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MEASURING AND MODELING SOLIDS MOVEMENT IN A LARGE, COLD
FLUIDIZED BED TEST FACILITY

Second Quarterly Report for January 1—March 31, 1980

By
Thomas J. Fitzgerald
Robert V. Mrazek
Stephen D. Crane

March 1980

Work Performed Under Contract No. AC01-79ET13152

Oregon State University
Department of Chemical Engineering
Corvallis, Oregon

U. S. DEPARTMENT OF ENERGY

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TRACER MEASUREMENT WITH SIMULTANEOUS BUBBLE DETECTION

A number of tests were performed which indicate that spreading ferrite tracer can be detected by inductor coils which simultaneously indicate the presence of gas bubbles. The same coil is used to measure both phenomena, and this allows the ferrite tracer signal to be corrected to eliminate signal dropouts which occur when a bubble surrounds the probe.

The technique takes advantage of a phase difference between the response to copper and the response to ferrite. Ferrite material in the vicinity of the probe causes the bridge shown in Figure 1-1 to become imbalanced, but the imbalance signal is nearly 90° out-of-phase with the driving signal. Copper causes an imbalance signal which is approximately 180° out-of-phase with the driving signal.

The inductor probe used in the tests was cylindrical in shape, about 1.25 cm long and 1.25 cm in diameter, with 460 turns of fine copper wire. The inductance in air was measured at 13.0 mH. The inductor coil wound on a 1.25 cm diameter rod was inserted vertically in a 7.5 cm diameter fluidized bed consisting primarily of copper particles.

Three runs were made in the copper-ferrite bed. The experimental conditions were:

Run #	1,2	3
Copper powder:	about 150 μm	150 μm - 450 μm
Ferrite powder:	about 74 μm	840 μm cast in polyester resin

The total slumped volume was ~500 ml; slumped height 9 cm. Ferrite was added 1 ml at a time until it equaled 1.2% and 1.8% of the bed volume for runs one and two, respectively. The base line for 100% copper with the bed fluidized yielded approximately 0.9V. The results of runs one and two

Figure 1-1. Inductive Bridge Circuit.

are illustrated in Figures 1-2 and 1-3. For run three, the bed contained 400 ml of solids with a slumped bed height of 7.6 cm. Larger ferrite particles were added in approximately 1 ml portions. The data for run three (Figure 1-4) showed more fluctuations than the first two runs, especially at lower concentrations of ferrite particles, probably because of statistical fluctuations in the number of ferrite particles included in the probes pick-up volume.

FIG 1-2

FIG 1-3

FIG -1-4

Figures 1-2, 1-3, 1-4. Illustrations of the Inductive Bridge Circuit Output Signals.

TRACER RELEASE AND COLLECTION

Construction of a prototype ferrite collection-release module was completed. The details of this prototype were given in the previous report (DOE/ET/13152-1). The active section of the module consists of a 10-inch length of alternate magnets and spacers to form a tube with an outside diameter of 1/2 inch. The active section was enclosed by a sleeve of latex or Mylar which could be inflated to release tracer or deflated to collect tracer.

A set of experiments was conducted to determine the collection and release characteristics of this device. These tests were carried out in a 30 cm by 30 cm bed loaded with 77 kg. of zirconia (average size of 300 μm) and 0.9 kg. of ferrite tracer (525 μm diameter), which produced an average bed depth of 22 cm. The horizontal tube containing the test section was centered in the slumped bed.

For tests of collection characteristics, the bed was fluidized for about one minute with no tracer collection to ensure a thorough mixing of bed material. Tracer was then collected for a measured period of time and the bed material was dumped. The collector was then removed and the amount of tracer collected was determined by weighing. The results of these tests for different superficial air velocities and collection times are given in Tables 2-1 and 2-2 for sleeves of latex and Mylar, respectively. The time listed in these tables includes dumping time, since collection was still occurring during this period.

The tests of the release characteristics were conducted in a similar manner. The bed was first fluidized for about one minute (with no tracer collection) to ensure good mixing. Tracer was then collected for a given period of time, released for an equal period of time, and the bed material

Table 2-1. Tests on Ferrite Collection-Release System with a Latex Sleeve

← Collection →		← Release →		← Net →
Collection Period, sec.	Ferrite Collected, g	Release Period, sec.	Ferrite Remaining, g	Ferrite released, g (collected-remaining)

Superficial Air Velocity: 1 ft/sec.

58	13.7	57	3.4	9.9
57	12.8 > 13.25	57	3.3 > 3.35	
109	14.0	106	5.2	8.35
105	13.9 > 13.95	107	6.0 > 5.6	

Superficial Air Velocity: 2 ft/sec.

59	9.8	57	3.5	5.95
57	9.9 > 9.85	60	4.3 > 3.9	
107	9.6	105	4.3	4.75
108	9.2 > 9.4	107	5.0 > 4.65	

Superficial Air Velocity: 3 ft/sec.

59	8.7	59	3.5	4.55
60	8.4 > 8.55	62	4.5 > 4.0	
108	6.2*	110	5.8	2.7
109	8.7 > 8.7	109	6.2 > 6.0	

* Anomalous result. Not used.

Table 2-2. Tests on Ferrite Collection-Release System with a Mylar Sleeve

<----- Collection -----> <----- Release -----> <----- Net ----->

Collection Period, sec.	Ferrite Collected, g	Release Period, sec.	Ferrite Remaining, g	Ferrite released, g (collected-remaining)
----------------------------	-------------------------	-------------------------	-------------------------	--

Superficial Air Velocity, 1 ft/sec.

55	24.1	50	6.7	16.45
55	18.6* > 24.1	53	8.6 > 7.65	
91	28.8	91	11.9	16.45
91	28.2 > 28.5	92	12.2 > 12.05	

Superficial Air Velocity, 2 ft/sec.

62	21.4* > 16.2	65	5.1	11.6
61	16.2	61	4.1 > 4.6	
91	14.2	96	5.9	9.3
90	14.8 > 14.5	101	4.5 > 5.2	

Superficial Air Velocity, 3 ft/sec.

59	19.4	59	3.6	15.0
62	19.1 > 19.25	61	4.9 > 4.25	
105	23.9	100	3.9	18.15
102	20.6 > 22.25	101	4.3 > 4.1	

* Anomalous result. Not used.

was dumped. The remaining tracer (on the device) was again determined by weighing. These results also are given in Tables 2-1 and 2-2, and, as before, the tabulated release time includes the dumping time.

The last column of data in Tables 2-1 and 2-2 represents the net release or collection of tracer, for even when the device is in the release mode, there will always be a residual amount of tracer attracted to it. The data indicate that a larger amount of tracer can be released into the bed or collected from the bed with the Mylar sleeve than with the latex sleeve. This offers the possibility of using fewer release-collection devices for a given amount of tracer in the bed.

COLD FLUIDIZED BED MODELING

Much work has been done recently using cold fluidized bed models to study many of the characteristics of hot fluidized bed combustors.

The advantages of cold fluidized bed modeling are many. Cold models are cheaper to construct and operate, easier to instrument and easier to modify for studying a variety of parameters. However, little work has been done to determine if cold models have the same hydrodynamic properties and other characteristics of hot fluidized bed combustors.

