furthermore, the ESP design for HTHP applications was currently under construction, and the added experience would undoubtedly strengthen the cost data base for future use in ASPEN.

The evaluation criteria scores for RC's electrostatic precipitator are given in Table 6-X.

6.10 Acoustic Agglomerator by State University of New York at Buffalo

Acoustic agglomerators are not collection devices. Their sole purpose is to precondition the process stream to produce a particle distribution that can be more easily handled by other cleanup systems.

The concept deals with various aspects of acoustic-aerosol interaction and acoustic propagation and attenuation in aerosols (Scittm, 1976). Theories have been developed for both small-amplitude and large-amplitude sound waves.

The effectiveness of an acoustic agglomerator is evaluated on the basis of the agglomeration time, which could be written as (Mednikov, 1965)

$$t_o = \frac{3}{K_a} \ln \left[\left(\frac{\rho_p}{\rho_{po}} \right)^{1/3} \left(\frac{a}{a_o} \right) \right], \quad (6.10)$$

where a is the particle radius and ρ_p is the particle density. The subscript "o" represents the conditions at time $t = 0$, and K_a is the empirical agglomeration constant, which is proportional to the square root of the acoustic intensity (Shaw and Wegrzyn, 1977).

The empirical data required to compute the acoustic agglomeration constant, K_a , are limited to cold-flow data only. The state of development has therefore been limited to laboratory models.

Testing of the acoustic agglomerator had been conducted at cold-flow conditions primarily by optical methods. Although the data appeared to be of good quality, much of the particle size information was inferred from opacity readings and lacked verification by a physical sizing method such as an impactor. A moderate temperature (650°K) and ambient pressure test facility was planned.

Table 6-X
Evaluation Results for Electrostatic Precipitator
by Research Cottrell

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	15
	2. Quality	6
	B. Supporting Data	
	1. Availability	8
	2. Quality	5
Criterion II.	<u>Cost Projection</u>	
	A. Availability	8
	B. Quality	7
Criterion III.	<u>Commercialization</u>	
	A. State of Development (process tested)	4
	B. Marketability	
	1. Low operating cost	2
	2. Low capital investment	2
	<u>Total Score</u>	57

The device was still in laboratory-scale development, and consequently no cost projection models or data bases were available.

The evaluation criteria scores for the acoustic agglomerator are given in Table 6-XI.

6.11 Phase I, Task 1 Recommendation

The final objective of the first quarter of work for this contract was to recommend specific hot-gas cleanup devices to use for further mathematical modeling and computer code development. The ASPEN code was selected as the executive code for modeling the performance of the hot-gas cleanup devices. Ten existing devices were evaluated according to three general criteria: performance, cost, and potential for future commercialization. The 100-point evaluation scale used was skewed heavily toward the state of the analytical models, with 60 points for technical performance, 25 points for cost projection, and 15 points for commercialization factors. Those devices with existing analytical models were rated considerably higher than those without. The evaluation results for all ten devices are given in Table 6-XII. Based on the results of the grading system, the following six devices were recommended for installation into the ASPEN executive code:

- Moving Granular Bed Filter by CPC
- Ceramic Bag Filter by Acurex
- Fixed Granular Bed Filter by Westinghouse
- Electrostatic Granular Bed Filter by GE
- Electrostatic Precipitator by Research Cottrell
- Electrocyclone by GE

METC subsequently approved this recommendation and five of the devices were modeled for the ASPEN code. The sixth device, the Westinghouse granular bed filter, was not completed.

Table 6-XI
Evaluation Results for Acoustic Agglomerator
by State University of New York

Criterion I.	<u>Performance</u>	<u>points</u>
	A. Basic Model	
	1. Availability	8
	2. Quality	5
	B. Supporting Data	
	1. Availability	5
	2. Quality	5
Criterion II.	<u>Cost Projection</u>	
	A. Availability	0
	B. Quality	0
Criterion III.	<u>Commercialization</u>	
	A. State of Development (process tested)	4
	B. Marketability	
	1. Low operating cost	1
	2. Low capital investment	1
	<u>Total Score</u>	29

Table 6-XII
Evaluation Results Summary

	<u>Performance</u>		<u>Cost</u> <u>Projection</u>	<u>Commercialization</u>	<u>Total</u>
	Basic Model	Supporting Data			
CPC	25	19	17	11	72
Acurex	15	20	18	11	64
Westinghouse GBF	22	15	17	10	64
GE-EGB	17	14	20	8	59
Research Cottrell	21	13	15	8	57
GE (3-Region)	21	10	14	10	55
APT	14	16	8	10	48
Westinghouse Ceramic	10	15	9	9	43
EXXON	0	14	8	7	29
S.U.N.Y.	13	10	0	6	29

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7. Description of the Five Recommended Models

In this section, we describe five of the six advanced HTHP gas cleanup devices that were selected for installation into the ASPEN executive code. The sixth device, the Westinghouse granular bed filter, was not completed. The ASPEN interface software and pressure drop model was coded but had not been debugged and validated by the end of the contract period.

7.1 Moving Granular Bed Filter

Granular bed filters have been used in several industrial installations: for the removal of particulate matter from the effluents from waste boilers in the forest products industry; in the cleaning of gases from clinker coolers in the cement industry; and in the cleaning of lime kiln effluents. Experimental studies have also been conducted for a wide variety of potential applications, including the removal of radioactive particles from the air in ventilation systems, the cleaning of waste gases from open hearth furnaces, and the simultaneous desulfurization and particulate removal from the gas from coal-fired boilers.

Granular bed filters consist of closely packed granules of river stone, alumina, or other materials that are capable of withstanding the environment of the flue gases to be cleaned. The sizes of the granules vary with the application, the usual sizes ranging from around 1.5 to 8 mm in diameter.

The physical arrangements of granular bed filters include a wide range of designs that permit the gas to flow through the filter medium and provide a means for removing the collected dust. Two basic methods of cleaning the filter medium are used. Fixed bed filters generally employ back flushing or mechanical agitation to dislodge the collected particulate matter and permit it to fall into a hopper for removal.

In moving bed filters, such as that shown in Figure 7-1, the filter medium and the collected dust are removed from the filter and the dust separated from the medium externally. The filter medium is cleaned in a disengagement chamber, where the dust and the granules are separated. The granules are reintroduced to the moving bed, and the dust falls into an ash hopper.

Figure 7-1. Sketch of CPC Moving Granular Bed Filter

The performance model of the moving granular bed filter takes into account the following particle collection mechanisms:

- (1) Inertial impaction
- (2) Direct interception
- (3) Gravitational sedimentation
- (4) Random diffusion

The efficiency of particle collection by granular bed filters is influenced by the temperature and pressure of gases as they alter the collection mechanisms. Inertial impaction depends upon the particle separation from a gas stream which is diverted around a collecting body. The efficiency of collection by impaction is a function of the particle mass, the particle velocity, the resistance of the gas to particle motion, and the frontal diameter of the collecting body. The resistance of the gas to particle motion increases with increasing temperature for relatively large particles due to the increase in gas viscosity. Increases in pressure would have a similar influence on viscosity; hence, a reduction in collection efficiency of large particles would be expected with an increase in pressure. Collection of small particles, on the other hand, would be enhanced by increasing temperature, as a result of the increased Brownian motion.

Steady-state equations for a one-dimensional moving granular bed have been developed by CPC. The method takes into account the dominant parameters of particle collection as identified above. These equations contain three terms that relate collection efficiency and pressure drop to operating conditions such as gas velocity, granule media characteristics, particle size, and local loading of suspended and captured dust.

Mathematical Model

Steady-state equations for a one-dimensional moving granular bed filter have been developed by CPC (Goren, 1979). The model requires prior knowledge of the device configuration.

The model has been developed for (1) the cocurrent configuration, in which the collector granules in the moving bed and the particle-laden gas flow in the same direction, and (2) the countercurrent configuration, in which the

granule media and the gas stream flow in opposite directions. The equations are based on mass balance on the sum of suspended and captured particles,

$$\frac{d(U_n + U_G N_G N)}{dz} = 0 , \quad (7.1)$$

mass balance on suspended particles alone,

$$\frac{d(U_n)}{dz} = -J N_G + R N_G , \quad (7.2)$$

and momentum balance on the gas,

$$\frac{dp}{dz} = -F N_G . \quad (7.3)$$

The variables are summarized as

- z = filter depth, m
- U = superficial gas velocity, m/s
- $n(z)$ = local concentration of suspended particles at depth, $1/m^3$
- U_G = superficial grain velocity, m/s
- N_G = number of grains per unit volume of bed, $1/m^3$
- $N(z)$ = local number of dust particles adhering to a single grain at depth z , $1/m^3$
- $p(z)$ = gas pressure, N/m^2
- J = rate of deposit onto a single grain including deposition onto its adhering dust particles, fraction
- R = rate of reentrainment, fraction
- F = drag force on a single grain including its adhering dust particles, N .

For cocurrent flow of gas and granular media, the + sign is used, and for countercurrent flow, the - sign is used in Equation (7.1). Expressions for J , R and F are complex functions of the operating conditions and geometrical variables. The remaining terms are interpreted as follows:

- Term $J N_G$: Rate of particle deposition per unit volume of bed
- Term $R N_G$: Rate of particle reentrainment per unit volume of bed
- Term $F N_G$: Drag force per unit volume of bed

The particle capture occurs on both the grain surface and on adhering dust particles. This effect is written as,

$$J = J_G + N J_1 , \quad (7.4)$$

where J_G is the rate of particle capture due to clean single grains and J_1 is the rate of particle capture due to all adhering particles. The present model expresses J_G and J_1 in terms of single grain capture efficiency, η_G , and single dust particle capture efficiency, η_1 , defined by the following equations:

$$J_G = \pi a_G^2 U n \eta_G \quad (7.5)$$

$$J_1 = (4\pi a_p^2) (3A_G U a_p / 2a_G) n \eta_1 , \quad (7.6)$$

where

a_G = collector granule radius, m

a_p = particle radius, m

n = number concentration of dust particles, $1/m^3$

η_G and η_1 are empirical correction factors that account for the realistic aerodynamics of the particles, including bounce-off and the influence of neighboring particles.

The assumption made here is that the gas flow rate must be much greater than the granule flow rate in the moving bed, i.e.,

$$U \gg U_G .$$

Wigton (1978) and Goren (1979) introduced an expression for the collection efficiency due to a clean single grain, η_G . The model is an addition of four terms representing the collection by impaction, interception, diffusion, and sedimentation mechanisms, viz,

$$\eta_G = f \left(\epsilon, K_{II}, K_I, K_D, K_{GS} \right) . \quad (7.7)$$

This equation corresponds to the experimentally determined values of η_G for clean granular media for a wide range of gas velocities and particle diameters.

The terms are described as

ϵ = void fraction of the granular bed

$$K_{II} \text{ , Inertial Impaction Parameter} = \frac{C_S \rho_p d_p^2 U_G}{9 \mu_G d_G}$$

$$K_I \text{ , Interception Parameter} = \frac{d_p}{d_G}$$

$$K_D \text{ , Diffusion Parameter} = \frac{C_S K T}{3\pi \mu_G d_p U_G}$$

$$K_S \text{ , Gravitational Sedimentation} = \frac{C_S \rho_p d_p^2 g}{18 \mu_G U_G} ,$$

where C_S = Cunningham slip correction factor, nondimensional
 d_p = particle diameter, m
 ρ_p = particle density, kg/m³
 d_G = granule diameter, m
 K = Boltzmann's constant, J/°K
 T = absolute temperature, °K
 μ = gas viscosity at temperature and pressure, N-s/m²
 g = acceleration due to gravity, 9.8 m/s²

These parameters are summarized and discussed in detail in Goren (1979).

