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Time-Dependent Behavior of Large-Particle Fluidized-Bed Hydrodynamics

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1. Summary

Research efforts during the period 3/1/81 through 8/31/81 were focused on two distinct areas. A novel description of the bubble rise velocity in a fluidized bed was developed which alleviates many of the limitations of commonly used design equations, such as that of Davidson and Harrison. The parameters of our new bubble velocity equation were evaluated using reported experimental results of bubble flow.

In other work supported by this grant, a successful approach to feedback control of the fluidized state has been demonstrated experimentally in our laboratory. The technique is practical in that it makes use of a simple differential pressure measurement and is carried out via a small on-line computer. A self-tuning regulator algorithm has been implemented which will adjust to various bed conditions.

2. Objectives

Progress during this period was related to the following objectives of the overall research program:

- a) to develop fundamental mathematical models of the dynamic behavior of the fluidization regime,
- b) to verify these models given experimental data,
- c) to develop on-line methods for estimation of fluidization parameters, and
- d) to investigate application of these models to the experimental feedback control of the fluidized state.

3. Research Activity

3.1 An Improved Description of Bubble Rise Velocity

During this period and primarily in the summer months of 1981, the Principal Investigator worked independently to extend the results of Weimer and develop a novel, more general, predictive equation for the bubble rise velocity in a fluidized bed. Such an equation filled an important gap in our overall mathematical model of the fluidization process. An overview of our findings is presented here and details are available in an appendix to this report.

Two-phase bubble models of fluidized beds require a predictive equation for the absolute upward velocity of the bubble phase. The widely used bubble velocity equation of Davidson and Harrison (1963) was derived based on the concept of a single bubble rising in an infinite medium. Hence, this equation cannot account for interference among rising bubbles reducing their velocity. In addition, there is no accounting for non-uniform flow and bubble volume fraction or for the bubble flow being less than excess flow above that at minimum fluidization. The inability to describe these effects causes significant deviations between actual bubble velocity and that predicted by the Davidson and Harrison equation.

An improved bubble velocity equation for bubbling fluidized beds has been developed. It has the form

$$\bar{u}_B = C_o \langle u_{Bo} \rangle + 0.71 \left[\frac{g \bar{d}_b (\rho_D - \rho_B) (1 - \langle \delta_B \rangle)}{\rho_D} \right]^{1/2}$$

Nomenclature for the above is:

- u_B : absolute bubble velocity
- C_o : distribution coefficient
- u_{Bo} : superficial velocity of bubble phase
- g : gravitational acceleration
- d_b : bubble diameter
- ρ_D : emulsion phase density
- ρ_B : bubble phase density
- δ_B : volume fraction bubbles
- $-$: weighted mean value
- $\langle \rangle$: average of cross-sectional area

This equation accounts for a number of factors which contribute to deviation of actual absolute bubble velocity from that predicted by the equation of Davidson and Harrison including

- interference between adjacent bubbles reducing bubble velocity
- influence of non-uniform bubble distribution either to increase (for bubbles concentrated near the axial centerline) or to decrease (for bubbles closer to the wall) the absolute bubble velocity, and
- reduction in bubble phase velocity which results from bubble flow being less than the excess flow above minimum fluidization.

The distribution coefficient, C_o , has been evaluated from available experimental data of a number of reported investigations of bubble flow. It has been shown that calculated values agree well with qualitative aspects of

its significance. In particular, bubbles are concentrated near the wall close to the inlet gas distributor and the local value of C_0 is less than one. As coalescence takes place and bubbles tend toward the axial centerline, the distribution coefficient approaches 1.5. Farther up the bed, C_0 can exceed 1.5 and tends to level off.

It can be concluded that this new bubble velocity equation represents a significant improvement over the widely applied equation of Davidson and Harrison. Also, since the new equation can be applied to accommodate either local (at any axial position) or bed volume-average bubble properties, it is immediately useful for interpreting experimental data and for inclusion in comprehensive models of fluidized bed hydrodynamics.

Finally, the improved bubble velocity equation can be compared to the equation of Davidson and Harrison. Under appropriate restrictive assumptions, the former reduces to the latter.

3.2 Feedback Control of the Fluidized State

During the first two months of this period, Dale C. Gyure completed his Master's thesis under the supervision of the Principal Investigator. The hydrodynamic state of the fluidized bed was quantized through use of on-line time series modeling of a simple differential pressure measurement.

The stochastic behavior of the differential pressure was modeled in time series form, and it was found that model parameters varied linearly with the extent of fluidization above the minimum. Through use of a recursive on-line estimation scheme, fitted parameter values were employed

in two different feedback control schemes to manipulate the gas feed rate to the bed. The first control scheme was a simple integral-only algorithm. This approach was quite successful; however, it did not have the adaptive properties which are desirable for changing bed conditions. Consequently, a self-tuning scheme was developed and shown to be feasible. Details of this work are presented in an appendix to this report.

Our progress in this area is important since we have demonstrated a practical technique for using information on time-dependent behavior of hydrodynamics to close a feedback loop on the fluidized state. This has not been accomplished before.

4. Projected Activity

In the period 9/1/81 through 2/28/82, the following research is planned:

- a) Our hydrodynamic modeling approach will be applied to the transition from bubbling to slug flow in order to round out our mathematical model to the upper end of the churn-turbulent flow regime.
- b) A computer simulation of a low-pressure, fluidized bed char gasifier will be developed integrating many of our previous modeling results in hydrodynamics. Michael D. Michonski, an M.S. student supported by this grant, will be working in this area. Simulation study results will not be forthcoming until the last six-month period of this grant however.

