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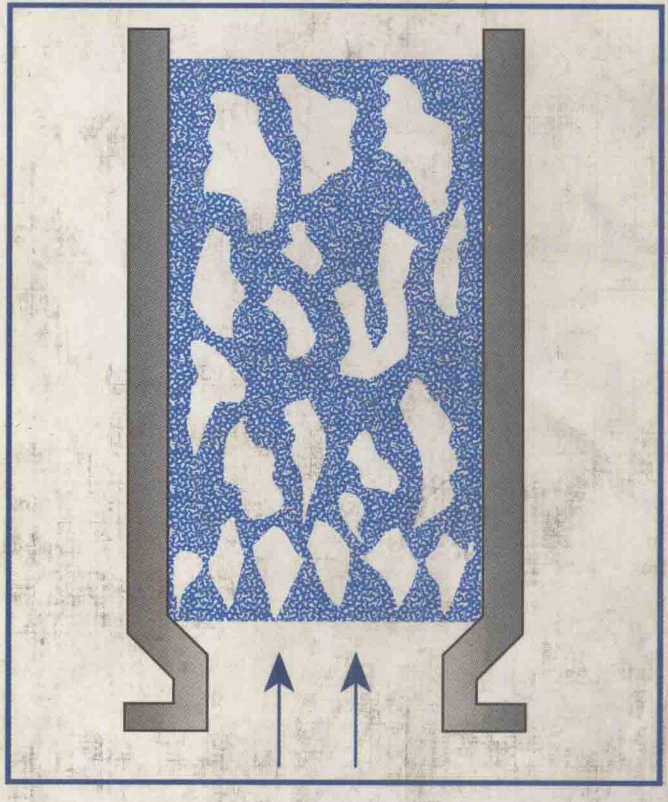
FLUIDIZED PROCESSES: Theory and Practice

Alan W. Weimer, Volume Editor



American Institute of Chemical Engineers

Fluidized Processes: Theory and Practice

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FOREWORD

This volume of the AIChE Symposium Series on Fluidization and Fluid-Particle Systems includes fifteen peer reviewed papers selected from those presented in eight sessions at the Annual Meeting in Los Angeles, November 17-22, 1991. Included is the plenary paper "Agricola Aground: Characterization and Interpretation of Fluidization Phenomena," as presented by Professor John Grace.

The fluid-particle systems covered in this volume include circulating fluidized beds, two-phase gas-solid and gas-liquid bubbling fluidized beds, three-phase magnetically stabilized fluidized beds, and cyclones. Applications are directed towards coal combustion, materials synthesis, and ethanol fermentation processes. Fundamental considerations include circulating fluidized bed hydrodynamics, voidage variations surrounding rising bubbles, kinematics of liquid fluidized beds, particle motion in bubble wakes, dynamic loading on and erosion of heat transfer tubes in fluidized beds, gas distribution, modeling of deterministic chaos, gas-liquid mass transfer, and experimental methods for measuring fluidized bed properties.

I would like to acknowledge the outstanding work of the chairmen and co-chairmen of the sessions for their planning and paper selection, and the reviewers whose valuable comments provided guidance for the publication of this volume.

Alan W. Weimer, volume editor
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CONTENTS

Plenary Paper

AGRICOLA AGROUND: CHARACTERIZATION AND INTERPRETATION OF FLUIDIZATION PHENOMENA

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Experimental results of different fluidization workers commonly show wide scatter, and this is usually ascribed to the complexity of particulate systems. A large measure of imprecision is accepted by the fluidization community. Simple rules of thumb are also often adopted. This paper provides some examples of cases where such simple rules and procedures may be misleading and where they may be useful. Failure to fully characterize particle properties lessens the value of experimental results to other workers. Lack of standardization procedures adds to experimental scatter. Neglected properties of solid particles, collapse tests where the results are affected by the windbox volume and distributor, and variations in procedures used to determine the onset of turbulent fluidization are employed as examples to illustrate these points.

While fluidization may have been practiced as an art by Agricola as early as the 16th century (see Leva, 1959), fluidization as a scientific and engineering endeavour has a life time of less than 50 years. Seldom has a technical field sparked such prodigious research effort, with tens of thousands of research papers, hundreds of millions of dollars and scores of symposia, conferences, workshops, books, etc., devoted to the investigation and exploitation of fluidization phenomena. Studies have taken place in industrial, academic and university environments. Fluidized beds have been investigated by researchers from a wide range of backgrounds - chemical and mechanical engineers, applied mathematicians, physicists, chemists, inventors. Individual projects have varied in scale from tiny tubes to enormous scale industrial reactors tens of metres in diameter.