As a first step toward addressing this question experiments were conducted to determine if dimensionless groups can be used to correlate fluidization hydrodynamics. Four dimensionless groups, first described by Scharff, M.F., et al, (1) were chosen for this study. They are:

$\frac{\rho_s}{\rho_g}$	the ratio of solid density to gas density.
$\frac{d_p}{L}$	the ratio of particle size to a characteristic bed dimension.
$\frac{U^2}{Lg}$	the ratio of superficial gas velocity squared to the product of the acceleration of gravity times a characteristic bed dimension.
$\frac{d_p U \rho_g}{\mu_g}$	the particle Reynolds number: the ratio of the superficial gas velocity times particle size to the kinematic viscosity of the gas.

If one sets all these dimensionless numbers to be identical when comparing two beds, then both beds should have similar fluidization hydrodynamics.

Procedure

To test this theory experiments were conducted using two cold fluidized bed models with identical dimensionless numbers. The experimental conditions were:

	Cork/Air	Sand/R12 Freon
	Bed #1	Bed #2
Bed width inches (cm)	19.0 (48.3)	2.22 (5.64)
Bed thickness inches (cm)	5.0 (12.7)	0.58 (1.47)
Distributor Plate	Glass Spheres	Lead Shot
(a) Diameter inches (cm)	0.5 (1.27)	0.079 (0.20)
(b) Height	9.0 (22.9)	1.05 (2.67)
Bed Media	Cork	Sand
(a) Mean particle size	2068 μ	240 μ
(b) Density	0.15	2.65
Fluidizing Gas	Air	R12 Freon (Dichlorodifluoromethane)
(a) Bed pressure, psia	≈ 14.7	61.0
(b) Temperature °F (°C)	77° (25°)	77° (25°)

Scaling Factors

$$\frac{L_1}{L_2} = 8.558$$

where: L_1 = a characteristic bed dimension for the cork/air system

L_2 = a characteristic bed dimension for the sand/Freon system

$$\frac{U_1}{U_2} = 2.925 \quad \begin{array}{l} \text{velocity/time} \\ \text{scaling factor} \end{array} \quad \begin{array}{l} U_1 = \text{the superficial gas velocity for} \\ \text{the cork/air system} \\ \\ U_2 = \text{the superficial gas velocity for} \\ \text{the sand/Freon system} \end{array}$$

The $\frac{L_1}{L_2}$ scaling factor of 8.558 was used to scale all of the physical dimensions parameters (i.e. bed dimensions, mean particle size) of the two beds. The densities were scaled by $(\frac{L_1}{L_2})^{.666}$.

The ratio of the superficial gas velocities, $\frac{U_1}{U_2}$, was used to determine if the fluidization hydrodynamics were similar. Pressure measurements were made across both beds at U_{mf} and various multiples of U_{mf} (up to $2.0 U_{mf}$) (Figure 2-1). These data were analyzed using a Fast Fourier Transform power spectrum to determine the frequency of the pressure fluctuations. The ratio of the average frequencies should be equal, to the predicted velocity/time scaling factor of 2.. The ratio of the measured superficial gas velocities at U_{mf} should also be equal to this value..

Movies of both systems were also taken for each experimental condition. The film speed was adjusted to take into consideration the differences in the superficial gas velocities. Since the ratio of the gas velocity of the cork/air system to that of the sand/Freon system is predicted to be 2.92 the movies of the sand/Freon system were taken at 54 frames per second vs 18 frames per second for the cork/air system. In real time the cork/air system is 2.92 times slower than the sand/Freon system.

Movies and pressure measurements were also taken in a slugging bed configuration using an 8" diameter cylinder for the cork/air system and a 1" diameter cylinder for the sand/Freon system. In addition, movies were made comparing tungsten carbide/water system with an expanded-polystyrene/air system with identical dimensionless number at velocities up to $4 U_{mf}$.

Results

Movies taken of the sand/Freon system vs. the cork/air system in two bed configurations (rectangular and cylindrical) and tungsten carbide/water vs. expanded polystyrene/air in a cylindrical bed configuration at U_{mf} and various multiples of U_{mf} showed similarities in bubble growth and solids flow patterns. In addition, in the slugging beds the slugs appear to be nearly equal in scaled length, they rise to the same scaled height, and they rise at nearly the same scaled velocity. The measured U_{mf} ratio, $\frac{U_{mf} \text{ Cork}}{U_{mf} \text{ Sand}}$ in the rectangular beds was within 20% of the predicted value of 2.92. The variation between the measured and predicted values may be a result of the angular shape of the cork particles which, because of their interlocking nature, yield a higher U_{mf} .

The ratios of the average frequency of pressure fluctuations for the cork/air vs. sand/Freon system in the rectangular bed configuration were near the predicted value of 2.92 (see Figures 3-2, 3-3, 3-4, 3-5, 3-6, 3-7, and Table 3-1). These frequencies were not calculated for the slugging bed configurations.

TABLE 3-1.

U_{mf}	1.0	1.2	1.4	1.6	1.8	2.0
Ratio of average frequency of pressure fluctuations	*	*	*	2.8	2.0	2.2

*Pressure fluctuations in the cork/air system were too low to measure.

Conclusions

The results strongly indicate that the four dimensionless groups can be used for scaling cold models. That is, the performance of one cold model

Figure 3-1. Rectangular Bed Configuration of Cold Scaled Fluidized Bed Model.

Figure 3-3. Frequency of Pressure Fluctuations at 1.6 U_{mf} Sand/Freon System with Average Measured Value Indicated. Ratio of Average Frequency is near Predicted Value of 2.9.

Figure 3-2. Frequency of Pressure Fluctuations of Cork/Air System at 1.6 U_{mf} with Average Measured Value Indicated.

Figure 3-4. Frequency of Pressure Fluctuations of Cork/Air System at $1.8 U_{mf}$ with Average Measured Value Indicated.

Figure 3-5. Frequency of Pressure Fluctuations of Sand/Freon System at 1.8 U_{mf} with Average Measured Value Indicated. Ratio of Average Frequency Near Predicted Value of 2.9.

Figure 3-6. Frequency of Pressure Fluctuations of Cork/Air System at $2.0 U_{mf}$ with Average Measured Value Indicated.

Figure 3-7. Frequency of Pressure Fluctuations of Sand/Freon System at 2.0 U_{mf} with Average Measured Value Indicated. Ratio of Average Frequency Near Predicted Value of 2.9.

can be used, reliably, to predict the performance of an exactly scaled second model even in extreme cases (i.e. tungsten/water vs. expanded polystyrene/air). It therefore seems likely that the fluidization hydrodynamics of a hot fluidized bed can be modeled in a scaled cold bed using the same dimensionless groups.

References

1. Scharff, M.F., et al., of JAYCOR. Final report to U.S. Department of Energy, Morgantown Energy Research Center, Contract EY-77-C-21-8156, "Project to Provide an Experimental Plan for the Merc 6' x 6' Fluidized Bed Cold Test Model," (1977).

SOLIDS MOVEMENT: DIRECT VELOCITY MEASUREMENT

A small glass recirculating-bed system has been designed and constructed as a facility for testing the feasibility of direct measurement of particle velocity. The background for this method was given in detail in the previous report (DOE/ET/13152-1).