Equation (7.6) defines the particle capture efficiency due to a single adhering particle, η , in a manner analogous to the definition of η_G in Equation (7.5). However, the cross-section available for capture, $4\pi a_p^2$, is based on the dust particle size, and the local gas velocity, $3A_G U a_p / 2a_G$, is evaluated at a distance a_p away from the no-slip grain surface. The factor A_G is a hydrodynamic factor relating the velocity gradient near the granule surface to the void fraction of the bed. For low-speed flows, A_G can be computed from simple hydrodynamic models of granular media. For example, Happel's model predicts $A_G = 38$ for a bed of spheres with fraction solids $(1-\epsilon) = 0.6$. For Reynolds numbers encountered in typical granular bed operations, fluid inertia plays a significant role in determining A_G . In

the absence of detailed theoretical and experimental analyses, Goren (1979) recommends that A_G be scaled with Reynolds number:

$$A_G = 38 (1 + 0.0194 \text{ Re}) , \quad (7.8)$$

where Re is the Reynolds number based on granule diameter $= \frac{d_G \rho u}{\mu}$.

The drag force, F , in Equation (7.3) can be approximated in a similar way. The drag force acts on both granules and dust particles, thus we may write

$$F = F_G + F_1 \quad (7.9)$$

and evaluate F_G and F_1 as follows:

$$F_G = 6\pi\mu a_G U B_G \quad (7.10)$$

$$F_1 = (6\pi\mu a_p) \frac{3 A_G U a_p}{2 a_G} B_1 , \quad (7.11)$$

where $6\pi\mu a_p$ is the drag force that would be exerted on a single isolated grain according to Stokes' law, B_G is a hydrodynamic factor that accounts for influence of neighboring particles, and B_1 is a correction factor known from theory to be 1.7 for creeping flow past a sphere on a flat wall. For low-speed flow, B_G can be computed from simple hydrodynamic models. For example, Happel's model gives the value of 85 for creeping flow past packed granules as Re approaches zero. To account for fluid inertia, Goren (1979) scales B_G to the Ergun equation:

$$B_G = 85 (1 + 0.0194 \text{ Re}) . \quad (7.12)$$

To complete the mathematical model, the boundary conditions are formulated as follows:

for cocurrent flow

$$n(0) = n_{in} \quad \text{and} \quad N(0) = N_{in} \quad (7.13)$$

for countercurrent flow

$$n(0) = n_{in} \quad \text{and} \quad N(L) = N_{in} \quad (7.14)$$

where L is the total depth of the bed.

The above set of equations can be expressed more compactly in dimensionless terms.

$$\begin{aligned}
 d(n^* + N^*)/dz^* &= 0 \\
 dn^*/dz^* &= -n^*(1 + \zeta N^*) + \xi N^* \\
 dp^*/dz^* &= -(1 + \zeta \beta N^*) \\
 n^*(0) &= 1 \\
 N^*(0) &= N_{in}^* \text{ for cocurrent flow}
 \end{aligned}
 \tag{7.15}$$

or

$$N^*(L^*) = N_{in}^* \text{ for countercurrent flow}$$

The dimensionless terms appearing in these equations are defined as follows:

$$\begin{aligned}
 n^* &= n/n_{in}, & P &= n_{out}/n_{in} \\
 N^* &= U_G N_G / U_{in}, & N_{in}^* &= U_G N_{in} / U_{in} \\
 z^* &= z \pi a_G^2 \eta_G N_G, & L^* &= L \pi a_G^2 \eta_G N_G \\
 p^* &= p a_G \eta_G / 6 \mu U_B, & \Delta p^* &= \Delta p a_G \eta_G / 6 \mu U_B \\
 \zeta &= J_1 U_{in} / J_G U_G N_G = 6 A_G a_p^3 U_{in} / a_G^3 U_G N_G \\
 \xi &= k / \pi a_G^2 U_G \eta_G, & \beta &= F_1 J_G / F_G J_1 = B_1 \eta_G a_G / 4 \eta_1 B_G a_p
 \end{aligned}
 \tag{7.16}$$

By solving the above set of equations we may express the penetration fraction, P , and pressure drop, Δp^* , as functions of L_x^* , ζ , ξ , N_{in}^* and β . These are written as

$$\begin{aligned}
 P &= f_1(L_x^*, \zeta, \xi, N_{in}^*) \text{ and} \\
 \Delta p^* &= L_x^* + \beta f_2(L_x^*, \zeta, \xi, N_{in}^*).
 \end{aligned}$$

Explicit formulas relating these groups are given below.

For countercurrent flow define

$$\begin{aligned}
 A &= (1 - \zeta P + \zeta N_{in}^* - \xi) / 2\zeta \text{ and} \\
 B &= \xi(P - N_{in}^*) / \zeta.
 \end{aligned}$$

Then the solution to Equation (7.15) has one of the following three forms:

If $C^2 = A^2 - B > 0$, then

$$L_x^* = \frac{1}{2C\zeta} \left\{ \ln \left(\frac{1+A-C}{P+A-C} \right) - \ln \left(\frac{1+A+C}{P+A+C} \right) \right\}$$

$$(\Delta p^* - L_x^*)/\beta = \zeta[N_{in}^* - A - P] L_x^* + \frac{1}{2} \ln \left(\frac{1+A-C}{P+A-C} \right) + \frac{1}{2} \ln \left(\frac{1+A+C}{P+A+C} \right). \quad (7.17)$$

If $A^2 - B = 0$, then

$$L_x^* = \frac{1}{\zeta} \left(\frac{1}{P+A} - \frac{1}{1+A} \right)$$

$$(\Delta p^* - L_x^*)/\beta = \zeta[N_{in}^* - A - P] L_x^* + \ln \left(\frac{1+A}{P+A} \right). \quad (7.18)$$

If $-D^2 = A^2 - B < 0$, then

$$L_x^* = -\frac{1}{D\zeta} \arctan \left(\frac{D(1-P)}{(1+A)(P+A)+D^2} \right)$$

$$(\Delta p^* - L_x^*)/\beta = \zeta[N_{in}^* - A - P] L_x^* + \frac{1}{2} \ln \left(\frac{(1+A)^2 + D^2}{(P+A)^2 + D^2} \right). \quad (7.19)$$

For cocurrent flow define

$$A = (1 + \zeta + \zeta N_{in}^* + \xi)/2\zeta \quad \text{and}$$

$$B = \xi(1 + N_{in}^*)/\zeta.$$

Then the solution to Equation (7.15) has one of the following three forms:

If $C^2 = A^2 - B > 0$, then

$$L_x^* = \frac{1}{2C\zeta} \left\{ \ln \left(\frac{1-A+C}{P-A+C} \right) - \ln \left(\frac{1-A-C}{P-A-C} \right) \right\}$$

$$(\Delta p^* - L_x^*)/\beta = \zeta(1 + N_{in}^* - A)L_x^* + \frac{1}{2} \ln \left(\frac{1-A+C}{P-A+C} \right) + \frac{1}{2} \ln \left(\frac{1-A-C}{P-A-C} \right). \quad (7.20)$$

If $A^2 - B = 0$, then

$$L_x^* = \frac{1}{\zeta} \left(\frac{1}{P-A} - \frac{1}{1-A} \right)$$

$$(\Delta p^* - L_x^*)/\beta = \zeta(1 + N_{in}^* - A)L_x^* + \ln \left(\frac{1-A}{P-A} \right). \quad (7.21)$$

If $-D^2 = A^2 - B < 0$, then

$$L_x^* = \frac{1}{\zeta D} \arctan \left(\frac{D(P-1)}{(1-A)(P-A)+D^2} \right)$$

$$(\Delta p^* - L_x^*)/\beta = \zeta(1 + N_{in}^* - A)L_x^* + \frac{1}{2} \ln \left(\frac{(1-A)^2 + D^2}{(P-A)^2 + D^2} \right). \quad (7.22)$$

The above set of equations is very complex and the problem of obtaining a set of solutions is compounded by the several computation paths that occur during the calculation process. The first step in formulating an algorithm for solving the problem is to categorize the equations and their terms as shown in Table 7-I. In that table, an invariant equation is defined as one that produces a constant value once the particle size is chosen whereas a variant is an iterative loop equation that produces changing values until a final solution is obtained.

The iteration control parameter is the penetration parameter, P . Researchers at CPC have found that a bisection technique is sufficient to obtain a set of solutions. A final solution is reached for each particle size when L_x^* converges to L^* . At that point, it is a trivial calculation to obtain the pressure drop and collection efficiency by using the following equations.

$$\Delta p_i = 6\mu U B_G \Delta p_i^* / a_G n_G$$

$$\eta_i = 1 - P_i \quad (7.23)$$

These equations provide solutions for a given particle size (i) and are sufficient for the collection efficiency. At present, however, there is no basic physical model available to obtain a bed pressure drop for multiple particle sizes except by averaging the particulate pressure drops. The researchers at CPC indicate that the technique is adequate for particle sizes less than 10 μm but could be inaccurate when used to evaluate device performance for HTHP streams with larger particle sizes.

Table 7-I
CPC Granular Bed Model Equation Parameters

User-Defined Constants

a_G	N_G
L	n_{in}
U_G	N_G

ASPEN Inputs

p	U
T	μ

User Variable

a_p

Iterative Invariants

N_G
 B_G
 L^*
 A_G
 n_1
 B_1
 B_1
 β
 ζ
 ξ
 N_{in}^*

Iterative Variants

P iteration control parameters
 $A(P)$
 $B(P)$
 $C(P)$
 $D(P)$
 L^*
 $L_x^* (P, A, C \text{ or } D)$
 $\Delta p^* (P, A, C \text{ or } D, L_x^*)$

7.2 Ceramic Bag Filter

Fabric filters have been used for many years to remove dusts from industrial process gases, notably from cement kilns and metallurgical operations. They have also been installed on a number of small power boilers, but it is only within the past few years that they have been considered for removing fly ash from large electric utility boilers burning coal.

For high-temperature, high-pressure applications, porous metal or woven ceramic material is used to withstand the severe physical environment.

The ceramic bag filter, such as that shown in Figure 7-2, is usually made up of cylindrical tubes of the order of 20 cm in diameter by 3 to 6 m long. A number of these tubes are suspended vertically in a pressure vessel. The bags are cleaned either by mechanical shaking or by a reverse flow of air often referred to as back flush.

Ceramic bag filters collect particles by the mechanisms of impaction, interception, and diffusion. However, after a new ceramic bag collector has been in operation for several minutes, collection mechanisms are different from those of classical filter theory. The reason for this change is that, after an initial phase of collection, the cake of filtered material that builds up on the bag surface and the interstices of the bag provides most of the filtration until it is removed in the part of the filtering cycle in which the filter is cleaned and regenerated.

As the cake grows thicker with time, the pressure drop across the filter increases. When a preset pressure drop across a compartment containing ceramic bags is reached, or when a preset time has expired, the compartment is isolated from the gas flow for bag cleaning. After the dust is removed, the isolated compartment is again exposed to the gas flow. A collection process now occurs which is similar to that of initial cleaning by an unused bag. But now the interstices of the filter material are loaded with particulate matter that is not dislodged by cleaning. After a number of cleanings, the bags build up an interstitial accumulation to a constant level, at which point the bag is considered to be broken in.

Because of this residual dust loading, the filter cake is more quickly reestablished when filtration commences. The pressure drop-versus-time relationship for a freshly cleaned bag is nonlinear until the filter cake is reestablished; this often takes several minutes.

Figure 7-2. Sketch of Acurex Ceramic Bag and Pressure Vessel

Factors that can influence ceramic bag filter performance with fly ash include the fabric structure, air-to-cloth ratio, maximum pressure drop before cleaning, method of cleaning, cleaning frequency, intensity of cleaning, and flue gas temperature and humidity. These factors can be broken down broadly into two categories: factors that are basic to the design of a bag filter system for optimum performance and factors that relate to the behavior of the system as installed. During the design of a bag house, questions of bag material, method of cleaning, air-to-cloth ratio, operating temperature, and humidity need to be considered. Once the bag house is operating, cleaning rate, duration and intensity of cleaning, and other variables related to the operation of the collector are adjusted to keep pressure drop before the cleaning below the maximum allowable.