The approaches followed have varied widely, encompassing, among others:

(a) studies of an experimental, empirical nature;

(b) simple mechanistic modelling work;

(c) complex theoretical approaches.

Studies employing one of the above approaches have, of course, often been pursued in conjunction with another approach.

Given the complexity of the subject and the diversity of background and interests of research workers who have had an interest in fluidization and its applications, it is not surprising that there has been no unanimity on a wide range of topics which are of fundamental importance in achieving understanding of fluidized beds. However, there has been a tendency, perhaps because particulate systems are inherently complicated, for the fluidization community to be mesmerized by a number of simple ideas, experimental techniques and generalities. Commonly these simple results have been very beneficial in advancing our understanding. At the same time, when they pass from useful observation or correlation to the realm of conventional wisdom or even established dogma, we risk oversimplification, lack of proper scientific rigour and obscurring of important factors which could enlarge our understanding of the field.

This paper gives some examples of cases where simple generalizations, which, while greatly useful in their own right, have grown to the status of domineering dogma and where some gentle poking at the underlying ideas or underpinning reveals serious limitations or shortcomings in accepted fluidization "doctrine."

GONE WITH THE WIND: FLUIDIZATION BEYOND THE TERMINAL VELOCITY

Early fluidization textbooks assumed that there was an upper bound on the fluidizing velocity. It was reasoned that if the superficial gas velocity exceeded the terminal settling velocity of individual particles, the bed would immediately be carried over and there would be no bed left. Fortunately, experimenters (not all at any rate) did not construct their equipment in such a way that they could not explore behaviour beyond the terminal settling velocity. If equipment and curiosity had been so limited, we would not now have the circulating fluidized bed and even the practical range of turbulent and bubbling fluidization would be out of bounds. As shown in Figure 1, adapted from earlier work (Grace, 1986), most practical fluidized beds operate beyond the terminal settling velocity, utilizing the limited saturation carrying capacity of the gas and efficient gas-solid separation devices accompanied by solids recycle to maintain concentrated suspensions of particles required to obtain favourable heat transfer, proper gas-solid contacting and high reaction rates.

This simple example illustrates the dangers of adopting too elementary an approach. A simple plausible sounding idea, if allowed to assume the status of doctrine, could have constrained the whole development of the field.

PARTICLE PROPERTIES

Conventional wisdom in the fluidization literature has suggested for many years that it is sufficient to characterize powders used for fluidization by means of only two properties: (i) the surface-volume mean particle diameter, usually estimated from sieve analyses as

$$d_p = 1 / \Sigma(x_i/d_{pi})$$

where x_i is the mass fraction of particles of characteristic size d_{pi}, and (ii) the particle density, ρ_p. As examples, all correlations used to predict minimum fluidization velocities use only d_p and ρ_p as the particle properties. Seldom do commonly used correlations for predicting such properties as bubble properties, heat transfer, reaction conversions, etc. make use of any other particle properties. The familiar powder groups introduced by Geldart (1972, 1973) also consider only d_p and ρ_p.

Despite the tendency to limit attention to these two key variables, there is considerable evidence that a host of other particle properties can play a role, sometimes a very significant role, in determining fluidized bed behaviour and performance. Some of the other properties, areas where these properties are important and key references are listed in Table 1. Many of the influences are not well understood, whereas others are beginning to be studied in detail. Some brief comments, based mainly on the references in Table 1, on some of these variables follow:

Angularity: It has long been understood that rounded particles are usually best from a fluidization point of view. Angular particles tend to flow less readily, and their higher apparent viscosity leads to smaller bubble wakes and hence to decreased solids mixing in the bubbling regime (Rowe et al, 1965). The rheological behaviour of angular particles is more non-Newtonian for smooth particles. There is little evidence of the influence of particle shape at high gas velocities.