A description of the flow system and the design calculations are given in the paper, "Flow Analysis of a Recirculating Fluidized Bed System," by L.R. Wood. This paper is included in Appendix 1.

PLUME MODEL FOR COAL COMBUSTION

Completion of the development of a "Plume Model" for coal combustion have been supported on this grant, for the behavior of the burning coal since the main effect of solids circulation is to reduce plume height and produce a more uniform bed. This work has been done by D. Park, and a draft of a paper giving the details of the model is included in Appendix 2.

APPENDIX 1

FLOW ANALYSIS OF A REGIRCULATING FLUIDIZED BED SYSTEM

At some point in any fluidized bed process, material must be transferred from the bed. In general, this can be done using a standpipe as an overflow point at the top of the bed, or beneath the bed as a drain. Assembly of the present recirculating system requires a pressure drop calculation for the three components of the system, the standpipe, the valve, and the bed itself. In the following pages, these components will be considered separately, then a description of the present system will be given.

Standpipe Flow Regimes

Extensive work on standpipe flow has been conducted by Leung.¹⁻³ Each regime is characterized by a given voidage and limits on the slip velocity. The regimes are PACFLO, TRANPACFLO, DENFLO, and LEANFLO. PACFLO occurs when the solids have a negative slip velocity and voidage equivalent to a vibrated packed bed. TRANPACFLO occurs when the slip velocity is positive but less than U_{mf}/ϵ_{mf} and the voidage is less than or equal to that seen at minimum fluidization. DENFLO is a state of fluidized flow where the slip velocity is greater than U_{mf}/ϵ_{mf} and voidage greater than ϵ_{mf} . LEANFLO shares these same definitions. The distinction between them is that at a given solids flowrate, the change in gas velocity with change in voidage is positive for DENFLO and negative for LEANFLO. A more practical differentiation is that DENFLO is characteristic of a high solids flux whereas LEANFLO is seen at small solids flux. The pressure drop equations for each regime are given below:

PACFLO

$$\Delta P = -z[k_1 |U_{SL}| + k_2 |U_{SL}|^2] \quad (1)$$

where

$$k_1 = \frac{150(1-\epsilon)^2}{(\phi_s d_p \epsilon)^2}, \quad k_2 = \frac{1.75 \rho_g (1-\epsilon)}{\phi_s d_p \epsilon}$$

and

$$U_{SL} = -U_g/\epsilon + w_s/\rho_s(1-\epsilon)A$$

TRANPACFLO

$$\Delta P = +z[k_1|U_{SL}| + k_2|U_{SL}|^2] \quad (2)$$

DENFLO, LEANFLO

$$\Delta P = \int_0^z \rho_s(1-\epsilon) y dz - \frac{w_s^2}{\rho_s(1-\epsilon)A^2} \Big|_0^z + \int_0^z \frac{\partial f w_s}{\rho_s(1-\epsilon)A^2 D} dz \quad (3)$$

Looking at these equations, it is clear that the non-fluidizing regimes are described by a modified Ergun equation. The fluidized regimes are a summation of the gravity pressure drop, acceleration pressure drop, and a wall friction term. In an industrial system, wall friction is small and has been neglected in the present case also. This is a reasonable assumption for almost spherical particles in a glass system.

Valve Analysis

According to Leung,¹ the valve can be modeled as an orifice in the line. In the latest paper, noted above, the orifice equations are derived assuming that voidage is uniform and equivalent to ϵ_{mf} ; that gravity field, inter-particle forces, and friction can be neglected; and that inertia of the gas is negligible.

If the solids are of Geldhart's type B,⁴ then the second assumption is reasonable. The last is also reasonable for all but extreme gas flows. The uniform voidage assumption seems sufficient since the length to diameter ratio of the valve is close to unity.

The result of the derivation yields the following design equations.

$$\Delta P_o = P_2 - P_3 = \pm [k_3 |U_{SL}| + k_4 |U_{SL}|^2] \quad (4)$$

where

$$k_3 = \frac{k_1 D_o A}{4A_o} \quad , \quad k_4 = \frac{k_2 D_o A^2}{24A_o^2}$$

and

$$\Delta P_o > 0 \text{ for } U_{SL} < 0, \quad \Delta P_o < 0 \text{ for } U_{SL} > 0$$

$$w_s = w_{so} \pm C_D (A_o/A) (\partial \rho_s (1-\epsilon_{mf}) |\Delta P_o|)^{1/2} \quad (5)$$

where $0.5 \leq C_D \leq 0.65$

and positive sign for positive ΔP_o

The term w_{so} is the flow of solids seen when there is no pressure drop through the valve. When the ratio of the orifice diameter to the particle diameter is less than 160, a correction must be applied. For most laboratory systems, this becomes important, and was first studied by Rausch in 1948. In a later paper by Zenz,⁵ the phenomena is explained in relation to hopper flow. In the text written by Levenspiel,⁶ an equation is given that allows calculation of w_{so} .

$$w_{so} = 0.161(d_{or}/d_p)^{2.746} C_w C_o g^{1/2} \rho_s (1-\epsilon) d_p^{2.5} (\tan \theta_R)^{-1/2} \quad (6)$$

Results of this equation for the present system are given in Figure 1.

Fluidized Bed Section

For the bed itself, the pressure drop is given by the gravity pressure drop, $\rho_s (1-\epsilon)gz$. The value of ϵ here must be calculated according to work done by Geldhart.⁷ In this calculation, a trial and error solution must be carried out to find the height of the bed at a given factor of U_{mf} and the minimum height at U_{mf} . Results of this very long trial and error are given in Figure 2.

There is an assumption of uniform voidage lurking beneath the development of Geldhart's equations. This is not a bad assumption when the bed height to diameter is less than two. Above this value, slugging may create problems.

Flow Criteria

The present system is shown in Figure 3. It has been specified that the bed should operate at approximately $1.5 U_{mf}$. The design has been done for a velocity range of $1.3 - 2.1 U_{mf}$, corresponding to a voidage of 0.460 to 0.481. The flow enters the bed at the bottom, causing the standpipe to be an overflow for the bed. The flow criteria for the system is,

$$\Delta P_{\text{standpipe}} + \Delta P_{\text{valve}} \geq \Delta P_{\text{bed}}$$

$$(P_2 - P_1) + (P_3 - P_2) \geq \Delta P_{\text{bed}}$$

or

Figure 1: Critical solids flow rate vs valve opening

Figure 2: Bed expansion vs multiple of U_{mf} .

Figure 3: Experimental recirculation system

Figure 4: Solids flow vs air flow for 46.0% bed voidage

Figure 5: Solids flow vs air flow for 48.1% bed voidage

$$\Delta P - \Delta P_o \geq \Delta P_{bed} \quad (7)$$

When the appropriate equations are chosen for each standpipe regime, the flow criteria are as given below:

PACFLO

$$2.8631 \times 10^5 \frac{g}{cm^2} - \left[175.56 \frac{f}{A_o} |U_{SL}| + .23487 \frac{f}{A_o^2} |U_{SL}|^2 \right] \geq 4.2104 \times 10^5 (1 - \epsilon_B) \quad (8)$$

TRANPACFLO

$$2.3878 \times 10^5 + \left[175.56 \frac{f}{A_o} |U_{SL}| + .23497 \frac{f}{A_o^2} |U_{SL}|^2 \right] \geq 4.2104 \times 10^5 (1 - \epsilon_B) \quad (9)$$

DENFLO/LEANFLO

$$4.2104 \times 10^5 (1 - \epsilon) - w_s^2 / 189.60 (1 - \epsilon) + \left[175.56 \frac{f}{A_o} |U_{SL}| + .23497 \frac{f}{A_o^2} |U_{SL}|^2 \right] \geq 4.2104 \times 10^5 (1 - \epsilon_B) \quad (10)$$

with w_s given by the valve equation, (5).