Mathematical Model

The conventional practice in modeling bag filters has been to assume a perfect filter with fractional penetration that is a function of particle size (Dennis and Wilder, 1975). The Acurex Bag Filter Model assumes a particle escape fraction of 0.005 or less (Shackleton et al. 1982). This assumption has been coded into the high-temperature, high-pressure ceramic bag filter. We do not agree with the assumption that particle collection efficiency is independent of particle size, but in lieu of any ceramic bag particle distribution test data to evaluate other models, such as conventional baghouse systems, we coded the Shackleton et al. assumption directly into the model. However, the computer model and code are structured so that the present single-statement equation for collection efficiency can be easily replaced by the proper set of statements when test data are gathered and a proper analytical model developed.

The pressure drop is empirically described as follows:

$$\Delta P = \zeta_d \mu_g \sigma U_f^2 t + P_o \quad , \quad (7.24)$$

where

ΔP	=	total pressure drop across filter, in. H_2O
ζ_d	=	empirical dust cake resistivity, nondimensional
μ_g	=	gas viscosity, centipoise
σ	=	particle loading, gr/ACF
U_f	=	face velocity, air-to-cloth ratio, ft/min
t	=	cleaning interval, min
P_o	=	baseline pressure drop, in. H_2O

The air-to-cloth ratio, U_f , is defined as the volumetric flow rate of the gas divided by the total filter surface of all bags except those that are being cleaned:

$$U_f = \frac{Q}{(N_{\text{cell}} - N_{\text{clean}}) A_{\text{bag}} N_{\text{bag}}} \quad (7.25)$$

where

Q	=	volumetric flow rate (m^3/s)
N_{cell}	=	number of cells in operation
N_{clean}	=	number of cells being cleaned
N_{bag}	=	number of bags per cell
A_{bag}	=	surface area of bag, m^2

In addition to the pressure drop across bag filters, the Acurex design requires auxiliary power to clean the bags at some time interval. This back flush system is device dependent and may involve a variety of mechanical systems. We give here but a general guideline for computing the auxiliary power requirement.

Mechanism: Cleaning with pulse jet

Pulse volume	=	3 times bag volume
Pulse interval	=	5 - 10 minutes
Pulse volumetric flow rate	=	(number of bags)(pulse volume)/(pulse interval)
Pulse ΔP	=	2 times operating pressure

We thus have

$$\text{Pulse Power Requirement} = \frac{6N B_v P}{t \eta_c} \quad (7.26)$$

where

N	=	number of bags
B _v	=	bag volume, m ³
P	=	operating pressure, N/m ²
t	=	pulse interval, min
η _c	=	compressor efficiency, fraction

As an example, for a 15-bag unit, 2.8 ft³/bag, 10 min cleaning interval, and 10 atm PFBC operating condition with 25% compressor efficiency, the auxiliary power requirement is that of a 65-horsepower compressor. Thus, both collection efficiency and pressure drop models are empirical and subject to interpretation of experimental data.

Testing of ceramic bags at Acurex is well advanced. Bags have been tested for chemical reactivity at the CURL facility (Leatherhead, England). Durability testing has been conducted in the laboratory. Individual bags were high-temperature, high-pressure tested at the Exxon miniplant, and a five-bag system was tested for 100 hours at the Westinghouse laboratory facility at PFBC conditions on resuspended ash. Problems with bag blow-through and bag cleaning were encountered but appear to have been resolved. Almost no fractional efficiency measurements have been made, with efficiency data being derived from a final total-flow filter.

7.3 Electrostatic Granular Bed Filter

A typical filter design concept for an electrostatic granular bed filter is shown in Figure 7-3. This filter design uses a moving bed of granules contained in an annular space by large retaining louvers. The gas is introduced in the center of the element where it is charged by a discharge electrode at a high electric potential. The particles in the gas flow are charged by intense bombardment of ions in this precharging chamber. The gas flow containing the charged particles then flows outward through the annular bed of moving granules where the particles are captured by impaction and electrostatic forces. This single-stage, inside-to-outside flow arrangement

Figure 7-3. Sketch of GE Electrostatic Granular Bed Filter

results in a lower face velocity at the outer diameter where the blowout may occur.

Collection of particulates in a packed bed of granular material by purely mechanical means has been studied by many investigators. Neglecting electrostatic effects (other than those which hold the particle to the collector after capture) for the purposes of this discussion, granular media collect aerosol particles by at least four mechanisms: inertial impaction, interception, diffusion, and sedimentation.

Collection by impaction occurs when the inertia of a particle causes it to depart from the gas streamline and collide with the collector. Theoretical models indicate that the Stokes' number is a satisfactory dimensionless parameter for these analyses.

Brownian diffusion caused by random motion of small particles being bombarded by gas molecules enhances the possibility of a particle being collected. Even though the particle may be generally following a gas streamline, random motion occasionally allows the particle to approach a collector surface and causes it to contact that surface and be captured.

From the physical description of these phenomena, it can be seen that inertial impaction most effectively operates on larger particles (with a significant mass/area ratio) whereas diffusion is more effective on small particles whose trajectory can be influenced by molecular motion.

In addition to the mechanical collection mechanisms, the electrostatic granular bed filter utilizes enhanced charging of dust particles.

Collection due to electrostatic forces relies on the particulate charge which occurs naturally from upstream processes. Frictional charging (or tribocharging) will occur in flow through pipes and cyclones and in high mixing environments. Collection occurs when the direction of travel of a charged particulate is changed by virtue of the electrostatic force, in opposition to the fluid flow field and viscous forces, and the particulate moves through the toward a collection surface. Although the force is directed perpendicular to equipotential surfaces, the particulate undergoes many changes in direction through a packed media bed, and has many opportunities to come in contact with the media. The electrostatic collection in the granular filter is similar to

an electrostatic precipitator with very narrow and winding walls and therefore increased effective surface area. Of course, the mechanical collection mechanisms of impaction and diffusion contribute significantly to the overall collection.

The electric field strength in the filter is a function primarily of the applied voltage and the geometry of the electrodes. Secondary effects include the presence of the media and the gas as highly resistive dielectrics. Thus, within the well-defined overall field, there are local changes in field strength and direction due to the media and, on a smaller scale, the particulate. Local gradients in field strength provide another small force on the particulate resulting in dielectrophoresis. The collection of particulates is therefore related to a number of factors. The actual collection is affected by one or more of the collection mechanisms: impaction, interception, diffusion, and electrostatic forces. The one or two mechanisms that dominate in an electrostatic granular bed filter will depend on the following physical parameters: particulate size, density, charge, collector size, and fluid velocity. In addition, reentrainment, the rubbing off and recapture of particulates from the media, affects the net collection efficiency.

Mathematical Model

Robinson (1971) and Melcher and Zahedi (1976) have attempted to model the electrostatic effects on particulates; however, there are few theoretical models describing the behavior of charged particles in a complex flow field. Besides solving for the fluid flow field, a detailed model must account for inertial effects and small variations in the electric field within the granular filter. The following simple model includes the major parameters affecting particulate collection, while assumptions are made regarding geometry and particulate loading.

It was shown previously that the total penetration is equal to the product of the penetrations due individual mechanisms. For a system in which both mechanical and electrical effects are important,

$$(1-\eta)_{\text{tot}} = (1-\eta)_{\text{elec}} (1-\eta)_{\text{mech}}$$

Considering the various collection mechanisms to represent η "stages" (each of efficiency η_i) which operate successively on the original dust load, the overall collection efficiency $\bar{\eta}$ can be written as:

$$(1-\bar{\eta}) = \prod_{i=1}^n (1-\eta_i) \quad (7.28)$$

or, considering the two mechanisms discussed above,

$$(1-\bar{\eta}) = \left[1 - \left(\frac{St}{\epsilon} \right) \right]^m \left[1 - \frac{4.36}{C} Pe^{-2/3} \right] \quad (7.29)$$

where m is an empirically determined exponent which depends upon the Stokes' number range.

In order to include the electrostatic effect within the system of equations, an expression is needed for the collection efficiency as a function of particle charge, average electric field, etc. An approach similar to the Deutsch equation (White, 1963) first developed for standard ESP's is used in which the force on a charged particle due to an electric field is equated with the viscous force when it reaches terminal velocity. Relating this migration velocity to the geometry of the collecting surface and flow field yields

$$\eta_{elec} = 1 - \exp \left(\frac{-1}{3\pi\mu} \frac{q_p C_S A_c E}{Q_f d_p} \right) \quad (7.30)$$

- E = average electric field strength, V/m
- A_c = collection area, m^2
- Q_f = volumetric gas flow rate, m^3/s
- C = Cunningham slip correction factor, dimensionless
- q_p = particle charge, C
- d_p = particle diameter, m
- μ = gas viscosity, N/m-s

Considering the geometry of a granular bed, A_c/Q_f is approximated as:

$$A_c/Q_f = 6(1-\epsilon)t/d_g U \quad (7.31)$$

U = gas superficial velocity, m/s
 ϵ = void fraction, fraction
t = bed thickness, m
 d_g = media granular diameter, m

Substituting, we get

$$\eta_{elec} = 1 - \exp \left[\frac{\frac{-2}{\pi\mu} (1-\epsilon)t E q_p C_s}{U d_g d_p} \right] \quad (7.32)$$

From the above it is noted that increased efficiency can be obtained by increasing electric field strength, particle charge, and filter thickness, or by decreasing: flow velocity and media granule diameter. An increase in temperature is seen to reduce efficiency through the increase in gas viscosity. According to the model, larger particles actually have increased efficiencies, as the particle charge is assumed proportional to the square of the diameter (surface area)(White, 1963).

The auxiliary power requirement is computed for the corona discharge rate at the wire electrode and the pneumatic media lift system for transporting the granular collectors. The Ergun's pressure drop model is also used for this device (Ergun, 1952).

7.4 Electrostatic Precipitator

A typical high-temperature, high-pressure ESP configuration, such as that illustrated in Figure 7-4, is made up of discharge electrodes, generally at a high negative potential, bordered by grounded collector pipe also serving as the pressure vessel. The particulate-laden flue gas enters the collecting area where the intense electric field established by the corona wire and collector pipe charges the particles. This charge acquisition is characterized (1) by negative ions in a negative corona field or positive ions in a positive corona field moving along field lines which terminate on the uncharged particle, or (2) by random collisions of particles and ions, or (3) by a

Figure 7-4. Sketch of Research Cottrell ESP

combination of both processes. The first process, known as field charging, is dominant for particles greater than 1 μm in diameter.

Submicron particles, on the other hand, are charged primarily by the second process, which is known as diffusion charging.

In ESPs, the incoming particles reach their saturation charge very quickly. As soon as any charge is acquired, they experience a Coulomb force that provides an acceleration in the direction of the electrostatic field.

Whereas the larger, more massive, particles drift steadily toward the wall, the fines respond quickly to the turbulent flow of the flue gas and are difficult to capture. The fluid dynamic forces tend to distribute the fines throughout the gas volume, with the result that a relatively small fraction of their time is spent near the collecting surfaces. However, because of the persistent Coulomb force, even the fines will eventually drift to the collector pipe and precipitate. After a sufficiently thick layer of material has been deposited, the agglomerated particles are dislodged from the pipe by mechanical rapping. They then fall into hoppers and are conveyed out of the precipitator facility.

Although irrigated precipitators have a potential of achieving even higher efficiencies than dry systems, the present model can be used only to simulate a dry wire-pipe precipitator.

Mathematical Model

The ESP geometry analyzed here is a coaxial wire-cylinder combination. The cylinder serves as the pressure vessel to contain the high-temperature gas with pressures up to perhaps 20 atmospheres. For this arrangement, the basic equation of electrostatic field can be written in a cylindrical coordinate system as

$$E = - \frac{\partial V}{\partial r} \quad (7.33)$$

$$\text{and} \quad \frac{1}{r} \frac{d}{dr} (rE) = \frac{\rho_i}{\epsilon_0} \quad (7.34)$$

It is assumed throughout these calculations that the passive electrode is at ground potential.