5. Publications

1. Gyure, Dale C., "A Stochastic Approach to Fluidized State Control,"
M. S. Thesis, University of Colorado, 1981.

**Appendix 1: An Improved Description of Bubble Rise Velocity --
Detailed Results**

Background

In applying bubble models to characterize fluidized bed hydrodynamics, one requires a predictive equation for the absolute upward velocity of the bubble phase in the bed. The commonly accepted equation is

$$u_B = u_o - u_{mf} + 0.711 (g d_b)^{1/2} \quad (1)$$

which was presented originally by Nicklin (1962) for gas-liquid systems and later by Davidson and Harrison (1963) for gas-solid fluidized beds. It is observed from the above equation that the absolute bubble velocity, u_B , is the sum of the natural rising velocity, $0.711 (g d_b)^{1/2}$, and the upward velocity of the dense phase gas between bubbles, $(u_o - u_{mf})$.

The form of the natural rising velocity equation,

$$u_{br} = K_b (g d_b)^{1/2} ,$$

results from application of the steady state mechanical energy balance to describe the rate of rise of a single spherical bubble in a low viscosity medium. The coefficient K_b is generally given the value 0.711 as obtained from experimental observations of Davies and Taylor (1950). However, a number of other investigators [Davidson et al. (1959), Donsi et al. (1972), Harrison and Leung (1962), Park et al. (1969), Rippin (1959), and Toei et al. (1966)] have assigned values to this coefficient ranging from 0.545 to 0.914. Nonetheless, Equation 1 has been used consistently in all recent modeling efforts for describing bubble phase velocity.

Davidson et al. (1977) have noted that the validity of Equation 1 is doubtful. The natural rising bubble velocity term, $0.711 (g d_b)^{1/2}$, can only be justified for the case of a single bubble and u_o approaching u_{mf} (from above). In addition, they note that the term, $(u_o - u_{mf})$, is questionable

since it has been obtained by analogy with slug flow. Furthermore, they enumerate reasons for the actual bubble velocity being either smaller or larger than that predicted by Equation 1. These are:

I. actual bubble velocity less than predicted

- interference between adjacent bubbles reducing velocity as for a slug rising more slowly in a smaller tube (Turner, 1966)
- it has been shown by Lockett et al. (1967) that percolation of gas can occur from bubble to bubble through the dense phase resulting in bubble flow being less than excess above minimum fluidization
- bubbles may be in regions of the bed where upward flow of gas is less than the cross-sectional average

II. actual bubble velocity greater than predicted

- bubble velocity can be augmented by coalescence: as a trailing bubble rises into the wake of a bubble above, the following bubble tends to rise faster (Harrison and Leung, 1962)
- bubbles may be located in the bed where the upward flow of gas is greater than the cross-sectional average

In this work, we will derive a bubble phase velocity equation, similar to (1), but from concepts quite different from those applied by Davidson and Harrison. This improved equation can account for most of the factors cited above.

Theoretical Development

Here, we consider a two-phase fluidized-bed system: a continuous dense phase consisting of solid particles surrounded by an interstitial gas and a dispersed bubble phase consisting of solids-free bubbles which develop and rise through the dense phase. Our primary concern is the unidimensional, upward, absolute velocity of the bubble phase.

Zuber and Findlay (1965) derived a general expression for the unidimensional velocity of the dispersed phase flowing through the continuous phase in any two-phase system. In terms of the fluidized-bed system, the weighted-mean, bubble-phase velocity is

$$\bar{u}_B = C_0 \langle j \rangle + \bar{V}_{Bj} \quad (2)$$

where the mixture volumetric flux density is given by

$$\langle j \rangle = \frac{Q_B}{A_T} = \langle u_{B0} \rangle \quad (3)$$

for a stagnant dense phase, and the distribution coefficient, C_0 , is

$$C_0 = \frac{\langle \delta_B j \rangle}{\langle \delta_B \rangle \langle j \rangle} = \frac{\frac{1}{A_T} \int \delta_B j dA_T}{\left[\frac{1}{A_T} \int \delta_B dA_T \right] \left[\frac{1}{A_T} \int j dA_T \right]} \quad (4)$$

This coefficient represents the effect of non-uniform bubble flow and volume fraction. The bubble-phase, weighted-mean, drift velocity term, $\bar{V}_{Bj}$, accounts for the effect of local relative velocity.

The bubble-phase drift velocity, V_{Bj} , is related to the local relative velocity between phases by

$$V_{Bj} = (1 - \delta_B) u_r \quad (5)$$

A relation for the bubble velocity relative to the dense-phase velocity, u_r , can be obtained from a mixture model developed by Ishii and Zuber (1979). This mixture model has been shown to describe bubbly, droplet, and particulate flow for gas-liquid, liquid-liquid, and solid-liquid systems in the Stokes, Viscous, Newton's, Distorted Fluid Particle, and Churn-Turbulent flow regimes by a uniform method. Here, we consider application to a gas-solid, bubbling, fluidized-bed system characterized by a bubble phase and a dense phase. Drag similarity criteria and mixture properties are introduced. The analysis takes into account the effect of both the presence and motion of other bubbles. Characteristics of the bubbles are reflected in the single-bubble rise velocity.

Ishii and Zuber (1979) have developed from fundamental concepts a relative velocity relationship for the dispersed phase in a two-phase, multi-particle system:

$$u_r |u_r| = \frac{8}{3} \frac{r_B}{C_D \rho_D} (\rho_D - \rho_B) g (1 - \delta_B) \quad (6)$$

A bubbling fluidized bed is characterized by turbulence and bubble coalescence. As the diameter of a bubble increases, the wake and bubble boundary layer can overlap due to the formation of a large wake region. A bubble can influence both the surrounding dense phase and other bubbles directly. Also, bubbles can become entrained in the wakes of other bubbles. This fluid behavior of large, deformed bubbles characterizes the Churn-Turbulent flow regime. These spherical-cap bubbles tend to manifest flat or inverted bases. Surface tension forces are unimportant, and the drag coefficient of a single, spherical-cap bubble in an infinite medium approaches a value of

$$C_{D_\infty} = \frac{8}{3} \quad (7)$$

independent of bubble size.