Asperities: Real particles are seldom smooth; they tend to have microscopic bumps, ridges and protruberances that can be referred to collectively as asperities. These asperities affect the minimum separation distance between particles and between particles and adjacent solid surfaces. Since van der Waals forces are a strong function of separation distance, asperities can play a significant role in determining interparticle forces, fluidizability and agglomeration (Seville, 1987). Since heat is transferred by conduction through the gas layer separating a particle and heat transfer surface, asperities are also believed (Decker and Glicksman, 1981) to be important in determining heat transfer coefficients from bed to wall.

Surface Composition: The surface composition of particles can influence absorption of gases on the particles, hence affecting interparticle separation and, thereby, van der Waals forces and low velocity fluidization behaviour, especially for fine particles. At high gas velocities there are indications (Louge, 1991) that changing the surface composition can alter the hydrodynamics of circulating beds, possibly due to changes in coefficient of restitution (see below) or different friction on the wall.

Coefficient of Restitution: The coefficient of restitution characterizes the fraction of the kinetic energy which is lost when particles collide. There is evidence (Senior, 1992) that this property can play a major role in determining movement of particles to and from the outer wall in circulating fluidized bed risers.

Binding Agents: Addition of small quantities of "sticky" liquids to a fluidized bed can result in local agglomeration or even total defluidization depending on such factors as the liquid viscosity, wettability of the particles, particle size, superficial gas velocity and means of dispersing the liquid. Similar agglomeration and defluidization can be produced by softening of particle surfaces, for example over critical high temperature ranges for certain materials where surface melting begins to occur.

Addition of Flow Conditioners or Spacers: It is sometimes possible to greatly improve the flowability and/or fluidizability of powders by the addition of fine particles which tend to keep the larger particles apart, hence decreasing van der Waals attractive forces. The effects of the added materials can be quite dramatic, sometimes sufficient to transform group C powders into group A behaviour.

Particle Size Distribution: For many years it has been recognized that wide particle size distributions and the addition of finer fractions are generally beneficial to fluidization behaviour. Among the beneficial effects are reduced bubble sizes, earlier transition to the turbulent regions of fluidization, and increased chemical conversions (Sun and Grace, 1990; Grace and Sun, 1991). These effects seem to be especially important for group A particles, and they have been widely recognized by industry, since loss of fines can result in serious deterioration in reactor performance. Some of the factors contributing to the beneficial influence of wide particle size distributions appear to be increased dense phase expansion, lower effective viscosities and the ability of finer particles to remain suspended within voids.

In our laboratory, we have recently begun to examine the influence of particle size distribution in the fast fluidization regime. Factors of importance appear to be interparticle collisions and the ability of different size fractions to respond to and to influence gas turbulence (Senior, 1982).

Electromagnetic Properties: Interparticle forces can arise from factors other than van der Waals forces such as electromagnetic forces. For example, some significant effects have been found on such items as the minimum bubbling velocity, particle motion and particle entrainment. Imposition of electrical or magnetic fields can cause dramatic hydrodynamic changes.

Electrostatic Effects: Particles can pick up electrostatic charges as they move past surfaces of different electrostatic potential. Relative humidity plays a strong role in dissipating these charges. Electrostatic charges, as well as being a potential hazard, can alter bed hydrodynamics, entrainment and the ability to feed or handle particulate solids.

Limited Self-Agglomeration: There exist some fine powders, such as aerogel catalysts (Chaouki et al, 1985; Li et al, 1990; Lauga et al, 1991), carbonaceous fibrous materials (Brooks and Fitzgerald, 1986) and hard metal particles (Pacek and Nienow, 1990) which exhibit an unusual property. At low gas velocities they exhibit C-type behaviour, but increasing the gas velocity triggers a sudden transition whereby the elementary particles gather into agglomerates consisting of hundreds or even thousands of elementary particles, and these fluidize readily, typically in a group B mode. The agglomerates are quite stable and can persist and even grow larger right into the fast fluidization regime (Li et al, 1990). There is poor understanding of what specific set of properties is needed to assure this kind of behaviour.

Summary Remarks: It is clear that there are many particle properties which can be very important aside from d_p and ρ_p. However, fluidization papers commonly do not bother to report any other properties, giving no indication, for example, of the particle shape or size distribution. For data to be as useful as possible to others in the field, it is essential that full characterization of such properties be reported and that any unusual characteristics be identified.