The bed material has the properties listed below:

Table 1. Bed Material Properties

Material:	Carbon steel shot
Size:	$d_p \approx 350$ micron, $200 \leq d_p \leq 500$ micron
Density:	7.87 g/cm^3
Sphericity:	0.8
Gas:	Air at 300 K, $\rho = 1.2 \text{ kg/m}^3$, $\mu = 1.85 \times 10^{-5} \text{ Pa}\cdot\text{s}$

For the case of PACFLO in the standpipe, no solution can be found that does not give physically impossible flowrates. For the TRANPACFLO case, no solution can be found that does not violate the assumptions of the regime (U_{SL} positive). Only for DENFLO can a solution be found. The operating diagrams are shown in Figures 4 and 5. Standpipe voidage up to 60% has been considered. Beyond this, solids flow exceeds a physically reasonable value. (At greater than 60% voidage, solids must flow at greater than 14,000 grams/sec. This means the standpipe volume must empty itself in less than 45 milliseconds.) In the following table, the actual values are given.

Table 2. DENFLO Calculations

Input Gas Flow cc/sec	Bed	Voidage Standpipe	% Valve Opening	Solid Flow g/sec
575	0.46	0.45	100	1,825
879	0.46	0.50	100	4,288
879	0.46	0.50	70	3,972
879	0.46	0.50	50	3,984
879	0.46	0.50	30	3,909
2,731	0.46	0.60	100	13,520
2,731	0.46	0.60	70	12,820
2,731	0.56	0.60	50	12,890
2,731	0.46	0.60	30	12,640
1,123	0.481	0.50	100	4,134
1,123	0.481	0.50	70	3,889
1,123	0.481	0.50	50	3,889
1,123	0.481	0.50	30	3,850
2,975	0.481	0.60	100	13,370
2,975	0.481	0.60	70	12,740
2,975	0.481	0.60	50	12,790
2,975	0.481	0.60	30	12,580

For the present system, LEANFLO is not possible because of the physical arrangement of the system.

Additional Data

Table 3. Valve Area vs. Fractional Opening

<u>f</u>	<u>Area, cm²</u>
1.0	4.9087
.7	3.5795
.5	3.3572
.3	2.4176

Table 4. Bed Air Inputs

Minimum Velocity 31.7 cm/s with
 Make-up Air at $1.3 U_{mf}$ of 181 cc/s
 and at $2.1 U_{mf}$ of 772 cc/s

The standpipe must be a 25 mm OD glass tube. The valve is a two inch section of tygon tubing of 25 mm diameter and a pinch clamp. The dimensions of the valve must be as stated, since the corrections, w_{SO} , are calculated using the stated values.

Nomenclature

A	-	Standpipe cross-sectional area
A_o	-	Valve cross-sectional area
C_w, C_o, C_D	-	Drag coefficients
D_o, d_{or}, d_p	-	Diameter, valve, orifice, particle
f	-	Fraction of open valve diameter
$\Delta P, \Delta P_o$	-	Pressure drop in standpipe, valve
U_{mf}	-	Minimum fluidization velocity, positive up, cm/s
U_g	-	Gas velocity, positive up, cm/s
w_s	-	Solid mass velocity, positive down, g/s
w_{so}	-	Critical flow at $\Delta P_o = 0$
z	-	Height of the regime considered, 54.6 cm
$\epsilon, \epsilon_{mf}, \epsilon_B$	-	Voidage, at minimum fluidization, of bed
ρ_g	-	Gas density
ρ_s	-	Solid bulk density
ϕ_s	-	Sphericity
θ_R	-	Angle of repose for the solids

References

- ¹Leung and Jones, "Flow of Gas-Solid Mixtures in Standpipes. A Review,"
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APPENDIX 2

MODELING OF FLUIDIZED BED COMBUSTORS

At present, the United States is studying and evaluating the atmospheric fluidized bed combustor (AFBC) as an alternative to the pulverized coal burner. Even a "small" AFBC (250 mw) would be enormous by chemical industry standards, consisting of 4 or 5 fluidized beds filled to a depth of about 1 m with limestone particles, and fluidized with air at 2 to 4 meters per second. For good distribution of coal one class of design visualizes coal entering through about a hundred feed points at the bottom of the bed.

A properly designed atmospheric fluidized bed combustor should:

- minimize the loss of unburned coal,
- minimize the release of pollutants: sulfur dioxide, nitrogen oxides and particulates,
- operate with a minimum of excess air, and
- maximize the utilization of limestone for sulfur dioxide retention.

These are competing requirements, and to predict the tradeoffs involved in different AFBC designs requires the use of some sort of model for the operation. In order to model the AFBC we have to understand many different kinds of physical and chemical processes occurring in the AFBC: flow of gases and solids in the bed, devolatilization of coal, elutriation of unburned char and limestone particles, attrition of solids in the bed, gas-solid reactions and gas-gas reactions within the bed and above the bed. Any generally model which tries to incorporate all these factors will be extremely complex. So a judicious choice of simplifications are needed if we are to develop a working model.

There are three main aspects to the operation of the AFBC: the combustion of coal, sulfur retention with limestone, and the fluidization itself. Each of these is quite a subject in itself but the problem is complicated

because these factors are interrelated. Nevertheless, the first thing we have to investigate is the fluidization in the bed because combustion and sulfur retention are processes occurring in the fluidized bed. The AFBC has unique features not found in other application of fluidized beds. First, high gas velocities, essential in keeping the size of the bed within reason, requires the use of large particles in the bed. At the same time the immersed tubes in the bed restricts the growth of bubbles therein. Thus the bed is in either the slow bubble regime or in turbulent-churning flow but certainly not in the fast bubble regime for which theories of fluidization were developed and used extensively in the past.

Next, there is so much more limestone than coal in the bed that the physical behavior of the bed is insensitive to the hold-up of coal, and is dependent primarily on the limestone and its products. This fact allows us to decouple coal combustion from the fluid dynamics of the bed.

Sulfur retention in the AFBC occurs primarily through the chemical reaction between calcined limestone (CaO) and sulfur dioxide, one of many combustion products of coal. Thus we have to know how and where in the bed does sulfur dioxide evolve in order to model its reaction with calcined limestone. And as the retention of sulfur dioxide consumes only a minute fraction of the inlet oxygen we can treat the sulfur retention after dealing with the combustion of coal, which dictates the sulfur dioxide formation. In this paper our interest is limited to the combustion of coal; sulfur retention is not treated here.

MODELS IN THE LITERATURE

Let us examine the critical assumptions underlying the models in the

open literature, listed in Table A2-1, and see whether they are useful for representing and modeling the large utility scale AFBC's.