The current per unit length at conductor, j_ℓ (amp/m), can be expressed as

$$j_\ell = 2\pi r \rho_i b E . \quad (7.35)$$

Substituting Equation (7.35) into (7.34) and integrating, the electrostatic field is given by

$$E(r) = \left[\frac{j_\ell}{2\pi\epsilon_0 b} + \left(\frac{r_0}{r} \right)^2 \left(E_c^2 - \frac{j_\ell}{2\pi\epsilon_0 b} \right) \right]^{1/2} . \quad (7.36)$$

The corona current-voltage relation is found by integrating Equation (7.36) following Townsend (1914), and neglecting second-order terms:

$$\frac{V_o - V_c}{V_c} \ln \left(\frac{r_1}{r_o} \right) = (1 + \phi)^{1/2} - 1 - \ln \frac{1 + (1 + \phi)^{1/2}}{2} . \quad (7.37)$$

Equation (7.37) can be easily solved for ϕ by functional iteration. Then E_c , the corona-starting field intensity, can be computed by

$$E_c = m \delta \left(A_g + B_g \sqrt{\frac{\delta}{r_u}} \right) \quad (7.38)$$

For negative corona in air, Robinson (1971) recommends the values $A_g = 32.2 \times 10^5$ V/m and $B_g = 8.46 \times 10^4$ V/m^{1/2}.

Thus, the local electrostatic field intensity $E(r)$ is computed by Equation (7.36) with prior knowledge of E_c from Equation (7.38), and j_ℓ from Equation (7-35).

When gases laden with suspended particulate matter are passed through an electrostatic precipitator, the particles acquire an electrical charge of the same polarity as that of the discharge electrodes. Two distinct particle charging mechanisms are generally considered to be dominant: (1) bombardment of the particles by ions moving under the influence of the applied electric field, commonly known as field charging, and (2) attachment of ionic charges to the particles by ion diffusion in accordance with the laws of kinetic theory, commonly referred to as diffusion charging.

In the formulation by Research Cottrell (Feldman et al., 1977) both field charging and diffusion charging were considered. In general application of ESPs, particle charging due to ion diffusion becomes significant for micron- and submicron-size particles.

Field charging has been calculated by Pauthenier and Moreau-Hanot (1932). In their formulation, the following is assumed:

- (1) The particles are spherical.
- (2) Particle spacing is sufficiently great to render negligible any particle-to-particle interactions.
- (3) The mean free path of the ions is small when compared to the diameter of the particles.
- (4) The electrical field intensity, except in the immediate neighborhood of a particle, is uniform throughout the volume of the precipitator.
- (5) The ion concentration is also uniform throughout the precipitator.

Regarding the first assumption, Smith and Penny (1961) have shown that departures from sphericity encountered in practice are not a serious source of error. The second assumption is true for most industrial and natural aerosols. The third assumption is generally valid, except for the finest particles or at considerably reduced pressures. The other assumptions are approached in various degrees of approximation.

The response time of field charging is less than 10^{-3} s, much smaller than the residence time of the particles in a precipitator pipe (~ 1 s). Thus, it is customary to assume that maximum field charging on the spherical particle is reached instantaneously. This maximum charge, commonly called the saturation charge, is written as

$$Q = 4 \pi \epsilon_0 \theta a^2 E(r) \quad (7.39)$$

where

$$\theta = 2 \frac{\kappa_p - 1}{\kappa_p + 2} + 1 \quad \text{with } \kappa_p \text{ as the dielectric constant.}$$

In the limit as $\kappa_p \rightarrow \infty$, θ may be assumed to have the value of 3, which is the universally accepted value for ESP application.

In addition to the ion-bombardment charge, a suspended aerosol particle in an ionized gas acquires a charge by virtue of the random thermal motion of the ions and their consequent collisions and attachments with the particles.

Viewing diffusion charging in terms of ion-current flow to a particle, and thereby generalizing the ion-bombardment mechanism to include the effects of

diffusion, Cochet (1956, 1961) derives the following equation that agrees with his experimental data.

$$Q = \left[\left(1 + \frac{\lambda}{a} \right)^2 + \frac{2}{1 + (\lambda/a)} \right] 4 \pi \epsilon_0 E(r) a^2 \frac{t}{t + \tau} \quad (7.40)$$

Equation (7.40) is an expression for combined field charging and diffusion charging. In the limit $\lambda \ll a$ for larger particles, and $t \gg \tau$, Equation (7.40) gives precisely the saturation charge due to ion bombardment. For smaller particles ($a \leq 1 \mu\text{m}$), the term λ/a will contribute to give additional charging due to ion diffusion.

The drag on a spherical particle moving at velocity w through a gas is

$$F_d = 6 \pi \mu a w \quad \text{for laminar flow,}$$

and
$$F_d = \frac{1}{2} C_D \pi a^2 \rho w^2 \quad \text{for turbulent flow approximation}$$

These have been empirically combined (Wallis, 1966) to give

$$F_d = 6 \pi f \mu a w \quad (7.41)$$

where f is the empirical correction factor based on the particle Reynolds number

$$f = 1 + 0.15 \text{Re}^{0.687} \quad \text{for } \text{Re} < 1000$$

with
$$\text{Re} = \frac{\rho_d w}{\mu}$$

and when Re exceeds 1000, f is empirically assumed to be constant at 18.262.

The drag force given by Equation (7.41) is balanced by the precipitator force

$$F_p = C_s E_p Q \quad (7.42)$$

where C_s is the Cunningham slip correction factor, E_p is the precipitation field intensity (i.e., the local electrostatic field intensity near the collector wall), and Q is the charge on the particle given by Equation (7.40).

Solving Equations (7.41) and (7.42) for the migration velocity, w , results in

$$w = \frac{6 \pi f \mu a}{C_s E_p Q} \quad (7.43)$$

Because w appears implicitly on the right-hand side of Equation (7.43) for the correction factor f , it must be computed by Newtonian iteration. However, the effect of w on the right hand side is relatively small (i.e., its contribution is $Re^{0.687}$), only the first few iterations are needed to assure convergence.

The migration velocity, w , is substituted into the well-known Deutsch-Anderson equation (White, 1963), and the expression for precipitator efficiency becomes

$$\eta = 1 - \exp \left(- \frac{A_c w}{U} \right) \quad (7.45)$$

where A_c is the total collecting area of pipe, and U is the superficial velocity of gas.

Traditionally, the pressure drop of an ESP has been considered to be negligible and has usually been ignored. However, we have included in the Research Cottrell ESP computer model the pressure drop equation based on gas flow through a rough wall pipe. Those equations are the following Reynolds (1974):

$$\Delta P = \frac{C_f g \rho U^2 L_p}{2\pi d_p^5} \quad (7.46)$$

$$C_f = \left[1.74 \ln(d_p / .0254) + 3.48 \right]^{-2} \quad (7.47)$$

where C_f is a dimensionless experimental skin friction coefficient, g is the gravitational constant, ρ is the gas density, U is the gas velocity, L_p is the pipe length and d_p is the pipe diameter.

7.5 Electrocyclone

The cyclone has been in use for particle control in various industries for nearly a century. It is perhaps the best known and least expensive device for particle collection. Consequently, the theory and practice of cyclone operation has been extensively documented.

Historically, cyclones were characterized by a cut size which was computed by equating the centrifugal force on a particle to the drag on the particle due to the radial gas velocity (Lapple, 1951; Stairmand, 1951). This theory is easily understood and widely employed.

In the most common arrangement for inertial separators, the flue gas enters the cyclone tangentially at the top of the cylindrical section and spirals downward into the bottom section. Because of the applied centrifugal force, the dust particles accumulate at the wall and are eventually removed from the mainstream. Thus, the system is governed largely by the equations pertaining to the centrifugal and aerodynamic forces acting on the particles.

Excellent survey papers and articles have been published by Jackson (1963); Caplan (1968); and Batel (1972). The interior flow fields have been explored by Ter Linden (1949); and Kelsall (1951). Standardized designs and practices have been defined by Stairmand (1951), and good agreement has been realized between experiment and the relatively recent theory of Leith and Licht (1972). This body of experience indicates that small cyclones operated at high swirl velocity give better separation efficiency, although consideration must also be given to coarse particle bouncing and limitations associated with system pressure losses and cyclone erosion. Hence, multicyclones (i.e., many small cyclones operating in parallel) and staging have been employed in situations where unusually clean gas is mandated.

Despite the extensive body of literature on the subject, predicting the performance of cyclones in actual service remains an art. This is partly because each application introduces new uncertainties about particle size distribution, state of agglomeration, dust loading, etc. However, major uncertainties are also a consequence of inadequacies in the analytical methods used. Theories appearing in the literature are at best semi-quantitative and are generally incapable of accurately accounting for the effect of design features such as inlet shape, scroll versus tangential inlet, or outlet duct diameter and engagement length, all of which are known to have strong influence on cyclone performance.

The General Electric electrocyclone concept is illustrated in Figure 1-5. The electrocyclone incorporates a clean air shield adjacent to the exhaust duct which prevents short circuiting of dust into the outlet. The clean air is provided by an axial flow separator/electrostatic charger device positioned in the exhaust duct of an upstream cyclone. This axial flow component uses a high-voltage corona electrode to charge the dust particles and improve the

Figure 7-5. Sketch of GE Electro-cyclone

separation efficiency. The vortex field in the axial flow component is provided by the swirling motion naturally present in the exhaust of a conventional reverse-flow cyclone. Within the electrocyclone, the separation of dust is strongly augmented by another high-voltage field applied with an electrode along the axis of the cyclone.

It is assumed that a principal advantage of the electrocyclone concept is that a large-diameter unit can be built to give the same efficiency as a small conventional cyclone. However, that remains to be proven. The large unit appears possible because the electrostatic separative force is independent of the cyclone diameter in contrast to the centrifugal force, which increases for larger cyclones.

A valuable advance in the analysis of cyclone performance has been made by General Electric on their three-region model. The model is the first analytical representation of cyclones to account for observed turbulent mixing between various internal regions. The model accounts for axial variations in the particle concentration and partially allows for radial variations by assuming a different concentration in the vortex core compared to either the inlet region or the main body of the vortex. The model has been validated by extensive comparisons with data from a variety of cyclone geometries and operating conditions.

Mathematical Model

The three-region cyclone model was based on the earlier Leith and Licht model (1972) and includes inherent turbulent flows present in the device. The model incorporates empirical turbulent back-mixing of the suspended particles and provides reasonable agreement with the data. Separate representations are included for (1) the entrance region, (2) the downflow (annular vortex) region, and (3) the upflow (core vortex) region. The model computes the inertial impaction due to tangential and radial components of the vortex flow. The effect of temperature is represented by change in viscosity, and the effect of pressure is not represented. The model also assumes a relatively low particle loading, and thus the effect of particle charge on electrostatic field calculations is assumed to be negligible.

In each region, turbulent mixing is assumed to maintain uniform radial concentration profiles. Thus, conservation of particles in each region requires that

$$Q_{vo} \frac{dn_1}{dz} = -\pi D_c \Gamma_w(z) \quad \text{Region 1} \quad (7.48)$$

$$\frac{d}{dz} [Q_{v(z)} n_2] = -\pi D_c \Gamma_w(z) - \pi D_{v(z)} \Gamma_v(z) \quad \text{Region 2} \quad (7.49)$$

$$-\frac{d}{dz} [Q_{v(z)} n_3] = \pi D_{v(z)} \Gamma_v(z) \quad \text{Region 3} \quad (7.50)$$

where $Q_{v(z)}$ is the axial volumetric flow rate, $\Gamma_w(z)$ is the particle flux to the cyclone wall, $D_{v(z)}$ is the diameter of the core region, and $\Gamma_v(z)$ is the flux of particles from the annular region to the core region (2→3).