JET/SPOUT STABILITY

There were vigorous debates some years ago regarding whether or not gas passing into the bottom of a bed of particulate solids from an upward-facing orifice could form stable jets or instead gave way to pulsating void formation right from the orifice itself. X-ray observations for fine particle systems (Rowe et al, 1979) indicated stable jets never formed for all conditions investigated. These observations were at variance with spouted bed experience where stable "spouts" are well known and much studied. A series of experiments reported by Chandnani and Epstein (1986) led to a critical ratio of orifice diameter to mean particle diameter, d_{or}/d_p of about 25, beyond which stable jets are never obtained. By working with fine particle systems, Rowe et al had operated with $d_{or}/d_p > 25$ where they indeed could not find stable jets. Subsequent X-ray observations in the same laboratory (Yates and Cheesman, 1984) as well as a broad range of other data from the literature (see Grace and Lim, 1987) indicate that this simple rule of thumb, i.e. stable jets only form if $d_{or} < 25\ d_p$, applies to a wide range of systems, with minimal effect of such factors as whether or not the orifice is alone, whether or not there is auxiliary gas and the Froude number of the incoming gas stream. This is, incidentally, one of the few instances where a finding from the spouted bed literature has proven to be directly applicable in fluidized bed work.

MINIMUM GAP WIDTH

In early experimental work performed as part of my Ph.D. thesis work (Grace, 1968), I was interested in the influence of vertical tubes and other fixed surfaces on fluidization behaviour. One feature I wished to explore was whether there was a minimum gap width between vertical surfaces below which flow would become fundamentally different. As a simple experiment a wooden ruler was lowered with a perfectly vertical orientation into a perspex column containing magnesite particles of mean diameter 274μm. The ruler was parallel to the front face and at different separation distances from the front face. When the ruler was sufficiently close to the front face, stable channelling occurred in the gap. Gas was drawn from the surrounding particulate phase into the gap between the ruler and the nearby front face of the column; this gas travelled upwards at high velocity carrying widely dispersed particles entrained from below on the edges of the gap.

In the experiments described, the channelling was initiated when the gap width was of the order of 10 times the mean particle size, and it never occurred if the gap width was $20d_p$ or more. The mechanism of channel formation seemed to involve initial blockage of particles as they tried to move into the gap region. Once the gap region had momentarily achieved a significantly higher voidage than the surrounding bed, the local resistance to gas flow was lowered leading to an increased gas velocity there; any particles which do enter the region are then entrained, and the phenomenon of gas channelling becomes established. Some indication of the "through flow velocity" in the gap is given by an analysis carried out in the same thesis (Grace, 1968) showing that a two-dimensional elliptical void of height L and thickness B has a throughflow velocity given by

$$U_{tf} = U_{mf}(L+B)/B$$

Since L>>B for the gap, U_{tf} is large, sufficiently large to keep the gap evacuated once the void is formed.

The key point for our purposes is that a gap of thickness less than about $10d_p$ between two fixed parallel vertical surfaces is such that particles no longer flow freely. Bridging apparently occurs, and instead of acting as a pseudo-continuum, local defluidization can occur as particles try to enter the gap. The magnesite particles used in the experiments described were of typical size distribution and particle shape. Clearly the critical multiple of d_p could be larger for particles of very wide size distribution or of extreme shapes, and it therefore seems wise to recommend gaps of at least $20d_p$, and preferably $50d_p$, in practical applications to avoid the channelling phenomenon.

When fluidization workers ignore this rule of thumb, something which happens with surprising frequency, especially for coarse particles, they pay a price. Some examples are as follows:

(a) "In two-dimensional" columns in which the thickness of the gap is only a small multiple of the mean particle size, the behaviour can be very different from a true two-dimensional column in which the thickness allows for unimpeded flow of dense phase particles, as well as of individual particles and gas.

(b) Arrays of tubes, horizontal or vertical, which are too close together cause particles to become wedged (or "bridge") in gaps, which may in turn cause hot spots, local defluidization and reduced heat transfer.

(c) Finned tubes are sometimes employed in fluidized beds to enhance heat removal. It has been found (e.g. Priebe and Genetti, 1977) that there is a marked deterioration in the effectiveness of the fins if the gap between adjacent fins is less than about $10d_p$.

Presumably particles bridge with smaller gaps, compromising the free movement and renewal of particles which are essential to the maintenance of high heat transfer coefficients.

(d) Square-nosed slugs form in beds of small dimeter, especially with coarse particles. Their occurrence may again be related to the bridging phenomenon.