Gas Flow in the Bed

As mentioned earlier AFBC's are operated in either the slow bubble regime or in turbulent-churning regime, but not in fast bubble regime. But as Table A2-1 shows, most model builders have used the two-phase theory developed for fluidized beds in the fast bubble regime with their severe gas bypassing. Thus, these two phase gas flow models, originally developed for catalytic reactors with their fine particles, are not applicable for coal combustion with their large particles, high gas velocities, and small bubbles.

Mixing of Solids

Good solid mixing occurs in fluidized beds and thus the assumption of uniform solid concentration in the bed has been widely used. But AFBC's are so large in cross-sectional area that we cannot expect this, and the problem is how many feed points are necessary for relatively uniform solid concentration (see Highley and Merrick). For sulfur retention we do not have serious problems because residence time of limestone is in the order of hour. But combustion of coal is much faster, so we have to compare the time scales involved. For a 0.9 m x 0.9 m particles take at least 15 seconds to be well mixed and the time for complete combustion of a char particle in the bed is of the order of minutes, about 5 minutes for 3000 μm particles and about a minute for 300 μm particles, and for still smaller particles the time is in terms of seconds. Therefore, if feed coal contains a significant amount of very small particles, say 50 μm , the AFBC is far

Table A2-1. Models on FBC.

Model Number and Authors	A	B	C	D	E	F
1. Avedesian & Davidson (1973)	I	a	I	c	I	No
2. Campbell & Davidson (1975)	I	a	I	c	I	No
3. Gibbs (1975)	I	a	I	a	I	Yes
4. Gordon & Amundson (1976)	I	a	I	c	I&III	No
5. Gordon & co-workers (1978)	I	a	I	c	III	No
6. Chen & Saxena (1977)	II	c	I	c	I	Yes
7. Horio & co-workers (1977)	I	a	I	c	II	Yes
8. Horio & Wen (1978)						
Rajan & co-workers (1978)	III	b	II	a	II	Yes
9. Baron & co-workers (1977)	I&IV	a&d	I	b	II	Yes
Baron & co-workers (1978)	I	a	I	c	I	Yes
10. Fan & co-workers (1979)	I	a	III	c	I	No
11. Park & co-workers (1979)	IV	d	I	d	II	Yes

See Table 2 for description.

E. Kinetics of Solid Coal Combustion

I. Film diffusion controlling.

II. Both film diffusion and surface reaction influence the rate, but homogeneous oxidation of CO to CO₂ is assumed to be very fast.

III. Rate of CO oxidation is comparable to that of other reactions.

F. Is Elutriation Considered?

from the perfect mixer for the combustion of coal if it has one feed point per square meter. But one advantage of using AFBC is that it can burn coal of relatively large size, and using of relatively large coal particles saves crushing cost. Thus for the combustion of char one feed point per square meter is not that bad. But the problem is complicated because of the volatiles in coal that evolve once the coal enters in the hot bed. If the dispersion of feed coal is much faster than the devolatilization then each coal particle will have its own tiny diffusion flame as long as its volatiles evolve and burn. However, if devolatilization is faster than the dispersion from a coal feed point, the volatiles will combine to form a large plume of unburned gas rising from the entry port. Since this plume can contain as much as 50% of the heating value of the fuel it is not something which we can or should ignore.

Measurements show that the time for devolatilization is less than one second for 200 μm particles and less than ten seconds for 3000 μm particles. Therefore we can see that one feed point per square meter is not enough if we want to do without the plume while burning coals not much bigger than 3000 μm . Some models simply ignore volatiles and some others assumed uniform release of volatiles throughout the bed.

Carbon Utilization

Unburned fuel can leave the AFBC in three ways:

- 1) with the spent bed material,
- 2) as unburned gases, CO and others, and
- 3) as elutriated char particles.

Results of Babcock and Wilcox studies and others show that there is negligible

char in the exit solids, and that unburned CO and other gases will rapidly burn to near completion above the bed. Thus the only significant contribution to unburned fuel is the elutriation of fine char particles. Thus models which ignore the elutriation loss are useless for correlating combustion efficiency with parameters like gas velocity or size of feed coal.

PLUME MODEL

As we discussed so far, none of the models are satisfactory. Here we propose a plume model which accounts for the general features of the AFBC and can explain some phenomena observed.

Assumptions of the Plume Model

The major assumptions of the plume model are as follows.

1. Plug flow of gas: For beds of large particles, high minimum fluidizing velocity ($0.5 \sim 3$ m/s), and small bubbles one cannot use any of the two phase models for gas flow with their cloud-surrounded bubbles. We have in fact a slow bubble bed or a churning-turbulent bed for which the best simple representation is straight plug flow.
2. Devolatilization of entering coal particles is instantaneous and occurs at the feed entry ports to give plumes of unburned combustible gases which rise through the bed at the same velocity as the surrounding gases.

This assumption represents the AFBC in which coal is fed from below by entrained air in discrete entry ports. We show later that other methods of coal feed can be treated as special

Figure A2-1 Sketch of the Main Features of Plume Model for the Bottom Coal Feed AFBC

cases of the entry port analysis, and that the entry port analysis with its plume of rising volatiles is the most complex situation to treat.

3. Combustion occurs at the boundary between volatiles and the surrounding oxygen-rich gas. This follows diffusion flame kinetics, the whole plume acting very much like a burning gas flame, with oxygen and combustibles moving to the plume boundary with an effective diffusion (or dispersion) coefficient which can be found by experiment.
4. The volatiles in the plume which pass through the bed unburned mix with oxygen-rich gas just above the bed to burn rapidly to completion.
5. The devolatilized solids are well mixed throughout the bed. They burn everywhere in the bed except in the plume. They maintain a constant density, the ash peeling off as it is formed. Hence reaction of the char follows shrinking particle kinetics and is first order with respect to oxygen.
6. The bed is uniform in temperature.
7. Large char particles shrink as they burn in the bed, and at any bed condition a distinct size exists above which no elutriation occurs. Below this size the chance of elutriation increases progressively as particles shrink.

Figure A2-1 sketches the main features of the bottom-fed coal burning AFBC.

Combustion of Char, Volatiles and Oxygen in the Bed

As the main form of fuel loss in an FBC is the elutriation of unburned

$$W(\bar{R}_i) = \frac{F_0(\bar{R}_0)\Delta R_i - W(\bar{R}_{i+1})R(\bar{R}_{i+1})\left(\frac{\Delta R_i}{\Delta R_{i+1}}\right)}{\frac{F_1}{W}\Delta R_i - R(\bar{R}_i) + \kappa(\bar{R}_i)\Delta R_i - \frac{3W(R_i)}{\bar{R}_i}R(\bar{R}_i)\Delta R_i} \quad (1)$$

the total weight of char in the bed by

$$W = \sum_i W(\bar{R}_i) \quad ,$$

The carryover of unburned char by

$$F_2 = \sum_i W(\bar{R}_i) (\bar{R}_i) \quad , \quad (2)$$

and the conversion of char in the bed by

$$X_b = 1 - F_2/F_{0,\text{total}} \quad (3)$$

Now put $F_1 = 0$, meaning that no unburned carbon leaves in the overflow solids. Also, notice that all the values are for elutriable particles. Particles that are too big to be elutriated can be incorporated into the largest size fraction with adjusted mass except $F_{0,\text{total}}$ which represents total char feed rate before the adjustment.