The radial particle velocity (u_{pw}) near the cyclone wall can be directly computed from a balance between inertial and Stokes' drag forces. Thus

$$6\pi \mu r_p u_{pw}(z) = \frac{\pi \rho_p r_p^3 u_{tw}^2}{r_c} \quad (7.51)$$

where μ is the gas viscosity, r_p is the particle radius, $u_{tw}(z)$ is the tangential gas velocity, and ρ_p is the density of the particle. Hence,

$$u_{pw}(z) = \frac{\rho_p D_p^2 u_{tw}^2}{9\mu D_c} \quad (7.52)$$

Similarly, the velocity of particles having migrated from the core vortex region into the outer annular region can be computed from the force balance equation as

$$u_{pv}(z) = \frac{\rho_p D_p^2 u_{tv}^2}{9\mu D_v} \quad (7.53)$$

The assumptions are (1) the tangential velocity does not vary axially (Caplan, 1968), and (2) the radial dependence of the tangential velocity is

given by a modified form of the free vortex in an inviscid fluid (Stairmand, 1951; Alexander, 1949) as

$$u_{t(r)} = u_{tw} (r_c/r)^m \quad (7.54)$$

where m is between 0.5 and 1.0.

With the model thus completed, Equations (7.49) to (7.51) become a set of coupled, variable coefficient, ordinary differential equations. The analytical solutions with appropriate matching inlet conditions from mass conservation, the axial concentration profiles in the inlet and outlet regions give the particle collection efficiency as

$$\eta = 1 - \left[K_0 - \left(K_1^2 + K_2 \right)^{1/2} \right] \exp \left[\frac{-\pi D_c u_{pw} d}{Q_v} \right] \quad (7.55)$$

with

$$K_0 \equiv \frac{r_c u_{pw} + r_v u_r + r_v u_{pv}}{2 r_v u_{pv}} \quad (7.56a)$$

$$K_1 \equiv \frac{r_v u_{pv} - r_v u_r - r_c u_{pw}}{2 r_v u_{pw}} \quad (7.56b)$$

$$K_2 \equiv \left(\frac{r_v}{r_c} \right)^{2m} \quad (7-56c)$$

When electrostatic enhancement is in effect, the particle velocities are modified by their Coulombic force. Assuming no particle-particle interactions, the force balance equation between the Stokes' drag and the sum of the inertial and Coulombic forces can be written

$$6\pi\mu r_p u_{pw} = \pi \rho_p r_p^3 u_{tw}^2 / r_c + q E_w \quad (7.57)$$

The electrode structure considered here is a coaxial rod of radius r_e extending the length of the cyclone body. Being insulated from the grounded cyclone wall, the electrode produces a radially directed electric field, approximated by

$$\bar{E}(r) = \frac{V_o}{r \ln(r_c/r_e)} \hat{r} \quad , \quad (7.58)$$

and so

$$E_w = \frac{V_o}{r_c \ln(r_c/r_e)} \quad , \quad (7.59)$$

where V_o is the imposed high voltage. The entering particles are assumed to be charged to their saturation value in a corona-producing electric field (E_c):

$$q = 12\pi \epsilon_o r_p^2 E_c \quad . \quad (7.60)$$

Using the above equations, Equation (7.39) can be simplified to yield

$$U_{pw} = \frac{2\rho_p r_p^2 u_{tw}^2}{9\mu r_c} + \frac{2\epsilon_o r_p E_c V_o}{\mu r_c \ln(r_c/r_e)} \quad (7.61)$$

Similarly, the particle velocity at the core vortex and outer annular vortex boundary can be written

$$U_{pv} = \frac{2\rho_p r_p^2 U_{tv}^2}{9\mu r_v} + \frac{2\epsilon_o r_p E_c V_o}{\mu r_v \ln(r_c/r_e)} \quad . \quad (7.62)$$

These modified velocities, Equations (7.61) and (7.62), can be substituted in the efficiency calculation, Equation (7.55), to model the electrostatic enhancement of the electrocyclones.

Pressure drop in a cyclone will be considered as that between gas inlet and gas outlet in Figure 7-5. A number of factors that contribute to the pressure difference between these points have been identified by Shepherd and Lapple (1939):

- (1) Loss due to expansion of the gas when it enters the cyclone chamber.
- (2) Loss as kinetic energy of rotation in the cyclone chamber.
- (3) Losses due to wall friction in the cyclone chamber.
- (4) Any additional frictional losses
- (5) Any regain of the rotational kinetic energy as pressure energy.

Most pressure drop theories consider factors 2 and 3 to be most important in determining cyclone pressure drop. Customarily, the cyclone pressure loss is expressed as a number of inlet velocity heads, ΔH . Inlet velocity heads can be converted to pressure drop in terms of static pressure head, ΔP , by:

$$\Delta P = \frac{v_g^2 \rho_G \Delta H}{2g\rho_L} \text{ in. H}_2\text{O} \quad . \quad (7.63)$$

The pressure drop prediction incorporated in the user model was formulated by Alexander (1949). Alexander derived the following expression for pressure drop, assuming all pressure losses are due to kinetic energy losses in the gas stream. The equation is written

$$\Delta H = 4.62 \frac{ab}{D De} \left\{ \left[\left(\frac{D}{De} \right)^{2n} - 1 \right] \left(\frac{1-n}{n} \right) + f \left(\frac{D}{De} \right)^{2n} \right\}, \quad (7.64)$$

where

$$n = 1 - \left[1 - \frac{(0.394 D_c)^{0.14}}{2.5} \right] \left(\frac{T}{283} \right)^{0.3} \quad (7.65)$$

and

$$f = 0.8 \left[\left(\frac{1}{n(1-n)} \right) \left(\frac{4-2^{2n}}{3} \right) - \left(\frac{1-n}{n} \right) \right] + 0.2 \left[\left(2^{2n} - 1 \right) \left(\frac{1-n}{n} \right) + 1.5 \left(2^{2n} \right) \right]. \quad (7.66)$$

Here, D_c is the diameter of the cyclone in centimeters, and T is the absolute temperature in degrees Kelvin.

8. Cost Models

The cost modeling task consisted of the following objectives:

- (1) To provide the best estimate of installation and operating costs for each device within the framework of existing data.
- (2) To design cost models that were responsive to changes in the physical size and capacity of the HTHP gas cleanup device.
- (3) To design cost models that provide data that are comparable with the results from the other models.

Certain constraints limited the degree of success possible in meeting these three objectives.

- (1) Generation of completely new cost models was beyond the scope and budget of this contract. Therefore, it was necessary to rely on existing cost data for these devices.
- (2) The availability of the HTHP gas cleanup devices for commercial application is highly variable. Some have been carried to the point of conceptual design in a reference power plant, while others are still operating at the bench or subpilot scale. Because of this, the sophistication of the manufacturers' own cost models varies from detailed to nonexistent depending on the manufacturer.
- (3) Different manufacturers have used different approaches in their cost modeling. All the manufacturers who provided detailed cost data did so with reference to a hypothetical PFBC combined-cycle power plant. There were two basic approaches taken by the manufacturers; some used a modular approach, which permitted the incremental costs due to the installation of the devices to be broken out separately, while others chose a modeling approach, which recognized the closely coupled nature of the devices in the overall plant design. This latter approach may yield results that are somewhat more accurate for the specific plants used in the calculation, but it is difficult to uncouple the cost of the cleanup devices themselves within such a model. Indeed, separating out the device costs from a coupled model would defeat the purpose of having such a model, since METC has little reason to expect improved accuracy from applying costs derived

for a particular situation in a coupled model to other plant sizes and configurations. For these reasons, a modular approach was adopted from the beginning.

- (4) Availability of detailed information depended on the degree of development of each device. This was especially true for the amount of detail available on the costing of labor, downtime, construction, etc. Manufacturers were generally helpful in providing such information, as was available, either through documentation previously supplied to METC or through individual phone calls. Table 8-I shows the source documents used in deriving the cost models.

In order to maximize the consistency between data supplied by different manufacturers for different reference plants, cost information was broken down to a modular basis. A module is the smallest combination of pressure vessel, supports, and cleaning elements for which data were supplied. Note that, using this definition, the reference plant data of several manufacturers were given for configurations involving several modules. Linear interpolation of costs was used for cases where users might require plant sizes which are not integral multiples of the reference plant data. This is reflected in the output as a nonintegral number of modules. For cases where the throughput capacity was below the capacity of a single module, a square root extrapolation was used. The choice of the square root function was arbitrary, and is intended to reflect economies of scale that can be realized from constructing a module of the minimum size. The nature of the code is such that, as better extrapolation functions become available, the code can readily be changed to accommodate them.

To divide the reference plant data into modules, it was necessary to define system boundaries, i.e., where a device began and where it ended within a plant. In all cases, the choice of hardware central to the device and its supporting structure was straightforward. Requirements for the associated hardware, however, were more individualized, depending on input loading conditions and output streams; some devices required an additional cyclone, another

Table 8-I
Source Documents for Cost Data in Hot-Gas Cleanup

<u>Device Name</u>	<u>Source Documents</u>
GE Electrocyclone	Final Report, Novel Methods for Respirable Dust Removal, U.S. Department of Energy Contract No. DE-AC02-79V-10143, September 1980. Report No. EV-10143-1.
Acurex Ceramic Bag Filter	Ceramic-Bag Filter System Commercial Feasibility Study, Interim Report. Prepared by General Electric Company Systems Programs Department for the Acurex Corporation under Contract No. DC59215A, October 1981.
CPC Moving Granular Bed Filter	Draft materials to be used in CPC final reports supplied to Research Technologies Group, Inc., by John L. Guillory, CPC, September 1982.
GE Electrostatic Granular Bed	Appendix C "Preliminary EGB Filter Cost Estimate." Material supplied to RTGI by R. Boericke.
Research Cottrell ESP	Development of a high-temperature/high-pressure electrostatic precipitator, Report No. EPA-600-7-77-132, November 1977.

one a separate bag house. Where operating conditions warranted their use, the cost of additional hardware is included with an appropriate annotation in the ASPEN HISTORY file.

Assignment of indirect costs followed no consistent pattern between manufacturers. A procedure similar to that followed by the General Electric Company was carried through all models because that procedure was tied to hardware costs and could be used for varying states of development to provide a consistent basis for cost comparison. No claim is made, however, that this procedure is more accurate than some alternative method. However, the general procedure was published by General Electric for three of the devices, including the GE electrostatic granular bed, électrocyclone and the Acurex ceramic bag. This procedure included the following assumptions:

- (1) The total direct costs are 34% of hardware costs.
- (2) The indirect costs are 50% of hardware costs.
- (3) The construction loan is 90% of the construction cost.

Finally, we should note that detailed estimates of operating labor costs, downtime, etc., were not available for most of the devices.

Table 8-II indicates the subjective confidence level that may be associated with the cost models. We emphasize that these judgements will (and properly should) change as these models are upgraded due to new developments in design and construction.

Table 8-II

Qualitative Levels of Confidence for Hot-Gas Cost Models
as of October 1982

Device Name	Confidence	Comments *
General Electric Electrocyclone	M - H	Less detail available than for the electrostatically enhanced granular bed, but design is relatively simple.
Acurex Ceramic Bag Filter	M - H	Considerable detail available, well-designed system, well-defined cost.
Combustion Power Company Granular Bed Filter	M	Little detail available for modular approach, but extensive development effort suggests major changes may be unlikely.
General Electric Electrostatic Granular Bed Filter	M - H	Considerable detail available to reflect future changes in design. Costs may be strongly influenced by media choices.
Research Cottrell Electrostatic Precipitator	L - M	Design rapidly evolving through major changes. Little detail available but company is generally experienced in ESPs for other applications.

KEY: L-low M-medium H-high

* - Assumes devices will perform as stated by manufacturer.