COLLAPSE TESTS

Since their introduction in the fluidization literature in the 1960's (Rietema, 1967), collapse tests have been widely used by fluidization researchers to characterize dense phase properties. The test has also been adopted in industry (e.g. Brown, 1990) for assessing the relative fluidizability of different catalysts and powders.

The general idea in a collapse test is that one fluidizes a material at some gas velocity in excess of minimum fluidization and then abruptly cuts off the supply of gas to the bed. The descent of the bed surface is then monitored over time as the bed collapses. The collapse is said to involve several successive stages:

(a) An initial stage where any bubbles present in the bed when the gas supply is abruptly interrupted escape from the bed.

(b) An intermediate stage where the gas present in the bed interstices is exhausted; particles have little or no interparticle contact during this stage as they squeeze closer and closer to each other.

(c) A final stage where particles do contact each other and eventually come to rest, supported by a network of particle-particle contacts.

Stage (b) is the one which is commonly analysed in detail. The results are employed to characterize the particulate material under investigation. Extrapolation back to time t_1, the instant when the gas flow was interrupted, is supposed to give information on the dense phase behaviour prior to interruption of the fluidization. Because the dense phase is expanded very little for Geldart group B and D particles, the collapse test technique is effectively limited to group A and AC powders.

Most workers who have performed and reported collapse tests have simply cut off the gas supply and allowed all excess gas in the system (in excess of that when atmospheric pressure is eventually reached) to escape by flowing upwards through the bed and bed surface. This simple method, shown schematically in Fig. 2(a), can be called a "single-vented system." The gas to be discharged includes not only that contained in the interstices of the particles, but also the excess gas which is found at t_1 in the plenum chamber below the distributor plate and in any piping downstream of the solenoid valve used to interrupt the flow. In these circumstances, it is clear that the amount of gas to be exhausted during stage (b) is a function not only of the dense phase expansion but also of: (i) the volume of the plenum chamber and of any piping downstream of the solenoid valve, and (ii) the pressure drop across the distributor plate, since a high Δp distributor will result in more moles of gas in the plenum chamber prior to gas interruption than a low Δp distributor. Clearly, the measurements cannot be simply a function of particle properties if equipment variables like plenum chamber volume and distributor pressure drop influence the results. Since fluidization workers have never been able to agree on any standardization of equipment, it is clear that collapse test results obtained in different columns have been subject to equipment-related factors which could have caused variations in results, even for two identical powders fluidized at the same superficial gas velocity with the same gas.

In order to counter this factor, a number of groups (e.g., Rowe et al, 1986; Park et al, 1991; Ham et al, 1991) have installed a second solenoid valve to vent the plenum chamber, as indicated by SV2 in Figure 2(b). This solenoid valve has been opened at the same instant as the gas flow has been interrupted via SV1. The idea seems to have been to bring the pressure in the plenum chamber to atmospheric (or to match the pressure in the freeboard region above the bed surface) as quickly as possible. In fact, in one case (Park et al, 1991) a low resistance connection was immediately established between the plenum chamber and freeboard region by means of a pipe. While massive venting of the plenum assures that no gas from the plenum chamber exits through the bed, it does not ensure that there is no reverse flow. After actuating both SV1 and SV2, the region above the distributor not only has a higher pressure than the freeboard, but it also finds itself at a higher pressure than gas in the plenum. As a result, gas flows downwards through the distributor. We may call this a "double over-vented system." Clearly the amount of gas above the distributor at $t = t_1$ which will reverse its direction and flow downwards through the gas distributor will depend on the distributor resistance, although the influence of the plenum chamber volume should now be minimal.

There is a third possibility, shown schematically in Fig. 2(c), which we introduced in a recent paper (Khoe et al, 1991). We may call this a "controlled double-vented system." There is again a second solenoid valve (SV2) to vent the plenum chamber. However, in this case there is a globe valve (GV) (or butterfly valve) on the plenum exhaust line to control the rate at which gas vents from the windbox during the collapse process. It is also essential to provide a means of measuring the pressure drop across the distributor plate. By trial and error, the globe valve is adjusted such that the pressure drop across the distributor plate is maintained as close as possible to zero during the venting/drainage/ collapse process. Alternatively, one could have this valve automatically controlled to maintain the pressure drop across the distributor equal to zero. Either method assures that there is negligible flow across the distributor, either upwards or downwards, as the bed surface collapses, hence assuring that the collapse is influenced by neither the plenum chamber volume nor by the nature of the gas distributor, except insofar as these influence the hydrodynamic conditions for $t < t_1$, i.e., prior to the instant when SV1 is abruptly closed and SV2 is simultaneously opened.