Now to calculate F_2 , we need the elutriation and shrinkage rate of particles. For $\kappa(\bar{R}_i)$, we use the expression of Colakyan¹⁸, good for any size of solids R_i

$$\kappa(R_i) = \kappa_0 \left(1 - \frac{u_t(R_i)}{u_o}\right)^2 \quad (4)$$

and the shrinking rate

$$R_{(R)} \equiv \frac{dR}{dt} \quad R = \frac{-12k_0 C_{O_2}^*}{\rho_c \xi_c}$$

For the particles of interest, i.e., of elutriable size, we have to account for both the film diffusion and surface reaction. Thus

$$k'' = \frac{1}{\frac{1}{2k_g} + \frac{1}{k_s}},$$

and by adopting the expression of Field¹⁹ we also have

$$k_s = 5.96 \times 10^2 T_s \exp \left[\frac{1.494 \times 10^6}{RT_s} \right].$$

The particle surface temperature is given by the energy balance

$$h(T_s - T_b) + \epsilon \sigma (T_s^4 - T_b^4) = -r_c (-\Delta H_r).$$

More than one stable state surface temperature is possible under some circumstances but the chance of realizing the high temperature stable state is very small for the particles of interest (see Appendix A). Calculations also show that k'' does not vary much with different size of particles and oxygen concentration. Therefore, we can use the simple space average concentration for oxygen without introducing much error. Notice that we assumed first order reaction rate with respect to oxygen.

At this point we must look at the combustion of volatiles in order to know the average oxygen concentration $C_{O_2}^*$ encountered by the well mixed

solids within the bed. Let A denote combustible volatiles and B oxygen. Then

Here ν is a stoichiometric coefficient characterizing the composition of volatiles and it is assumed that there is no change in moles by the reaction.

Neglecting axial dispersion we obtain the following equations by mass balances for oxygen and combustible volatiles:

INSIDE THE PLUME

$$\frac{C_A}{z} = \frac{D_A}{u_o r} \frac{\partial}{\partial r} \left(r \frac{\partial C_A}{\partial r} \right) \quad (6)$$

$$\text{I.C.} \quad C_A = C_{A_0} \text{ for } 0 \leq r < L_0$$

$$\text{B.C.} \quad \frac{\partial C_A}{\partial r} = 0 \text{ at } r = 0$$

for all $z > 0$

$$C_A = 0 \text{ at } r = L_i(z)$$

OUTSIDE THE PLUME

$$\frac{C_B}{\partial z} = \frac{D_B}{u_o r} \frac{\partial}{\partial r} \left(r \frac{\partial C_B}{\partial r} \right) - \frac{k}{u_o} C_B \quad (7)$$

$$\text{I.C.} \quad C_B = C_{B_0} \text{ for } L_0 < r \leq L_2$$

$$\text{B.C.} \quad C_B = 0 \text{ at } r = L_i(z)$$

for all $z > 0$

$$\frac{\partial C_B}{\partial r} = 0 \text{ at } r = L_2$$

Figure A2-2 Typical Concentration Profile of Volatiles (A) and Oxygen (B) in the Vicinity of the Plume

ON THE PLUME SURFACE

$$D_A \frac{\partial C_A}{\partial r} \bigg|_{r = L_1(z)^-} = -v D_B \frac{\partial C_B}{\partial r} \bigg|_{r = L_2(z)^+} \quad (8)$$

Here we assume that volatiles evolve from a circular region on the bottom of the bed centered on the coal feed port. The radius of that circle, L_0 , is determined by the amount of volatiles that move upward at the same velocity as the fluidizing air. The radius of the plume at any elevation z is set by Eqs. (6), (7) and (8). Typical concentration profiles are sketched in Figure A2-2.

We also assume that the gases rise in plug flow with a certain lateral dispersion which is surprisingly small and not too different from that found in a packed bed. We adopt this picture as a result of the experimental studies by Jovanović²⁰ on the spread of a plume of tracer gas. In essence, he found that the time averaged concentration profiles showed a large amount of lateral dispersion. However, the true instantaneous dispersion of gas out from the axis of the meandering plume was much smaller. And it is this instantaneous coefficient which is to be used in the above equations for D_A and D_B .

We can simplify the problem by assuming that $D_A = D_B$ because small scale eddy diffusion and not molecular diffusion is the main mechanism of the gas dispersion. Applying the classical treatment of Burke and Schumann²¹, we treat the concentration of combustible volatiles as a negative oxygen concentration. Oxygen and combustible volatiles have to move to the flame front (surface of the plume) stoichiometrically so that Eq. (8) is satisfied. Then C_A that satisfies Eq. (6) can be applied to Eq. (7) if we put $C_A = -v C_B$. Dropping subscripts A and B, we can now combine Eqs. (6) and (7).

Letting $\hat{C} = \frac{C}{C_0}$, $\hat{z} = \frac{z}{H}$, and $\hat{r} = \frac{r}{L_2}$, we obtain the final equation

$$\frac{\partial \hat{C}}{\partial \hat{z}} = \left(\frac{HD}{u_0 L_2^2} \right) \frac{\partial}{\partial \hat{r}} \left(\hat{r} \frac{\partial \hat{C}}{\partial \hat{r}} \right) - \frac{Hk}{u_0} \hat{C} I \quad (9)$$

where

$$I = \begin{cases} 0 & \text{if } \hat{C} \leq 0 \\ 1 & \text{if } \hat{C} > 0 \end{cases}$$

$$\text{I.C.} \quad \hat{C} = \begin{cases} -vC_{A0}/C_0 & \text{for } 0 < \hat{r} < S_0 \\ C_{B0}/C_0 & \text{for } S_0 < \hat{r} \leq 1 \end{cases}$$

$$\text{where} \quad S_0 = L_0/L_2$$

$$\text{B.C.} \quad \frac{\partial \hat{C}}{\partial \hat{r}} = 0 \text{ at } \hat{r} = 0 \text{ and } \hat{r} = 1 \text{ for all } \hat{z} > 0$$

In this notation the surface of the plume is where $\hat{C} = 0$ and the new concentration profile is as shown in Figure A2-3. Eq. (9) can be solved to give the mean oxygen concentration in the bed without much difficulty using numerical methods with given initial and boundary conditions.

The value of k in Eq. (9) represents the local oxygen disappearance rate constant and is constant throughout the bed since we have assumed mixed flow for solids in the bed. But notice that the value of k is determined by the given bed conditions and feed coal and this value must be found by trial and error.

By solving Eq. (9) with an assumed value for k we can obtain

Figure A2-3 Concentration Profile of
Oxygen (+) and Volatiles (-)

$$X_v = 1 - 2 \int_0^{\hat{L}_1(1)} \hat{C} \hat{r} \, d\hat{r} / (S_0^2 \gamma C_{A_0}/C_0) \quad (10)$$

$$X_{O_2} = 1 - 2 \int_0^1 \hat{C} \hat{r} \, d\hat{r} / [(1-S_0^2) C_{B_0}/C_0] \quad (11)$$

For convenience define X_p as follows

$$X_p \equiv 1 - 2 \int_0^1 \hat{C} \hat{r} \, d\hat{r} / [(1-S_0^2) C_{B_0}/C_0] \quad (12)$$

Then from an oxygen balance the conversion of char in the bed is

$$X_b = \frac{1}{1 + \left[\frac{1-X_v}{X_{O_2} - X_p X_v} \right] \left[\frac{1}{1+E} - X_p \right]} \quad (13)$$

where E is the excess air ratio.