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9. Validation

For mathematical performance models to have meaningful existence, they must be validated using actual experimental data and evaluated for their limits of application. However, we must also keep in mind that the experimental data for HTHP gas cleanup devices are scarce and are in a state of flux for the following reasons:

- (1) Since most HTHP gas cleanup device concepts are in an early stage of development, the experimental data base for cleanup efficiency is sometimes not statistically significant.
- (2) Since experimental techniques for sampling and measuring HTHP flue gas are also in the state of development, interpretation of the data is often difficult.
- (3) Since there are only a limited number of PFBC test beds, it becomes increasingly difficult to design, fabricate, schedule, and test a cleanup device to meet the constraints of the PFBC test beds.

Thus, it is important to approach validation of the mathematical performance models by analyzing both the experimental data for the sampling and measurement techniques used and the mathematical models for inclusion of the appropriate physics. In this section, we summarize the sampling and measurement techniques, the current status of experimental device verification process, and the evaluation of the mathematical performance models.

9.1 Sampling and Measurement Techniques

The data required for determining the efficiency of control devices in collecting fly ash are obtained by sampling flue gas and measuring the physical and chemical properties of the ash. Only within the past few years have adequate sampling and analytical techniques become available, and even these techniques are continually being evaluated and improved. There is still no complete agreement among those who use the techniques with regard to the interpretation of some of the results.

The relevant techniques are those used to measure the mass and particle-size distribution of fly-ash collected by sampling the flue gas upstream and downstream of the pollution control device. In the methods used at present,

mass is measured by weighing. The particle-sizing techniques employ inertial impaction, light scattering, diffusion, and electrical mobility of the particles for their characterization. The applicability of these techniques and some of the problems involved in their use on the effluent gas streams from industrial processes are discussed below, as are some improved techniques under development by the Environmental Protection Agency (EPA) and others.

9.2 Particulate Mass Concentration

Mass concentrations of particulate matter in flue gas are measured by drawing a sample of gas through a probe and filter and weighing the collected material. This is the basis of EPA Method 5, the manual that is used in determining compliance with federal regulations (Hamil et al., 1974). There is some difference of opinion as to how the results should be interpreted, especially with regard to condensation of vapors in the probe and filter box. Some investigators making performance tests of control devices on emission sources prefer to use a sampling train that differs from that used in Method 5 in that the filter for trapping particulate matter is located in the stack instead of outside the stack at the end of the sampling probe (proposed Method 17, Lytle, 1976). Thus, condensation of vapors in the probe and filter box is avoided. Another advantage of the alternative system is that the particles are trapped before they enter the probe, from which they may not be adequately recovered despite the probe-washing procedure specified in Method 5.

The only instrumental method that is regularly used for continuous measurement or monitoring of mass emissions is the opacity monitor or light transmissometer, which is commercially available and which has been demonstrated to be usable in power plant stacks. Opacity is only an indirect measure of mass and must be related to it empirically for any specific installation (Lester, 1976; Beutner, 1975).

Other methods that are being considered for monitoring mass emissions include the following, which are mostly still in the development stage (Knapp, 1976; Nader, 1975):

- (1) Light scattering by suspended particles. Designs for measuring forward scattering that have been investigated so far require rather complicated optics. Instrumentation that measures back-scattering

light in the stack is already commercially available. The light source is a light-emitting diode incorporated into a probe inserted into the stack.

- (2) Deposition of electric charge on an electrometer sensor by impinging particles (the triboelectric effect). This technique has been used both in the stack and on extracted samples. Better results are obtained on sources that emit particles relatively uniform in size and charge. Probe losses have been a problem. Some instruments using this approach are already on the market.
- (3) Attenuation of beta radiation by deposited particulate matter. This technique requires an extractive system in which some kind of filter is used to collect the sample. The equipment that has been used so far for interfacing with the flue gas is mechanically complicated and requires excessive maintenance. Probe losses have also been a problem.
- (4) Change in vibrational frequency of a quartz crystal on deposition of particles (piezoelectric microbalance). This method is subject to probe losses and loss of sample by particles not adhering to the surface after deposition. It is subject to interference from gas and moisture. Commercial devices based on this method have been used for monitoring low concentrations of particles in ambient air, but the applicability of the method for monitoring stack gases remains to be demonstrated.

9.3 Particle Size and Size Distribution

These techniques may be classed according to the physical mechanism that is used to obtain the data: inertial, optical, diffusional, and electrical. In their present state of development, instruments employing these techniques are used to complement each other in covering the particle size range below 10 μm . Inertial devices are used for the larger particle sizes ($>0.5 \mu\text{m}$). They include cascade impactors and cyclones. Diffusional and electrical instruments are best suited for the smaller sizes ($<0.3 \mu\text{m}$), and optical counting equipment is applicable for intermediate sizes (McCain, 1974; McCain et al., 1974).

9.4 Sampling from High-Temperature, High-Pressure Gas Streams

Two approaches to the problem of characterizing the particulate emissions in high-temperature, high-pressure gas streams are possible: (1) the instruments may be placed inside the duct to take samples under typical gas conditions, or (2) a probe may be employed to cool the sample and reduce the pressure. Each method has disadvantages. Presumably filters would be adequate means for measuring mass concentration for in situ sampling, but there are not particle sizing instruments for which performance under conditions of high-temperature and pressure has been established. If a probe is used, it is likely that the nature of the aerosol will change when the gas stream is cooled and the pressure reduced, for example, by the condensation of high-boiling-point organic substances.

9.5 Sources of Error

Errors may occur in measurement or in calculation, and also may be intrinsic due to random variations. Table 9-I lists the sources of measurement errors. Some are unavoidable and irreducible, but some can be avoided or reduced to a small value.

From the experimentalist's point of view, all gas cleanup performance data must be carefully evaluated on the basis of (1) sampling and measurement techniques, (2) types and characteristics of dust used, (3) applied temperature, pressure, and electric charge, if any, and (4) the Reynolds number regimes of gas flow characteristics, i.e., laminar, turbulent, and/or boundary layer flows.

9.6 Summary of Model Verification

In this section, we present a summary of validation studies performed to verify the mathematical performance models. The basic approach has been to (1) gather the experimental data, (2) choose a sample case that is general and similar to PFBC conditions, (3) run the computer simulation for the selected case, and (4) compare the results. The verification has been carried out for the electrocyclone, the ceramic bag filter, the electrostatic precipitator, the moving granular bed filter, and the electrostatic granular bed filter.

Table 9-I
Sources of Measurement Errors

1. Errors due to probe
 - a. Anisokinetic sampling velocity
 - b. Probe not parallel to gas flow
 - c. Dust collected on walls or probe and sample line
 - d. Condensation in probe or sample line
2. Errors due to sample collector
 - a. Efficiency less than 100%
 - b. Condensation in collector
3. Errors in sample flow
 - a. Meter inaccuracy
 - b. Leaks
4. Errors in weighing
 - a. Hygroscopic loss
 - b. Loss of sample in handling
 - c. Scale inaccuracy
5. Errors in integrating
 - a. Variations over time or space of dust load or velocity
 - b. Integration error
 - c. Too few traverse points
6. Errors due to personnel
 - a. Reading and recording errors
 - b. Failure to comply with requirements
 - c. Calculation errors

9.6.1 Moving Granular Bed Filter

An extensive data base using the atmospheric fluidized-bed combustor exists at Combustion Power Company. A detailed experimental analysis was carried out during the 1000-hour test in which cascade impactors were used to obtain fractional efficiency data with respect to particle size.

During the 1000-hour test series, the GBF was coupled to a conical bottom 0.46 m^2 (5 ft^2) atmospheric fluidized-bed burning a mixture of Illinois #6 coal and dolomite. The complete pilot unit is referred to as Model 4. Typical operating conditions of the Model 4 subpilot unit are detailed in Table 9-II. Four test segments amounting to continuous operation of the GBF for 100-, 200-, 300-, and 400-hour periods for a total of 1000 hours were run in the fluidized bed combustor. Propane inline air heaters were used to maintain temperatures in the small subpilot system. Due to its physical size, heat losses are significant when attempting to maintain a constant temperature of 870°C .

Metered streams of coal and dolomite enter the spouting fluid bed with the fluidizing air. The design of the fluid bed was intended to produce particulate rather than achieve combustion efficiency. Particulate-laden combustion gases pass first through the recycle cyclone where large particles are returned to the fluid bed. The gases then pass through a primary "separator" cyclone, removing large particulates and providing a more representative size distribution to the GBF. Cleaned combustion gases leave the GBF at 840°C (1550°F), pass through the exhaust particle sampling duct and out to the atmosphere.

Cascade impactors were used to determine the inlet and outlet particulate size distributions. A modified EPA Method 5 procedure was used for total loading measurements and was compared to the total catch of the impactors on the inlet and outlet streams of the GBF. Cooling air was introduced in both sampling ports to lower the temperature of the isolation valves. At the inlet to the GBF, the impactor or the EPA 5 filter was located behind a heated sampling probe inside a temperature-controlled oven. An S-pilot tube was used to measure the local velocity in the duct. A ball valve capable of higher temperatures was located on the GBF outlet sampling port, thus allowing a closer coupling of the impactor to the sampled stream. Both inlet and outlet sample streams were passed through four stainless steel impingers in an ice bath and then through a flow controller, ensuring isokinetic sampling rates.

Table 9-II
CPC 1000-Hour GBF Test Conditions

System Temperature	1120°K	(1550°F)
Inlet GBF Pressure	122 kPa	(17.6 PSIA)
Granular Bed Depth	0.61 m	(2 ft)
Inlet Particulate Loading	6.3 g/m _O ³	(0.66 GR/ACF)
Gas Flow Rate	0.18-0.20 m _O ³ /s	(400-450 SCFM)
Granular Collector Diameter	2 mm	(0.075 in.)
Granular Collector Flow Rate	0.19 kg/s	(25 lb/min)
Coal Feed Rate (Illinois #6)	16 g/s	(2 lb/min)
Dolomite Feed Rate*	6.5 g/s	(0.86 lb/min)

*Dolomite was used for sulfur suppression and to increase particulate concentration.

Impactors were operated between 430-500°C (800-930°F) while sampling the inlet and outlet gas streams.

The basic performance data from the 1000-hour test series are described in Table 9-III. At a pressure drop between 6.0-7.0 kPa (24-28 in. H₂O), the overall particle capture efficiency of the GBF remained near 99% for all tests with an inlet gas particulate loading of 6.0 g/m³ (0.63 gr/ACF) at a gas flow rate of 0.18 m³/s (1600 ACFM) at 840°C (1540°F).

Andersen impactors were used to obtain the aerodynamic particle size distributions. Impactor size classification data converted to fractional efficiencies of the granular bed filter, along with the results from the computer simulation using typical data (Table 9-II), are presented in Figure 9-1. The upper and lower limits of the efficiency band represent measured impactor fractional efficiencies from the test series, and the solid line indicates the analytical model prediction. Repetitive agreement between the experimental tests and analytical predictions with the experimental data depicts the internal consistency of the mathematical model for capture efficiency calculations. The minimum in fractional collection efficiency near 0.7 μm in Figure 9-1 is predictable from the theory of particle capture (Moresco et al., 1979; Rao et al., 1975), where inertial impaction capture is decreasing with particle size and diffusional capture is beginning to become a dominant mechanism of capture. The location of this minimum provides further credence to the measured aerodynamic diameters.

The predicted overall capture efficiency also shows reasonable agreement with the experimental value. The predicted pressure drop, on the other hand, shows a large discrepancy. It should be noted here that the mathematical performance model gives reasonably accurate fractional (as well as overall) efficiency data for a relatively low input dust loading. Furthermore, the pressure drop prediction is sensitive to the input void fraction. The reasons for the importance of the void fraction are (1) the pressure drop equation is a strong function of the nondimensional parameter describing the effective granular bed depth, and (2) the effective granular bed depth is in turn a strong function of the void fraction.