In considering a multi-*fluid*-particle (here, the particles are bubbles), it is necessary to account for the effect of other particles (bubbles) on the flow field. Consequently, the drag coefficient, C_D , for a multi-particle system differs from C_{D_∞} . In order to model the increased drag effect due to the presence of neighboring particles, Ishii and Zuber (1979) assumed a simple similarity criterion between single- and multi-particle systems based on an appropriate Reynolds number and derived the drag coefficient in a multi-particle (fluid or solid) system for the Churn-Turbulent flow regime. It is

$$C_D = \frac{8}{3} (1 - \delta_B)^2 \quad (8)$$

Hence, the relative velocity can be obtained from Equations 6 and 8 as

$$u_r = \left[\frac{\frac{8}{3} r_B (\rho_D - \rho_B) g}{\frac{8}{3} (1 - \delta_B) \rho_D} \right]^{1/2} = 0.71 \left[\frac{g d_b (\rho_D - \rho_B)}{\rho_D (1 - \delta_B)} \right]^{1/2} \quad (9)$$

The drift velocity of the bubble phase is now derived from Equations 5 and 9 to be

$$v_{Bj} = 0.71 \left[\frac{g d_b (\rho_D - \rho_B) (1 - \delta_B)}{\rho_D} \right]^{1/2} \quad (10)$$

It can also be shown that the *weighted-mean* drift velocity is given by

$$\bar{v}_{Bj} = 0.71 \left[\frac{g \bar{d}_b (\rho_D - \rho_B) (1 - \langle \delta_B \rangle)}{\rho_D} \right]^{1/2} \quad (11)$$

Finally, the improved bubble velocity equation is obtained from Equations 2, 3 and 10 as

$$\bar{u}_B = C_0 \langle u_{B0} \rangle + 0.71 \left[\frac{g \bar{d}_b (\rho_D - \rho_B) (1 - \langle \delta_B \rangle)}{\rho_D} \right]^{1/2} \quad (12)$$

in terms of average and weighted-mean values. A more detailed derivation of the above result is given elsewhere (Weimer, 1980).

Experimental Evaluation of Distribution Coefficient

The distribution coefficient can be interpreted as a local value dependent on such factors as axial distance up the bed and gas velocity, or it can be viewed as an average value which is constant throughout the bulk of the fluidized bed and range of gas velocities considered--for a particular flow regime. Zuber and Findlay (1965) conclude that available experimental data show the distribution coefficient to remain constant for established flow profiles. However, at the time of their study, direct local measurements of bubble flow had not been made. We will show here that evaluations of such reported local measurements made from bubbling, gas-solid, fluidized beds indicate that the distribution coefficient varies with bed height and diameter, and inlet gas velocity.

In either case, the numerical value associated with the distribution coefficient indicates whether the bubble volume fraction is uniform or non-uniform over the bed cross-section. For a non-uniform coefficient, the value also indicates whether the bubbles are concentrated near the wall or near the centerline. In particular, if the bubbles are uniformly distributed across the bed cross-section,

$$\delta_{B_w} = \delta_{B_c} = \langle \delta_B \rangle \quad (13)$$

and

$$C_o = 1 \quad (14)$$

If the volume fraction of bubbles is greater toward the axial centerline, that is, if

$$\delta_{B_c} > \delta_{B_w} \quad (15)$$

then

$$C_o > 1 \quad (16)$$

and vice versa.

Zuber and Findlay (1965) have presented an extensive investigation of average volumetric concentration in two-phase flow systems. Their analysis of the distribution coefficient is presented in much detail, and this is recommended reading for those interested.

Local Values

The development of miniature capacitance probes [Werther and Molerus (1973a), Werther (1974a)] has contributed to a better understanding of local bubble behavior in gas-solid fluidized beds. These devices have made possible the investigation of radial bubble volume fraction and flow distributions [Werther and Molerus (1973a,b), Werther (1974a,b,c, 1976), Burgess and Calderbank (1975), Calderbank et al. (1976), and Wittman et al. (1978)]. From the data now available, we can evaluate the distribution coefficient, C_0 , directly via the definition given in Equation 4. Each integral of Equation 4, representing an average over the bed cross-section, may be determined from

$$\langle F \rangle = \frac{1}{A_T} \int_{A_T} F dA_T = \frac{2\pi}{\pi r_{\max}^2} \int_0^{r_{\max}} (rF) dr \quad (17)$$

for

$$F = \delta_B, j, \text{ and } \delta_B j$$

These integrals are evaluated numerically by quadrature using data taken from the above-referenced investigations.

Calderbank et al. (1976) have used miniature capacitance probes to make local measurements of bubble volume fraction, δ_B , absolute bubble rise velocity, u_B , and bubble diameter, d_b . Their bed was 0.155 m in diameter, and during operation the expanded height averaged 0.55 m. The fluidized solid had a minimum fluidization velocity of 0.011 m/s. The gas superficial velocity was maintained at 0.031 m/s. Local measurements were taken at elevations above

the gas distributor of 0.3, 0.4 and 0.5 m, and along a radius of the bed to provide a complete traverse of the bubble dispersion. Their data are summarized in Figure 1. In viewing these profiles, it is apparent that bubbles are located closer to the axial centerline, and hence we should expect a distribution coefficient greater than unity. This is shown quantitatively in Table 1 where the three integrals of Equation 4 have been evaluated numerically by quadrature. The distribution coefficients obtained, 1.36, 1.92 and 1.96, for increasing bed heights indicate that bubbles move toward the axial centerline as they travel up the bed.

Wittman et al. (1978) have made local measurements in a 0.19 m diameter fluidized bed. The collapsed bed height was 0.57 m, and the fluidized particles had a minimum fluidization velocity of 0.161 m/s. The inlet gas superficial velocity was varied in the range of two to four times that at minimum fluidization. Measurements were reported at axial distances of 0.055, 0.165, 0.372 and 0.565 m from the inlet gas distributor. Once again radial measurements were made, and the results are shown as radial profiles in Figure 2. It appears that bubbles are concentrated near the wall at heights of 0.055 m and 0.165 m; whereas, at greater distances up from the gas distributor (0.372 m and 0.565 m), bubbles approach the axial centerline. Calculations summarized in Table 2 yield values of the distribution coefficient from 1.07 to 2.56. The effect of height within the bed is clearly significant as C_0 increases up the bed, that is, as bubbles tend toward the centerline. The effect of superficial gas velocity is not as clear. At a particular bed height, the distribution coefficient *does* vary with changes in gas velocity; however, the effect does not appear to be predictable.