This X_b has to be the same as the one from Eq. (3). Thus using Eqs. (1), (2), (3), (9), (10), (11), (12) and (13) we can find ka dn conversions by a trial and error procedure.

Combustion Above the Bed

On entering the freeboard region, unburned combustible gases rapidly mix with oxygen giving essentially complete combustion within a short distance above the top of the bed. This results in an abrupt temperature jump. Although much sparser than in the bed there are many particles moving in all directions in the freeboard. Thus there is the combustion of particles though its contribution is much smaller than that of in the bed owing to much smaller char hold-up in the freeboard. But the elutriation rate constant is,

by the definition, the measure of the probability of a particle to be elutriated out of the bed against to stay in the bed and in the freeboard. Therefore using Eq. (1) we already lumped the combustion of char in the freeboard, which burst into the freeboard and eventually settle down back to the bed, with the combustion of the particles in the bed. Rigorously speaking, Eq. (1) and others have to be modified to account for this but it is not necessary as char combustion in the freeboard has minor effect on the combustion in the bed. The remaining problem is how to treat those elutriating particles. But it is evident that combustion of elutriating particles comes into the picture only when the freeboard is high enough so that there is region where practically no down coming particle is present. But it is unlikely that any AFBC will have such high freeboard, thus let us forget about the further combustion of elutriating particles farther up in the freeboard, which at any rate is insignificant.

The overall carbon efficiency is given by

$$X_{ce} = 1 - (1 - X_b)f_1 \quad (14)$$

where f_1 is the ratio of fixed carbon to total carbon in the feed coal.

Effects of Transport Air

So far we assumed that no air enters into the bed with coal through the feed port. But actually air is used as carrier gas for the transport of coal into the bed and its amount can be significant to the total fluidizing air.

Entering the bed coal will release volatiles which will burn as coal particles are dispersed in the transport air. Thus if the transport air is sufficient in amount to burn all the volatiles all the volatiles will combust

near the feed port and there will be no plume of unburned volatiles. But we expect that usually the amount of transport is not sufficient for the complete combustion of volatiles (around 10% of total air in B & W pilot plant operation). In this case all the oxygen in the transport air will be quickly consumed near the feed port and there will be adjustment in the value of v while the size of plume increases by the amount of the transport air. Also, we have to modify Eqs. (10), (11) and (12) accordingly.

$$X_v = 1 - \frac{2[1-f_2-f_3(1+E_x)]}{1-f_2} \int_0^{S_0} \hat{C} \hat{r} \hat{dr} / (S_0^2 \gamma C_{A_0}/C_0) \quad (10')$$

$$X_{O_2} = 1 - 2(1-f_3) \int_0^{S_0} \hat{C} \hat{r} \hat{dr} / (1-S_0^2) (C_{B_0}/C_0) \quad (11')$$

$$X_p = 1 - 2(1-f_3) \int_0^1 \hat{C} \hat{r} \hat{dr} / (1-S_0^2) (C_{B_0}/C_0) \quad (12')$$

where f_3 is the ratio of transport air to the total air to the bed.

Results of Calculations

For the typical coal of properties shown in Table A2-2 let us first compare the "plume" model with the "plumeless" model. By "plumeless" model we mean the extreme case where the particles of feed coal disperse uniformly throughout the bed before any devolatilization, and the release of the volatiles and their subsequent combustion is complete before the onset of the char combustion. In the "plumeless" model, the average oxygen concentration in the bed is given as

$$C_{O_2}^* = \frac{C_{B_0} [1 - (1-X_b)f_2]}{(1+E_x) \ln \left[\frac{x}{(1-X_b)f_2 + E_x} \right]} \quad (15)$$

Table A-2-2. Properties of Coal Used for Calculation

Total carbon:	73% (moisture free)
Fixed carbon:	49% (moisture free)
Volatile matter:	43.7% (moisture free)
Volume of volatiles:	$0.25 \text{ m}^3/\text{kg}$ of coal
Stoichiometric air for the combustion of coal:	317 mol air/kg coal
f_2 :	0.576
γ calculated using the above values:	2.53
Heating value of coal:	31000 kJ/kg coal (dry basis)
Density of char:	1000 kg/m^3
Maximum (possible) temper- ature rise from combus- tion of volatiles:	944.4k

where f_2 is the ratio of stoichiometric oxygen for char combustion to that for total coal combustion.

We can now find X_b from Eq. (15) together with Eqs. (1), (2) and (3), although the procedure again involves trial and error. This X_b of the "plumeless" model is always smaller than that of the "plume" model because the complete combustion of volatiles in the plumeless bed results in a lower average oxygen concentration in the bed. Calculations show that the maximum difference in the carbon efficiency is about 5% for the most sensitive coal feed size, and that in most practical cases the difference is much smaller. This is a useful finding for it now allows us to rapidly estimate char efficiency without needing to solve the plume model, Eq. (9). All we need to do is solve the very much simpler Eqs. (1), (2), (3) and (15).

Figures A2-4, A2-5 and A2-6 show the effect of different superficial gas velocities, excess air and elutriation rate constants on char combustion efficiency in the bed which is fed different sizes of typical coal particles whose properties are given in Table A2-2. All the calculations in this paper are performed for a bed temperature of 1116 K. These figures all show that char conversion is nearly complete for both small feed particles and large feed particles, while intermediate size particles have lower conversion.

The reason for this somewhat unexpected behavior is as follows. For very small feed particles reaction to completion is more likely than elutriation, hence high combustion efficiency. For intermediate sized particles, elutriation is more likely than reaction, hence lowered conversion of solids. But for very large feed particles most of the solids are consumed before the particles get down to elutriable size, hence high combustion efficiency once again. This prediction shows that there is one critical intermediate size of

feed solids which gives minimum char efficiency in the bed.

Figure A2-7 shows that one parameter, $HD/u_o L_2^2$, determines the conversion of volatiles in the bed. X_v is insensitive to X_b and excess air level which affect the average oxygen concentration in the bed. Notice that for sufficiently large values of $HD/u_o L_2^2$ volatiles burn completely in the bed and the plume closes and disappears within the bed.

The adiabatic temperature jump in the freeboard is shown in Figure A2-8.

Discussion and Conclusion

The plume model is developed here to represent a tube filled AFBC with large particles in which air-entrained coal enter in a number of feed ports from below. It assumes that the volatiles are rapidly released from the coal at the feed entry ports to rise as plumes of combustible vapors.

Graphs have been prepared (Figures A2-4-A2-9) to display the predictions of this model for narrow size cuts of a typical coal feed (Table A2-2). For a feed of wide size distribution use these single size charts and properly sum.

Figures A2-4, A2-5 and A2-6 show that the lowest carbon efficiency always comes with an intermediate size of feed coal, not with very large or very small feed sizes. Thus the coal feed to the AFBC should try to avoid this critical size.