Thus, the limits of application for the mathematical performance model of the moving granular bed filter are (1) verification of the model has been

Table 9-III
Typical Experimental Result from 1000-Hour High-Temperature
Granular Bed Filter Test

Gas Flow	0.18	(1590 ACFM)
Media Rate	0.23 Kg/s	(31 lb/min)
Pressure Drop	6.8 kPa	(27.1 in. H ₂ O)
Inlet Temperature	1110°F	(1538°K)
Inlet Loading	6.9	(0.72 GR/ACF)
Outlet Loading	0.48	(0.005 GR/ACF)
Overall Efficiency	99.2%	
Total Hours on Coal	179	
Total Hours on Steady State	107	
Cascade Exposure Hours	177

Figure 9-1. Collection Efficiency, Counterflow Moving Granular Bed Filter

limited to the high-temperature, atmospheric pressure data obtained by CPC, (2) the inlet duct loading should be limited to about 10 g/m^3 (1.0 grains/ACF), and (3) the model requires accurate input data for the inlet dust loading and the measured void fraction.

9.6.2 Ceramic Bag Filter

A substantial data base exists for the ceramic bag filter developed by Acurex. The following summarizes the status of the experimental efforts accomplished to date.

Ambient Temperature and Pressure Filter Media Tests

A large number of ceramic fiber filter media candidates were subjected to a series of filtration tests using dioctylphthalate (DOP) smoke and AC fine test dust (Shackleton and Kennedy, 1978). Of the ceramic paper, woven, and blanket felt materials tested, blanket felt consisting of small-diameter fibers ($3 \mu\text{m}$) appears most promising because of good filtration performance and relatively high strength.

HTHP Mechanical Durability Test

Three ceramic filters were subjected to 50,000 rapid cleaning pulses, simulating approximately 1 year of operation at high temperatures and pressures (Shackleton and Kennedy, 1978). These tests showed that the fine fiber filters were undamaged by pulse cleaning loads and that the fly-ash dust cake was deposited mostly on the media surface.

HTHP Filtration Test

Ceramic filters about 10 cm (4 in.) in diameter and 0.46 m (1.5 ft) long were tested for a period of 200 hours at high temperatures (approximately 815°C) and pressures (500 to 920 kPa). Filter face velocities ranged from 2.5 cm/s (5 ft/min) to 9 cm/s (18 ft/min). Cleaning pulses occurred offline about once every 10 minutes during testing. Overall particle collection efficiencies in all cases were better than 99.95% and generally better than 99.98%, while baseline pressure drops were maintained at less than 10 inches of water (Shackleton, 1979).

Slipstream Tests at the EPA/Exxon Miniplant

Nine 0.14 m^2 (1.5 ft^2) ceramic bag filters were evaluated under PFBC conditions (850°C , 1000 kPa, 900 mg/m^3 dust loading) at the EPA/Exxon mini-plant (Ernst et al., 1979). Pressure drops of under 2 kPa (8 in. H_2O) have been maintained for over 6 hours at face velocities of 7.5 cm/s (15 ft/min). Collection efficiencies ranged from 96 to 99.5%.

Chemical Stability Test at National Coal Board CURL

Samples of Saffil filter media were exposed for 1000 hours in the hot gases from the PFBC at CURL (Leatherhead, England). The Saffil before and after exposure was examined by SEM and X-ray. Both methods show the fibers to be physically and structurally unchanged by conditions present in the PFBC stream (Unpublished Data, 1980).

Westinghouse Tests Under Simulated PFBC Conditions

Ceramic bags, 0.2 m (8 in.) in diameter by 1.52 m (5 ft) long, were tested in a five-module, 4.6-m^2 (50-ft^2) filter unit under simulated PFBC conditions (111 kPa, $427\text{-}827^\circ\text{C}$) at Westinghouse. Measured overall collection efficiencies were 99.5 to 99.8% at a filter face velocity of 5 cm/s (10 ft/min). Difficulty was experienced in bag cleaning, with evidence of nonuniform cleaning of all bags. Increases in overall baseline pressure drops were observed over a period of about 100 hours of testing. However, when blowback pressures were increased from 1482 kPa (200 psig) to 2861 kPa (400 psig), better but not complete control of pressure drop was achieved. The nonuniform bag cleaning and observed increases in baseline pressures are attributed to the of cleaning parameters in these tests. A series of tests are underway at Acurex to optimize pulse cleaning parameters (Unpublished Data, 1980).

It has been reported (Private Communication with R. Chang, 1981) that because of extremely high filtering efficiency of the ceramic bag, only the overall efficiency data taken by the final absolute filter over several hours of operation are available. A cascade impactor has been used in an attempt to obtain a set of fractional efficiency data with respect to particle size, but the process took over 10 hours to accumulate enough sample. Because of this sampling time requirement, the Acurex team was unable to get a statistically

meaningful correlation between the input dust loading and the output impactor sample. Thus, a critical comparison between the experimental result and the mathematical prediction has been made using the pressure drop data and the overall filtering efficiency. Table 9-IV summarizes the Acurex test results.

As the results indicate, the analytical predictions for the pressure drop are in reasonable agreement with the experimental results. The collection efficiency has been empirically determined at 99.9%, and the results reflect that agreement.

9.6.3 Electrostatic Granular Bed Filter

The granular bed filter design discussed in Section 7.3 was completed in 1981. The purpose of the design was to provide a method to test a granular bed filter (no electrostatics) in a high-temperature environment with PFBC-type contaminants. To date, however, no experimental data using the electrostatic enhancement under PFBC conditions have been reported by General Electric.

The proposed test for verifying the effects of electrostatic enhancement will be conducted under the following conditions:

Bed Temperature:	1600°F
Bed Pressure:	Atmospheric
Inlet Dust Loading:	2.8 gr/ACF
Flow Rate:	500 ACFM

The granular bed design parameters are:

Height:	6.3 m (21 feet)
Outer Bed Diameter:	1.3 m (51 inches)
Inner Bed Diameter:	0.7 m (27 inches)
Bed Depth:	0.3 m (12 inches)
Media Diameter:	2 mm
Applied Voltage:	Variable

Figure 9-2 illustrates the predicted performance under these conditions. At the applied voltage of 30 kV, only 89.9% overall efficiency is predicted with more than 10% of particles under 3 μ m penetrating the bed. At increased applied voltages, an increase in overall collection efficiency as well as in

Table 9-IV. Summary of Acurex Test Results

	1	2	9	10
Operating Conditions				
Pressure (PSIG)	120	120	120	120
Flow rate (ACFM)	37	36	23	29
A/C Ratio	15	15	10	13
Dust Type	Exxon	Exxon	Curtis-Wright	Westinghouse
Dust Loading	0.4 g/ACF	0.4 g/ACF	0.37 g/ACF	0.26 g/ACF
Bag Parameters				
Size (ft x in)	3' x 3"	3' x 3"	3' x 3"	3' x 3"
Screen	SS	SS	SS	SS
Fiber	Saffil	Saffil	Saffil	Saffil
Pulse Parameters				
Pulse Reservoir	400	400	400	400
Size (in ³)				
Pulse Pressure (PSIG)	270	270	250	---
Pulse Length (ms)	100	100	100	100
Pulse Interval (min)	11	30	15	15
Test Duration (hr)	4.5	7.5	7	15.6
Experimental Results				
Pressure Drop	5/1.75	1/3	7.5/1	7.5/1.2
max/min (in. H ₂ O)				
Collection Efficiency	--	99.986	99.983	99.89
Analytical Prediction				
Pressure Drop	5.2/1.75	7.2/3	7.4/1	7.3/1.2
max/min (in. H ₂ O)				
Collection Efficiency	99.9	99.9	99.9	99.9

Figure 9-2. Predicted Collection Efficiency for Electrostatic Granular Bed Filter Over Varying Applied Voltages

fine particle capture efficiency can be observed. Without the electrostatic enhancement, however, only a 72% overall collection efficiency with poor fine particle control was achieved.

In summary, the electrostatic granular bed with the cross-flow design would require electrostatic augmentation to achieve effective removal of fine particles, but further evaluation must await experimental data currently being processed by General Electric. As more data become available, the limits of application, the pressure drop correlation, and the corona stability will be defined.

9.6.4 Electrostatic Precipitators

Because ESPs have been popular for industrial particulate control, much information is available for ESP performance at conventional temperatures and pressures. However, at high-temperature, high-pressure conditions, only clean gas corona data have been generated and no collection testing has been done.

Thus, for the validation effort we have examined the existing precipitator performance data and chosen a set that may partially reflect the severe environment in which the HTHP precipitators must operate.

The data set chosen here is a hot-side precipitator installed on a pulverized-coal boiler burning a southwestern (low-sulfur) coal (McCain et al., 1975). The fractional efficiency data were taken by three mass particle counters: a diffusion battery was used to measure the inlet and outlet mass fraction for particle diameters ranging between 0.005 and 0.2 μm ; an optical particle counter was used for particle diameters ranging between 0.5 and 3.0 μm ; and a cascade impactor was used for particle diameters less than 0.5 μm . The hot-side precipitator operated at 630°F with average current density at the tubular collector wall of 35 $\mu\text{a}/\text{ft}^2$. The electrical conditions were reported to be optimum and a high specific collection area of 430 $\text{ft}^2/1000 \text{ ACFM}$ resulted in an overall collection efficiency of 99.3 percent. The average specific particle density (of mostly ash and unburned coal particles) was 1.77 g/cm^3 .

To simulate the hot-side precipitator, a 15-ft long pipe with an inside diameter of 8 in. and a centrally located discharge wire of 0.109 in. in diameter was specified. Operating conditions were at 630°F and 1 atmosphere.

Applied voltage was varied until the matched condition of $35 \mu\text{a}/\text{ft}^2$ at the collector walls was obtained.

The result of the simulation run is presented in Figure 9-3. The fractional efficiency measured by the three different techniques is appropriately labeled and the solid line indicates the numerical prediction. As depicted in the figure, the fractional efficiency results show a reasonable agreement for the hot-size precipitator, with a slight overestimation for smaller particle size ranges. For particles smaller than $0.2 \mu\text{m}$, diffusion charging is the dominant charging mechanism for the particle, and a higher collection rate is observed for that size range. Particles larger than about $3 \mu\text{m}$, on the other hand, are charged primarily by field charging, which increases for the larger particles. As indicated in the figure, more than 99.5% of the particles with diameters greater than $3 \mu\text{m}$ were collected. The micron-size particles ($0.2 < d_p < 3 \mu\text{m}$), however, are most difficult to collect as neither diffusion charging nor field charging is an effective charging mechanism at this particle size range. The dip in the fractional efficiency curve clearly indicates these phenomena.

Although the simulation run for the hot-side precipitator shows favorable agreement with the experimental data, any conclusions must be limited to low to moderate temperatures and atmospheric pressure conditions. As has been stated before, some corona data have been obtained for HTHP conditions, but only clean gases were used. Thus, the effects of particles on the electric condition at high temperatures and pressures remain unknown. A continued effort to validate the current mathematical model at HTHP conditions is therefore strongly suggested.

In summary, the model shows reasonable agreement with the actual experimental data for limited cases. The model does not take into consideration the effects of particle reentrainment and varying particle resistivity. The model assumes a low particle loading (less than $1.0 \text{ grain}/\text{ACF}$), which simplifies the modification of the electrostatic field due to the charged particles. The pressure drop model using the rough walled duct approximation gives favorable agreement with the conventional data, but it is also unproven at HTHP conditions. A more detailed analysis (both experimental and computational) will help to improve the model as well as assist in the future design of ESPs with a greater range of applications.

Figure 9-3. Comparison Between the Measured and Computed Fractional Collection Efficiency for a Hot-Side Electrostatic Precipitator Installed on a Pulverized-Coal Boiler

9.6.5 Electrocyclone

Detailed evaluations of the GE electrocyclone computer model were conducted using the test data taken from reports of experiments conducted at CURL in Leatherhead, England. The reports are DOE/ET 14129-1101 (Vol. 1), DOE/ET/14129-1102 (Vol. 2), and DOE/ET/14129-1103 (Vol. 3). Those reports contain the results of six tests carried out on the pressurized, fluidized-bed facility where the test combustor exhaust was split into two streams, each of which passed through a series of three cyclones before encountering candidate turbine alloy test samples. Six distinct cyclones were used in these tests. An earlier computer cyclone model (R. A. Bajura and S. K. Beer, 1981) used the results from Test 4 (the CURL reports, Vol. 2, pp. 135-178). The same test data were used to conduct the evaluations reported below.