It should be noted that some of our calculated values of the distribution coefficient are much higher than those reported by previous investigators. Zuber and Findlay (1965) found values of C_0 ranging from 0.85 to 1.5 for

Figure 1 Variation of Bubble Phase Volume Fraction and Superficial Bubble Phase Velocity with Axial and Radial Position in the Fluidized Bed as Determined by Calderbank et al. (1976).

Table 1

Summary of Calculations for Determining Distribution
Coefficient, C_o , for Fluidized Bed System
of Calderbank et al. (1976)

<u>Bed Height, z(m)</u>	<u>0.3</u>	<u>0.4</u>	<u>0.5</u>
$\langle j \rangle$ (m/s)	0.903	1.11	1.16
$\langle \delta_B \rangle$	0.0307	0.0354	0.0365
$\langle \delta_B j \rangle$ (m/s)	0.0376	0.0756	0.0830
$C_o = \frac{\langle \delta_B j \rangle}{\langle \delta_B \rangle \langle j \rangle}$	1.36	1.92	1.96

Figure 2 Variation of Bubble Phase Volume Fraction and Superficial Bubble Velocity with Axial and Radial Position in the Fluidized Bed as Determined by Wittman et al. (1978).

Table 2

Summary of Calculations for Determining Distribution Coefficient,
 C_o , for Fluidized Bed System of Wittman et al. (1978)

Bed Height, z(m)	0.055			0.165			0.372		0.565		
Ratio, u_o/u_{mf}	2	3	4	2	3	4	2	3	2	3	4
$\langle j \rangle$ m/s	0.496	1.14	2.32	0.738	2.03	3.87	0.589	1.65	0.723	1.60	3.08
$\langle \delta_B \rangle$	0.0137	0.0251	0.0441	0.0130	0.0291	0.0569	0.00990	0.0241	0.0110	0.0202	0.0404
$\langle \delta_B j \rangle$ (m/s)	0.00730	0.0355	0.114	0.0118	0.0833	0.244	0.00850	0.0603	0.0188	0.825	0.233
$C_o = \frac{\langle \delta_B j \rangle}{\langle \delta_B \rangle \langle j \rangle}$	1.07	1.24	1.11	1.23	1.41	1.11	1.46	1.52	2.36	2.56	1.87

gas-liquid systems. The value of 1.5 corresponds to a pronounced parabolic profile. As shown in Table 2, we have calculated directly values of the local distribution coefficient as high as 2.56. This indicates that bubbles in bubbling gas-solid fluidized beds do not travel up the bed in established flow profiles such as the familiar ones which develop for laminar flow in circular ducts.

Bubbles are known to travel up the bed following preferred paths (Geldart, 1970, 1971), and, in general, these paths approach the centerline. If most bubbles are near the centerline and few are close to the wall, values of the distribution coefficient greater than 1.5 are conceivable. Also, as the rate of coalescence decreases up the bed, a point would be reached at which the distribution coefficient would level off. The calculations summarized in Table 1 corroborate this. At an axial distance of 0.3 m above the distributor, C_0 takes on a value of 1.36. As bubbles continue to move toward the centerline, the coefficient increases to 1.92 at 0.4 m and 1.96 at 0.5 m.

An important question that we have yet to address is: Does the distribution coefficient remain constant for an established flow regime of a given system? As we mentioned earlier, Zuber and Findlay (1965) attest that for fully established and constant flow profiles, the value of C_0 is fixed. From an analysis of the slug flow regime (Weimer, 1980), we would have to agree. However, for the churn-turbulent flow regime, we have shown from reported data that the distribution coefficient may take on values from 1.1 to 2.6. The flow regime is established as churn-turbulent; yet the bubble flow profile is not and local values of the distribution coefficient vary up the bed.

All calculations reported here have resulted in values of the distribution coefficient greater than unity. However, we have been limited in our analysis to axial elevations greater than 0.05 m. Results presented by Werther (1974a,b,c, 1976) and Werther and Molerus (1973a,b) show qualitatively that

the bulk of bubbles close to the inlet gas distributor are adjacent to the wall. This implies a distribution coefficient less than unity. However, the above-cited work does not present enough information to allow direct calculation of the distribution coefficient in this lower region of the bed.

To summarize our findings concerning local values of the distribution coefficient as a function of bed height: close to the inlet gas distributor, bubbles are concentrated near the wall and C_0 takes on values < 1 . As coalescence takes place and the larger bubbles tend toward the centerline, C_0 approaches 1.5. Farther up the bed and in the absence of slug flow, bubbles are not near the wall and larger values of C_0 are attained. Furthermore, the value of C_0 tends to level off at high elevations in the bed. For a transition to slug flow, the distribution coefficient will decrease as bubbles grow and spread out across the bed cross-section (Weimer, 1980).

Average Values

Thus far we have been concerned with local values of the distribution coefficient. These values have been determined from local bubble measurements taken at various axial and radial positions within the fluidized bed. However, data from local bubble measurements are scarce and the experimental technique for obtaining such data is quite involved. This, combined with the fact that no equations are available to predict a priori local values of C_0 , leads us to consider values averaged throughout the bed and over the range of inlet gas velocities considered. Furthermore, average values of the distribution coefficient are readily determined from the more common measurements of visible bubble flow (i.e. superficial bubble phase velocity) and expanded bed height.