The plume behavior, whether it breaks the surface of the bed, the temperature jump above the bed, concentration variations across the bed, etc., are all governed by one dimensionless group $HD/u_o L_2^2$ which accounts primarily on the spacing of feed ports in the bed. As an example, Figure A2-7 shows that doubling the distance between feed ports lower the volatiles burned in the bed from 90% to about 50%.

Figure A2-4 Conversion of Char in the Bed as a Function of Size of Coal Feed and Elutriation Constant

Figure A2-A1

Figure A2-A2

Figure A2-A3

Figure A2-A4

Figure A2-5. Conversion of Char in the Bed as a Function of Size of Coal Feed and Superficial Gas Velocity.

For a given coal feed, the carbon efficiency depends only on superficial gas velocity in the bed, the excess air, and elutriation rate constant.

The special case and simplification of this model views the coal as being uniformly distributed all over the bed before the volatiles are released. Here analysis is very much simpler (Eqs. (1), (2), (3) and (15) only) not involving plumes and no volatiles leaving the bed. This plumeless model should reasonably represent AFBC using large feed particles, introduced across the top of the bed and then rapidly mixed by large scale convective flow of solids.

The analysis of this paper shows that even in beds with plumes the simpler plumeless model can be used with negligible error to calculate carbon efficiency in the bed. However, for volatile efficiency, temperature jumps, composition variations across the bed, the complete plume model must be used.

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APPENDIX A. TEMPERATURE OF BURNING COAL CHAR PARTICLES

It is well known that more than one stable temperature is possible for burning coal particles (Field, et al). But the question is at which temperature, if there be more than one, are these burning coal particles in the AFBC.

For the modeling of AFBC's we need a simple but useful representation of this process. As our concern is for char particles of elutriable size we concentrate on the combustion of small particles, <1mm. In addition, if we assume that only the homogeneous reaction $\text{CO} + 1/2\text{O}_2 \rightarrow \text{CO}_2$ occurs away from the surface of the particle the energy and mass balance is greatly simplified.

In general we have three different regimes of behavior (see Figures A2-A1).

Regime A: Only one high temperature steady state. Here the particle is much hotter than the rest of the bed, the reaction rate constant is very high and consequently the overall combustion rate is controlled by film diffusion.

Regime B: Here we notice three possible char temperatures X, Y, and Z. At X the combustion is slow and reaction controlled, at Z the particle is hot, reaction is fast and diffusion controlled. Point Y is unstable and physically unrealizable. If a particle enters regime B at high temperature it moves to the high temperature point Z. If it comes from low temperature it settles to point X.

Regime C: Only one steady state. Here the particle temperature is close to that of the bed, the reaction rate constant is low and the overall combustion rate is slow and chemical reaction controlled.

The regime in which a char particle finds itself depends on the ambient temperature, oxygen concentration, size of particle and the reactivity of the particle. Since a particle shrinks and moves about the bed into widely different oxygen environments (high at bottom of bed, low at the top, zero in

Figure A2-6 Conversion of Char in the Bed as a Function of Size of Coal Feed and Excess Air

the plume) it experiences frequent changes in regime during its lifetime. This can be conveniently displayed as follows.

First, adopting the approach of Field, et al. we evaluate the overall combustion rate of individual particles in a bed at 1116K. This is shown in Figure A2-A2 as a function of oxygen concentration and particle size.

Figure A2-A3 is a more useful representation of these results. It shows that regime A can only be attained for large particles at high oxygen concentration, while regime C can be found for the entire range of particle size at low oxygen concentrations.

Since coal particles enter the bed cold those smaller than 1 mm have no chance of attaining the higher temperature point of regime B (point 2). However particles larger than 1 mm could well enter regime A (see point 1 of Figure A2-A3). These remain at high temperature even when they move to point 2. But they are quenched if they move further into regime C (point 3). In addition, once quenched they cannot go back to the higher temperature even when they return to regime B (from point 4 to point 5). A very few particles may be "lucky" enough to stay in regime B throughout their lifetimes (from point 1 to point 6), but most do not because the lifetime of particles in the bed is longer than their mixing time, and the fraction of bed in regime C is larger than the fraction in B and A.

Thus with a bed at 1116 K most of the elutriable particles are at the low stable temperature, close to that of the bed.

But the picture is different at higher bed temperature as shown in Figure A2-A4. Here particles in regime C (point 1) have a good chance of entering regime A (point 2) thereby experiencing a temperature jump. They then maintain this high temperature even if they enter regime B (point 3),

Figure A2-7 Conversion of Volatiles in the Bed

Figure A2-8 Temperature Jump
In the Bed

Figure A2-9

as long as they don't return to regime C. Since the fraction of elutriable solids in regime A is significant the change of having hot elutriable solids is not negligible in this bed.

LIST OF SYMBOLS

C, C_A, C_B	Gas concentration [kg-mol/m ³]
$C_{O_2}^*$	Average oxygen concentration in the bed [kg-mol/m ³]
D, D_A, D_B	Gas dispersion coefficient in the bed [m ³ gas/m bed]
E_x	Excess air ratio
$F_0(\bar{R}_i)$	Char feed rate of size $\bar{R}_i$ [kg/s]
F_1	Char outflow rate [kg/s]
$F_2(\bar{R}_i)$	Elutriation rate of char of size $\bar{R}_i$ [kg/s]
f_1	Ratio of fixed carbon to total carbon in coal
f_2	Ratio of oxygen needed for char combustion to total coal combustion
f_3	Ratio of transport air to
H	Height of the bed [m]
h	Convective heat transfer coefficient [kJ/m ² Ks]
k	Reaction rate constant for the disappearance of oxygen in the bed [1/m ³ bed·s]
k''	Reaction rate constant for char $\frac{\text{kg-mol carbon}}{\text{kg-mol } O_2/\text{m}^3} \frac{1}{\text{m}^2 \text{ char surface} \cdot \text{s}}$
k_s	Rate constant for surface reaction [1/m ² of char surface·s]
$L_0, L_1(Z)$	Radius of plume; at the bottom of the bed, at height Z, respectively [m]
L_2	Radius of modlue for one feed coal point [m]
R	Radius of shrinking coal particle
$\bar{R}_i$	Average radius of particles in i^{th} size interval [m]
r	Radial distance from the coal feed point [m]

R	Rate of particle shrinkage [m/s]
S_0	L_0/L_2
T_b	Temperature of the bed [K]
T_s	Surface temperature of a burning char particle [K]
u_0	Superficial gas velocity in the bed [m/s]
u_t	Terminal velocity of an elutriating particle [m/s]
$W(\bar{R}_i)$	Char weight of i^{th} size interval [kg]
X_b	Conversion of char in bed
X_{ce}	Overall carbon efficiency
X_{O_2}	Conversion of oxygen in bed
X_v	Conversion of volatiles in bed
z	Vertical distance in the bed [m]

Greek Letters

ϵ	Emissivity of char particles
ν	Stoichiometry coefficient
κ_0	An elutriation constant, [s^{-1}], see Eq. (4)
$\kappa(\bar{R}_i)$	Elutriation rate constant for particles of size $\bar{R}_i$ [s^{-1}]
ξ_c	Mass fraction of carbon in char
ρ_c	Density of char [kg/m^3]
σ	Boltzmann coefficient

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