The schematics of the six cyclones and critical dimensions are shown in Figures 9-4 through 9-7. Tables 9-V and 9-VI are summaries of the operating conditions and critical dimensions of the test devices. The temperature listed for the tertiary cyclone on Stream 2 in Table V is assumed to be a typographical error in the CURL report. The primary and secondary cyclones on Stream 2 are constructed coaxially; that is, the secondary cyclone is contained within the primary. The diameters of the primary underflow and exit pipes were adjusted to reflect that configuration.

Figures 9-8 through 9-13 are corresponding plots of the results. Overall, agreement with the CURL grade efficiency data is not good below 10 μm . However, there is a great deal of uncertainty about grade efficiency measurements below 10 μm . Although every effort was made by the personnel at CURL to record correct grade efficiency data, the reports show large uncertainties about the measurement results. For example, grade efficiencies for 2 μm particles range from 10 to 90% among the different measuring systems (Coulter, Bahco, Pilat and PMS). For particles greater than 10 μm in diameter, the spread in grade efficiency, as measured by the various techniques, was only 10%. In general, the Coulter system is considered very good above 10 μm but its uncertainty increases rapidly below that size.

Primary Cyclone

Mass flow = 2.0 lb/s
 Inlet velocity = 60 ft/s
 Body diameter = 20 inches

$a/D = 0.46$
 $b/D = 0.21$
 $A/D = 0.30$
 $S/D = 0.90$
 $h/D = 1.32$
 $H/D = 3.80$
 $B/D = 0.40$

Secondary Cyclone

Mass flow = 2.0 lb/s
 Inlet velocity = 100 ft/s
 Body diameter = 15.4 inches

$a/D = 0.46$
 $b/D = 0.21$
 $A/D = 0.19$
 $S/D = 0.90$
 $h/D = 1.32$
 $H/D = 3.80$
 $B/D = 0.40$

Figure 9-4. Dimensions of Primary and Secondary Cyclones on Stream 1

Tertiary Cyclone

Mass flow = 20 lb/s

Inlet velocity = 145 ft/s

Body diameter = 13 inches

$a/D = 0.5$

$b/D = 0.19$

$A/D = 0.5$

$S/D = 0.5$

$h/D = 1.5$

$H/D = 4.0$

$B/D = 0.38$

Figure 9-5. Dimensions of Tertiary Cyclone on Stream 1

Primary Cyclone

Mass flow = 1.2 lb/s
 Inlet velocity = 65 ft/s
 Body diameter = 36 inches

$a/D = 0.18$
 $b/D = 0.09$
 $A/D = 0.24$
 $S/D = 0.74$
 $h/D = 0.83$
 $H/D = 1.80$
 $B/D = 0.22$

Tertiary Cyclone

Mass flow = 1.2 lb/s
 Inlet velocity = 110 ft/s
 Body diameter = 10 inches

$a/D = 0.5$
 $b/D = 0.25$
 $A/D = 0.4$
 $S/D = 0.75$
 $h/D = 2.0$
 $H/D = 4.0$
 $B/D = 0.4$

Figure 9-6. Dimensions of Primary and Tertiary Cyclone on Stream 2

Secondary Cyclone

Mass flow = 1.2 lb/s

Inlet velocity = 105 ft/s

Body diameter = 10 inches

$$a/D = 0.5$$

$$b/D = 0.25$$

$$A/D = 0.5$$

$$S/D = 0.75$$

$$h/D = 2.0$$

$$H/D = 4.0$$

$$B/D = 0.4$$

Figure 9-7. Dimensions of Secondary Cyclone on Stream 2

Table 9-V

Summary of the CURL Cyclone Operating Data
Used in Evaluating the GE Model

Cyclone	---- Stream 1 ----			---- Stream 2 ----		
	1	2	3	1	2	3
Temperature						
Degrees F	1450	1440	1430	1420	1400	*1500
Degrees K	1061	1055	1050	1044	1033	1089
Pressure						
PSIA	90.4	90.1	86.3	90.4	90.3	89.4
N/m ²	623K	621K	595K	623K	623K	616K
Gas Density						
lb/ft ³	0.134	0.135	0.129	0.136	0.138	0.129
kg/m ³	2.15	2.15	2.07	2.18	2.20	2.07
Particle Density						
lb/ft ³	201	203	207	207	202	204
kg/m ³	3221	3237	3276	3311	3230	3262
Gas Flow						
lb/hr	7170	7170	7170	4390	4390	4390
kg/s	0.903	0.903	0.903	0.553	0.553	0.553
Particle Loading						
lb/hr	77.8	6.12	1.42	51.8	8.23	3.13
kg/s (x 0.001)	9.81	0.771	0.179	6.53	1.04	0.394

*Temperature is assumed to be a typographic error in the CURL report.

Table 9-VI

Dimensions of CURL Cyclones

STREAM 1 CYCLONES	Pri	Sec	Ter
Cyclone inlet axial extent (inches)..... :	9.20	7.08	6.50
Cyclone inlet radial extent (inches)..... :	4.20	3.23	2.47
Cyclone diameter (inches)..... :	20.0	15.4	13.0
Underflow diameter (inches)	8.00	6.16	4.94
Length of overflow (inches)..... :	18.0	13.9	6.50
Length of cylinder plus cone (inches)..... :	76.0	58.5	52.0
Cyclone pipe exit diameter (inches)..... :	6.00	2.93	6.50
Vortex strength (range 0.5-1.0) dimensionless..:	0.70	0.70	0.70

STREAM 2 CYCLONES	Pri	Sec	Ter
Cyclone inlet axial extent (inches)..... :	6.48	5.00	5.00
Cyclone inlet radial extent (inches)..... :	3.24	2.50	2.50
Cyclone diameter (inches)..... :	36.0	10.0	10.0
Underflow diameter (inches)	*6.84	4.00	4.00
Length of overflow (inches)..... :	26.6	7.50	7.50
Length of cylinder plus cone (inches)..... :	64.8	40.0	40.0
Cyclone pipe exit diameter (inches)..... :	*7.66	5.00	4.00
Vortex strength (range 0.5-1.0) dimensionless..:	0.70	0.70	0.70

* Diameters adjusted for coaxial configuration

Figure 9-8. Comparison for CURL Test 4, Stream 1, Primary

Figure 9-9. Comparison for CURL Test 4, Stream 1, Secondary

Figure 9-10. Comparison for CURL Test 4, Stream 1, Tertiary

Figure 9-11. Comparison for CURL Test 4, Stream 2, Primary

Figure 9-12. Comparison for CURL Test 4, Stream 2, Secondary

Figure 9-13. Comparison for CURL Test 4, Stream 2, Tertiary

For the secondary and tertiary cyclones, the CURL grade efficiencies for particle sizes greater than a characteristic size are considered to have extremely large error bands because the relative amount of collected dust was extremely small. Those characteristic sizes were

Stream 1:	Secondary:	4 μm
Stream 2:	Secondary:	11 μm
	Tertiary:	8 μm

The relatively flat CURL grade efficiency data of the Stream 1 tertiary cyclone is believed to be real, although there was concern about a leaky flange and its effect on performance.

Although the overall collection efficiency predictions do not agree well with the CURL data, the Alexander (Leith and Mehta, 1973) pressure drop model reasonable agreement. The largest difference occurs with the Stream 2 tertiary data, where the pressure drop was predicted by the model to be 1.16 psi (1.3% of inlet), and it was measured to be 2.32 psi (2.6% of inlet).

Sensitivity studies were conducted for the Stream 1 primary cyclone by varying the gas and particle densities and the vortex strength and applying electrostatics. The reference case for all studies is listed under the gas density section. The results from the density variations are as expected. If the gas density increases, there will be a greater drag on particles moving to the wall, hence lower collection efficiency. Conversely, when particle density increases, the particles will be subject to a greater centrifugal force, and this results in increased collection efficiency. The vortex strength affected only the collection efficiency of particles from 1.5 to 11.4 μm . The reference vortex strength was 0.7.

The CURL test reports show that electrostatic enhancement was not an important factor. The change in collection efficiency was negligible when the system was charged with 30 kV. Tests conducted with the model produced the same results.

Overall, the results of the cyclone evaluations are mixed. There is good agreement between the CURL data and the model predictions for pressure drop and electrostatic effects, but considerable doubt about collection efficiency predictions for particle sizes less than 10 μm .

10. Conclusions

This report represents the culmination of an extensive investigation into the development, coding, and evaluation of mathematical performance models for particulate control devices. The work has involved (1) a literature survey as well as facility visitations to determine the current status of device development, (2) evaluation of available performance models and experimental data, (3) evaluation of available data on the effects of alkali compounds at high temperatures, (4) coding of the computer models that are computationally efficient and input/output-compatible with the ASPEN executive code, (5) development of cost analysis based on the available (but limited) information, and (6) evaluation of mathematical models for their limits of the application and appropriateness of the physics involved.

From the literature survey and facility visitations, much information on the advanced gas cleanup systems has been obtained and examined. In order to make a recommendation as to which initial set of gas cleanup devices be included for development of system simulation models, a detailed set of evaluation criteria was established. Based on these criteria, the following five devices were chosen to be incorporated into the ASPEN executive code:

- Moving Granular Bed Filter
- Ceramic Bag Filter
- Electrostatic Granular Bed Filter
- Electrostatic Precipitator
- Electrocyclone

The design of software for this contract followed the life-cycle stages of (1) system analysis, (2) system specifications, (3) algorithm design, (4) detail design, (5) coding, (6) testing, (7) maintenance, and (8) operation. Validation studies show that the codes accurately represent available models. In addition, the codes are self documented, easy to use and read, and easy to maintain. A major development was the creation of a common framework for building computer models of cleanup devices for the ASPEN code. The framework and implementation procedures are applicable to any cleanup device and are not restricted to PFBC systems.

Many questions still remain in the area of alkali control in PFBC gases. From the available literature, we have concluded that (1) the presence of

alkalis in PFBC environments can be modeled, at least as far as their production from dolomite is concerned, (2) the removal of alkalis for hot-gas cleanup systems can be modeled because the forms of the equations (Michaelis-Menton) are available, but the required parameters are not all available, (3) Westinghouse is the only company that measured both Na absorption and K desorption, and the Na-K displacements on a getter granule may thus create an additional alkali problem in a PFBC environment.

As for the model validation effort, it became apparent that the experimental data for the advanced gas cleanup devices are limited in amount and are in a state of flux. This is attributed to the fact that (1) most advanced gas cleanup device concepts are in an early stage of development, and the experimental data base for cleanup efficiency sometimes is not statistically significant, (2) the experimental techniques for sampling and measuring the particulate-laden gases in HTHP environments are also in a state of flux, and the interpretation of the data is often difficult, and (3) there are only a limited number of PFBC test beds, and only the electrocyclone and the ceramic bag filter have been tested under HTHP conditions.

Thus, future model validation efforts must be carried out in parallel with the actual hardware development. To effectively define the limits of model application, more experimental data at simulated PFBC conditions are required. In particular, for the devices that use electrostatic enhancement (ESP, electrocyclone, and electrostatic granular bed), a practical understanding of the effects of the particle space charge at varying temperatures and pressures is essential. A continued effort to validate the current mathematical models at a greater range of operating conditions based on the experimental effort currently underway is therefore strongly suggested. A better understanding of the characteristics of these advanced gas cleanup devices, both from experimental observations and the basic physical point of view, will directly impact the future development of emission control devices. Although the models show reasonable agreement with the actual experimental data, the range of application and validation data are limited. These devices are affected by such temperature-dependent variables as the corona generation, stability, and dust resistivity. Thus, more well-defined experimental data in these areas will benefit both the hardware design and the mathematical model development.

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