The weighted-mean absolute bubble phase velocity is given by Equation 12. We now consider the absolute bubble phase velocity and the superficial bubble phase velocity to be mean values averaged throughout the bed volume for a

particular inlet gas velocity. These mean values are denoted as $\bar{u}_B$ and $\bar{u}_{Bo}$, respectively. The distribution coefficient is interpreted to be constant throughout the bed volume and for all inlet gas velocities considered. This constant value, which represents an average over both the bed volume and all inlet gas velocities, is represented by $\bar{C}_{om}$. Furthermore, Ishii and Zuber (1979) claim that the drift velocity is constant throughout the system (bed volume and inlet gas velocities) for the case of churn-turbulent bubble flow. Hence, it will be denoted as $\bar{V}_{Bjm}$. With these assumptions,

$$\bar{u}_B = \bar{C}_{om} \bar{u}_{Bo} + \bar{V}_{Bjm} \quad (18)$$

Equation 18 is represented by a straight line on a plot of mean absolute bubble velocity, $\bar{u}_B$, versus mean superficial bubble velocity, $\bar{u}_{Bo}$. The slope of this line yields $\bar{C}_{om}$ and the intercept is given by $\bar{V}_{Bjm}$.

Values of the superficial bubble phase velocity can be readily determined at various axial distances above the gas distributor from standard experiments. Averaging these values over the bed volume,

$$\bar{u}_{Bo} = \frac{1}{L_f} \int_0^{L_f} \langle u_{Bo} \rangle dz \quad (19)$$

With a measurement of the expanded height of the fluidized bed, we have the bubble volume fraction averaged throughout the bed for a particular gas flow

$$\bar{\delta}_B = 1 - \frac{L_{mf}}{L_f} = \frac{1}{L_f} \int_0^{L_f} \langle \delta_B \rangle dz \quad (20)$$

The absolute bubble velocity averaged throughout the bed is given by

$$\bar{u}_B = \bar{u}_{Bo} / \bar{\delta}_B \quad (21)$$

This method of determining average values for the distribution coefficient, $\bar{C}_{o_m}$, will be illustrated. It requires experimental data in which simultaneous measurements of visible bubble flow and expanded bed height are reported. This type of data is plentiful.

Geldart (1976) has considered the bed expansion and visible bubble flow for sand-like powders of various mean sizes and collapsed bed depths. The fluidized bed was 0.305 m in diameter and bed heights at minimum fluidization were varied in the range $L_{mf} = 0.10$ to 2.0 m. Here we consider the particular case of particles of mean size $\bar{d}_p = 128 \mu\text{m}$ and inlet superficial gas velocities exceeding minimum fluidization by 0.02, 0.04, and 0.06 m/s. The resulting mean superficial bubble velocity is plotted against the mean absolute bubble velocity in Figure 3. A least squares fit of these data yields an average distribution coefficient of $\bar{C}_{o_m} = 1.39$.

McGrath and Streatfield (1971) considered the fluidization of large particles in shallow gas-fluidized beds. The fluidized bed was rectangular, 0.305 m x 0.156 m, and 0.133 m deep at minimum fluidization. The fluidized particles had a mean diameter of $\bar{d}_p = 1540 \mu\text{m}$ and a minimum fluidization velocity of $u_{mf} = 0.55$ m/s. A plot of mean superficial bubble velocity against mean absolute bubble velocity is shown in Figure 4. A least squares fit of these data yields an average distribution coefficient of $\bar{C}_{o_m} = 1.29$.

This method of determining a value for the distribution coefficient averaged over the bed volume and operating gas velocities has been successfully applied to gas-liquid (Zuber and Findlay, 1965) systems and gas-solid (Staub, 1980) transport lines. It provides a simple means of determining the average distribution coefficient which can be easily applied in modeling.

Values of the average distribution coefficient which we have determined here, $\bar{C}_{o_m} = 1.29$ and 1.39, appear to be reasonable in representing average bubble profiles throughout the bed. Both values are qualitatively representative

Figure 3 Mean Absolute Bubble Velocity Versus Mean Superficial Bubble Velocity for Experimental Data of Geldart (1976).

Figure 4 Mean Absolute Bubble Velocity Versus Mean Superficial Bubble Velocity for Experimental Data of McGrath and Streatfield (1971).

of bubble profiles in shallow fluidized beds where it is improbable that bubbles could travel up the axial centerline in such a short distance. Hence, values of the local distribution coefficient would be expected to be less than $C_0 = 1.50$.

Comparison of Improved Bubble Velocity Equation to Conventionally Applied Bubble Velocity Equation

It is appropriate to compare the improved bubble velocity equation, Equation 12, to the conventionally applied bubble velocity equation, Equation 1.

1. It is apparent that the improved bubble velocity equation predicts an absolute bubble phase velocity less than that given by Equation 1 when

1. interference between adjacent bubbles reduces their velocities (i.e. as the average bubble phase volume fraction, $\langle \delta_B \rangle$, increases, the weighted-mean bubble drift velocity, $\bar{V}_{Bj}$, decreases),
2. the average superficial bubble velocity, $\langle u_{B0} \rangle$, is less than the total inlet gas flow minus that required to incipiently fluidize the bed, and
3. bubbles on the average are closer to the wall (i.e. $C_0 < 1$).

The improved bubble phase velocity equation, Equation 12, predicts a bubble phase velocity which is greater than that given by Equation 1 when bubbles on the average are closer to the bed axial centerline (i.e. $C_0 > 1$), and this effect is greater than the slowing effects of bubble interference and of superficial bubble velocity being less than excess flow above minimum fluidization.

These differences result because the two equations have been derived from different concepts. As discussed earlier, the commonly accepted equation for absolute bubble velocity, Equation 1, has been developed partially from analogy with slug flow and partially by applying the steady-state mechanical energy balance to describe the rate of rise of a single spherical bubble in a low viscosity medium (Davidson and Harrison, 1963). The bubble coefficient,

K_b , was originally inferred from experimental results.

On the other hand, the improved bubble velocity equation, Equation 12, has been developed by considering fundamental transport properties of the two phases, accounting for both the effect of non-uniform flow and bubble volume fraction as well as the effect of the local relative velocity between phases. Furthermore, consideration has been given to the influence of neighboring bubbles on the flow behavior of a characteristic bubble within the flux (i.e. considering multi-particle fluid properties).

A comparison of the two equations is best made by simplifying Equation 12 to obtain the conventionally applied bubble velocity equation of Davidson and Harrison (1963), that is, Equation 1. Assuming bubbles to be distributed evenly across the bed cross-section at any axial location, the distribution coefficient takes on the value

$$C_0 = 1 \quad (22)$$

Assuming the relative velocity of the dispersed bubble phase to be unaffected by the presence of neighboring bubbles

$$u_{r_\infty} = u_r (1 - \delta_B)^{1/2} \quad (\text{Ishii and Zuber, 1979}) \quad (23)$$

From Equation 5

$$u_r = \frac{V_{Bj}}{1 - \delta_B} = 0.71 \left[\frac{gd_b(\rho_D - \rho_B)}{\rho_D(1 - \delta_B)} \right]^{1/2} \quad (24)$$

Combining Equations 23 and 24

$$u_{r_\infty} = 0.71 \left[\frac{gd_b(\rho_D - \rho_B)}{\rho_D} \right]^{1/2} \quad (25)$$

With the dispersed bubble phase assumed to be described by a single bubble in an infinite medium unaffected by the presence of neighboring bubbles

$$v_{Bj} = u_{r_{\infty}} \quad (26)$$

Assuming all flow in excess of that required for minimum fluidization to contribute directly to the formation of bubbles

$$u_{Bo} = u_o - u_{mf} \quad (27)$$

Equations 2, 3, 22, 25, 26, and 27 can be combined to yield

$$\bar{u}_B = u_o - u_{mf} + 0.71 \left(\frac{g \bar{d}_b (\rho_D - \rho_B)}{\rho_D} \right)^{1/2} \quad (28)$$

in terms of weighted mean values. If,

$$\rho_D \gg \rho_B \quad (29)$$

then

$$(\rho_D - \rho_B) / \rho_D \approx 1 \quad (30)$$

and Equation 28 is identical to Equation 1 for the bubble phase.

NOTATION
(SI Units)

A_T	= cross-sectional area of the bed
C_o	= distribution coefficient defined in Equation 4
$\bar{C}_{o_m}$	= distribution coefficient averaged over bed volume and all operating conditions
C_D	= drag coefficient (multi-fluid particle)
d_b	= bubble diameter (volume equivalent sphere)
d_p	= particle diameter
g	= gravitational constant
j	= velocity of the center of volume of dilute and dense phase, i.e. volumetric flux density of the mixture
K_b	= bubble coefficient used in Equation 2
L_f	= expanded bed height
L_{mf}	= bed height at minimum fluidization
Q_B	= bubble volumetric flow
r	= radial distance
r_B	= bubble radius (equivalent spherical volume)
r_{max}	= radius of bed (tube)
u_o	= superficial gas velocity
u_{br}	= bubble natural rising velocity
u_B	= absolute velocity of bubbles, bubble phase
u_{Bo}	= superficial velocity of bubble phase
u_{mf}	= minimum fluidization velocity
u_r	= relative velocity = $u_B - u_D$
V_{Bj}	= drift velocity of bubble phase

Greek Letters

δ_B = volume fraction-bubbles

δ_{Bc}, δ_{Bw} = bubble volume fraction--at axial center-line, at wall

ρ_B, ρ_D = density of bubbles, dense phase

Superscripts

— = denotes an arithmetic-mean value or a weighted-mean value, i.e.

$$\bar{F} = \frac{\langle \delta_B F \rangle}{\langle \delta_B \rangle}$$

~ = denotes integrated average value over bed volume, i.e.

$$\tilde{F} = \frac{1}{L_f} \int_0^{L_f} \langle F(z) \rangle dz$$

Subscripts

∞ = denotes property of a single bubble, unaffected by neighboring bubbles

c = denotes value at cross-section centerline

m = denotes average of value over all operating conditions

Brackets

$\langle \rangle$ = average over cross-sectional area, i.e.

$$\langle F \rangle = \frac{1}{A_T} \int_{A_T} F dA_T$$

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Appendix 2: Feedback Control of the Fluidized State --

Detailed Results

Motivation

In operating industrial fluidized-bed units, the state or extent of the fluidization regime is not controlled in a feedback sense. As bed conditions change, such as particulate loading and size distribution, open-loop adjustments may be made by an operator often to the gas flow to the bed. If these adjustments are not made, the bed may drift into an undesirable fluidization regime or even collapse. Our goal in this work was to control in a feedback sense the extent of fluidization of an experimental bed. Furthermore, it was desired to use only a commonly available process measurement, bed differential pressure, and not more exotic instrumentation, such as radiation densitometers or capacitance probes.

The behavior of the bed differential pressure for a small silica sand laboratory fluidized bed is illustrated in Figure 1 (Robertson, 1979). The differential pressure is recorded as the signal to the control valve for inlet air flow is ramped linearly. The first, smooth portion of the curve represents the packed-bed state, and nonlinearity there is due to the equal percentage characteristic of the valve. At an output just greater than 600, the bed levitates and fluidizes. A threshold in differential pressure is observed at this point. Thereafter, the signal manifests an increasingly noisy character; however, the average value remains relatively constant. This average, smoothed, or "DC" value corresponds to the weight of the material fluidized and may be used in a feedback control scheme to manipulate particulate inflow or outflow. The noise content of the differential pressure is shown in Figure 2 where each trace has roughly the same mean value. The lowest curve on the figure corresponds to a mild, quiescent state of fluidization near minimum fluidization; whereas the highest curve represents fluidization at or near slug flow. At this point, the focus of our previously-stated goal may be shifted. By controlling the nature or extent of "noisiness" of the differential

Figure 1. Fluid-Bed Pressure Drop Versus Air Flow Rate (105 second sweep).

Figure 2. Differential Pressure Fluctuations at Constant Air Flow Rates

pressure signal, it is possible to maintain a desired state of fluidization. The first problem is to quantify the "noisiness."

Experimental Unit

The work reported hereafter was carried out in the Chemical Engineering Laboratories of the University of Colorado utilizing a specific fluidized-bed unit, associated instrumentation, and real-time computer system. A schematic diagram is presented in Figure 3. The bed column is plastic (pmma), 0.3 m O.D. (1 ft) and 3.7 m (12 ft) high. The particulate material was silica sand with a mean size of 300 μm . Superficial gas velocity at minimum fluidization was 0.1 m/s. The computer system is a Data General Nova 3/12 with Computer Products RTP process interface. This system uses a diskette-based operating system and multi-tasking FORTRAN with real-time extensions. Extensive details of both the bed unit and computer system are presented elsewhere (Weimer, 1980, Gyure, 1981).

Stochastic Modeling

Since the differential pressure measurement across the fluidized bed is relatively constant in mean value for different gas flow rates, it is possible to view this signal as a stationary series when it is sampled by an on-line computer. A simple measure of the "noisiness" of this series is an estimate of the variance, and this is shown in Figure 4 for different inlet air flow rates. The results were generated by a recursive estimator which was allowed to compute at steady levels of air flow. The sampling interval was 200 ms. A relatively linear increase in variance with air flow is observed. This recursive estimator does not track rapid changes in the fluidization regime well; in fact, the variance estimate inflates unrealistically during such changes.

In order to provide a more complete model of the differential pressure

Figure 3.
Instrument and
Equipment Sketch

THE PECURSIVE ESTIMATOR :

$$DP = \lambda \cdot (\text{CURRENT PRESSURE DROP}) + (1-\lambda) \cdot DP$$

$$DPS = \lambda \cdot (\text{CURRENT PRESSURE DROP})^2 + (1-\lambda) \cdot DPS$$

$$\text{VARIANCE} = (DPS - DP^2) / (1-\lambda)$$

$$\lambda = \text{SMOOTHING PARAMETER} = 0.2$$

Figure 4. Pressure Drop Variance vs Air Flow and the Fluidized State

and its dependence on flow rate, time series analysis methods were applied (Box and Jenkins, 1976). An appropriate model form for a wide range of air flows is the following autoregressive form

$$DP(t+1) = \beta(1)DP(t) + \beta(2)DP(t-1) + \beta(3)DP(t-2) + e(t+1)$$

By applying a recursive least squares algorithm on-line, parameter estimates for the above model were computed at different inlet air flow rates. The sample interval was again 200 ms. At each flow rate, 2400 iterations (8 min.) were carried out and estimates were logged at 15 second intervals (every 75 iterations). The logged parameter estimates were averaged for each flow rate, and these results are presented in Figure 5. Again, we note the linear dependence of the model parameters on air flow and, consequently, on the fluidized state. Variance estimates and standard errors for the parameters are presented in the table below for two different flow rates.

	Flow = 0.020 Kg/s		Flow = 0.038 Kg/s	
	s^2	S.E.	s^2	S.E.
$\hat{\beta}^*(1)$	0.022	0.15	0.021	0.14
$\hat{\beta}^*(2)$	0.046	0.21	0.050	0.21
$\hat{\beta}^*(3)$	0.022	0.15	0.023	0.15

The uncertainty of the parameter estimates appears to be relatively independent of flow rate, and the $\hat{\beta}^*(2)$ estimate is less certain than the other two.

An additional observation of importance during the tests was that the recursive least squares algorithm handled rapid transitions in flow without the problems encountered with the recursive variance estimator. Evidence of

Figure 5. Average Time Series Model Coefficients vs Air Flow and the Fluidized State

this is presented in Figures 6 and 7 where the response of the $\hat{\beta}(1)$ and $\hat{\beta}(2)$ estimates respectively is shown to a square pulse in air flow. The exponential forgetting factor for the algorithm was 0.975 and estimates are plotted every 15 s. Note also the greater uncertainty of the $\hat{\beta}(2)$ estimate. Also, the FORTRAN code for recursive least squares is quite compact (35 statements), and our small minicomputer could execute it every 200 ms without problem. This scheme provided robust parameter estimates which are directly related to the noise character of the differential pressure signal and to the extent of fluidization in the bed.

Simple Feedback Algorithm

In order to prove feasibility of feedback control of the fluidized state, a simple control scheme was devised and is shown in Figure 8. The concept is to adjust air flow so that a parameter estimate approaches a desired value. The control algorithm chosen was the position-mode floating (or integral-only) controller--noted for slow, sure control action. We did not want to generate a direct proportional response to the rapid variations of the parameter estimate. The floating controller tends to smooth this parameter noise.

This algorithm was tested using both $\hat{\beta}(1)$ and $\hat{\beta}(2)$ as the controlled parameter, and the results are presented in Figures 9 and 10 respectively. The success of this approach in terms of regulating the mean value of the parameter estimate is apparent. We can also report, based on qualitative observations, that the fluidized state changed in lock step with the mean parameter estimate changes. Equations for the control algorithm are presented on the figures. The air flow signal varies more in Figure 10 than in Figure 9: this is due again to the greater uncertainty of the $\hat{\beta}(2)$ estimate. Response of the controller to an unknown disturbance is shown in Figure 11. At time ≈ 16 minutes, the parameter estimate drops and the flow adjusts to bring it back into line.

Figure 6. Open Loop Autoregressive Parameter Response to Changes in the Fluidized State. ($\hat{B}(1)$ shown)

Figure 7. Open Loop Autoregressive Parameter Response to Changes in the Fluidized State. ($-\hat{\beta}(2)$ shown)

FIGURE 8 Simple Feedback Control Scheme.

$$\text{ALGORITHM : } E(i) = \beta_{SP} - \hat{\beta}(1)_i$$

$$\%_{\text{OPEN}} = 0.08 \left[\sum_i E(i) \right] + \%_{\text{BIAS}}$$

Figure 9. Fluidized State Response to Autoregressive Time Series Parameter Set Point Changes. Control Based on $\hat{\beta}(1)$.

$$\text{ALGORITHM : } E(i) = \beta_{SP} - \hat{\beta}^*(2)_i$$

$$\%_{\text{OPEN}} = 0.08 \left[\sum_i E(i) \right] + \%_{\text{BIAS}}$$

Figure 10. Fluidized State Response to Autoregressive Time Series Parameter Set Point Changes. Control Based on $\hat{\beta}(2)$.

$$\text{ALGORITHM : } E(i) = \beta_{sp} - \hat{\beta}(1)_i$$

$$\%_{\text{OPEN}} = 0.0446 \left[\sum_i E(i) \right] + \%_{\text{BIAS}}$$

Figure 11. Fluidized State Response to Autoregressive Moving Average Time Series Parameter Set Point Changes. Control Based on $\hat{\beta}(1)$.

The success of this simple, floating algorithm prompted us to consider a more sophisticated approach involving adaptive control.

Adaptive Algorithms

Given that control decisions from an adaptive algorithm would be less frequent than those from the simple floating controller, we implemented a high-performance slave controller for flow regulation. This was accomplished with a Dahlin algorithm, and results of step changes in flow setpoint are shown in Figure 12.

Figure 5 suggests a static linear relationship between the parameter estimates and the air flow rate. If it is possible to identify the slope and intercept of a model

$$\overline{\beta(1)} = \overline{m}^* \overline{\text{FLOW}} + \overline{b}^*$$

and track these values as bed conditions gradually change, the model may be inverted to yield a proportional control law for flow rate setpoint:

$$\text{FLOW}_{\text{sp}} = \frac{\beta(1)_{\text{sp}} - \overline{b}^*}{\overline{m}^*}$$

Here, parameter "noisiness" is not that much of a concern since $\overline{m}^*$ and $\overline{b}^*$ are obtained from recursive estimation of the model involving smoothed values, $\overline{\beta(1)}$ and $\overline{\text{FLOW}}$. This adaptive scheme is shown in Figure 13. Note the difference in sampling intervals. Performance of this algorithm is depicted in Figures 14 and 15. Controller response appears slower for a setpoint change in $\beta(1)$ than for the simple floating algorithm. Reasonable values of $\overline{m}^*$ and $\overline{b}^*$ are estimated; yet, their covariance matrix grows large during steady operation. This is especially true for the slope parameter ($\overline{m}^*$) uncertainty. With reference again to Figure 5, we see why this occurs. For the recursive estimation of $\overline{m}^*$ and $\overline{b}^*$, there is little information available when operating at a

ALGORITHM :

$$\% (t) = 154.2E(t) + 19.7E(t-T) + 44.1E(t-2T) + 40. \left[\sum_{i=3}^{\infty} E(t-iT) \right]$$

PROCESS GAIN = 0.0038 KG/S/%

SAMPLE INTERVAL = 200 MS

DAHLIN TIME CONSTANT = 1.0 S

Figure 12. Dahlin Flow Controller Response to Set Point Changes in Flow. Deadtime Included in the Controller Design.

Adaptive Algorithm Flow Chart

Figure 13.

ALGORITHM : SELF TUNING REGULATOR

$$FLOW_{SP} = \frac{[\beta(1)_{SP} - \hat{b}^*]}{\hat{m}^*}$$

Figure 14. Adaptive Controller Response to Autoregressive Time Series Parameter Set Point Changes

ALGORITHM : SELF TUNING REGULATOR

$$FLOW_{SP} = \left[\beta(1)_{sp} - \hat{b}^* \right] / \hat{m}^*$$

Figure 15. Covariance Matrix Stability and Fluidized State Model Updating

steady flow condition, and the estimates cluster around one point of the graph. Any straight line estimate for such a cluster will be more difficult as time progresses and older information is forgotten by the least squares algorithm. We proposed two remedies for this and implemented one here.

First, estimates of $\hat{m}^*$ and $\hat{b}^*$ could be carried out during abrupt changes in bed behavior only. This should yield adequate information for stable estimation as demonstrated during the setpoint transitions of Figure 15 where the covariance matrix values are noted. A disadvantage of this approach is that gradual changes in bed behavior would be missed until such abrupt changes occur. Another approach, the one which we implemented, is to remove a degree of freedom from the linear model by pinning down the intercept at a fixed value; that is,

$$\hat{p}(1) = \hat{m}^* \text{FLOW} + B$$

where B is known and fixed. Control using this simplified scheme is shown in Figure 16. Flow response is particularly sharp and accurate at the setpoint change, and the variance estimate of $\hat{m}^*$ appears stable and convergent. In practice it would also be wise to update the fixed intercept during abrupt changes in bed operation.

Concluding Remarks

We have demonstrated that it is possible to control the fluidization state in a feedback sense through manipulation of inlet gas flow. The extent or state of fluidization is inferred from the stochastic component of a common differential pressure measurement. On-line estimation of parameters of an autoregressive model for the differential pressure fluctuations showed these parameters to be related to the state of fluidization. Feedback control was implemented using a small, real-time computer system. A simple floating algorithm and two adaptive schemes were tested and were found to control the fluidized bed successfully.

ALGORITHM : SELF TUNING REGULATOR

$$FLOW_{SP} = \left[\beta(1)_{SP} - \beta \right] / m^*$$

Figure 16. An Adaptive Model With One Degree of Freedom Checks Covariance Matrix Growth

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Nomenclature

b	intercept in parameter/flow model
B	fixed intercept
DP	differential pressure fluctuation
DPS	smoothed square of DP
e, E	model error
FLOW	inlet air flow rate
m	slope in parameter/flow model
s^2	variance estimate
S.E.	standard error
t	sample index
%OPEN	valve command signal
%BIAS	bias value for floating algorithm
β	parameter of time series model
λ	smoothing parameter for recursive variance estimator

Subscripts and Superscripts

sp	setpoint
$\hat{\cdot}$, *	estimated quantities
—	smoothed or averaged quantities