Pressure profiles for the three cases are shown schematically in Figure 2. Note that in case (a), the single-vented system, the pressure in the plenum chamber remains above that in the bed during the collapse and there continues to be upwards flow through the distributor at $t > t_1$. For case (b), the double over-vented system, there is a reverse flow through the distributor for $t > t_1$. For case (c), on the other hand, there is negligible pressure drop across the distributor and hence negligible flow through the distributor (upward or downward) during the collapse process.

In our laboratory we have been examining the collapse behaviour of beds of fine glass beads and FCC particles of different particle size distribution with the three different venting modes described above, two different plenum chamber volumes (214 and 1274 cm^3), two different superficial gas velocities (11 and 19 mm/s), and two different distributors, one with twice the resistance of the other. These experiments have been carried out with a column of diameter 100 mm with a settled bed depth of 450 mm.

Figure 3 shows some results for wide size distribution FCC particles with a relatively high pressure drop distributor and the larger plenum chamber volume (1274 cm^3). The superficial gas velocity prior to closure of the air supply by a solenoid valve was 19 mm/s. Particle properties are given in Table 2. It is seen that there is a significant difference between the results for the three cases discussed above. The single-vented system gives the slowest collapse, taking nearly 24s for the bed surface to fall to the minimum fluidization level. For the double over-vented system, collapse is significantly faster, the corresponding times to reach H_{mf} being only 15s. The controlled double-vented system gives intermediate results, with the corresponding time for the bed surface to reach H_{mf} being 20s. Extrapolating each of the linear portions back to time 0 gives significantly different intercepts, i.e., one would infer different particulate phase expansion depending on which of these venting methods one employed. Slopes of the linear portions also differ somewhat. Corresponding pressure drops across the distributor during the collapse process determined by a manometer are plotted in Figure 4. Note that controlled double-vented system does produce a pressure difference near 0 during the collapse, in contrast to the single-vented and double over-vented systems which gave over-pressures and under-pressures, respectively, in the windbox during bed collapse.

When the plenum chamber volume was decreased to 214 cm^3, (see Figure 5) the two lowermost curves of Figure 3 are essentially unchanged, whereas the curve for the single-vented systems moves downwards much closer to the controlled double-vented result. In the limit of zero plenum chamber volume, it is clear that the single-vented system should give the same result as the controlled double-vented system since there is no excess gas to be discharged from the windbox. Over-venting of the windbox can clearly be worse than not venting the windbox at all: typically the pressure drop across the distributor plate during fluidization is of the order of 30% of that across the bed; this means that gas can readily easily flow downwards through the grid after SV1 and SV2 are both opened in Figure 2(b). Note that a low Δp distributor is helpful for the single-vented system, whereas a high Δp distributor gives results closest to the controlled venting case for over-venting.

To summarize, if collapse tests are to give results which are representative only of the particles and their state of fluidization prior to gas interruption, one should vent the plenum chamber in a controlled manner to maintain zero distributor pressure drop (and hence negligible flow from or to the windbox) during the collapse process. Over-venting can be worse than single venting. If single venting is used, the volume of the windbox and the distributor pressure drop should both be as small as possible. Alternatively, analytical procedures are available (e.g. Tung and Kwauk, 1982; Geldart and Wong, 1985) to partially correct for excess gas in the windbox.

EXPERIMENTAL DETERMINATION OF THE ONSET OF TURBULENT FLUIDIZATION

Differences exist not only between the interpretation of what constitutes the turbulent regime of fluidization, but also in how to determine the onset of the regime. Methods employed to identify the transition in increasing order of sophistication are:

(a) Observation through the transparent wall of a column or by means of X-rays, looking for breakdown of coherent bubbling or slugging.

(b) Determination of bed surface level and hence of bed expansion. One then plots bed expansion vs. superficial gas velocity and looks for a change of slope.

(c) Measurement of the amplitude of pressure fluctuations in some manner. One then plots some measure of pressure fluctuation intensity vs. superficial gas velocity and determines the velocity (U_c) at which the fluctuations reach a peak, or the velocity (U_k) at which they level off again.

(d) Insertion of a capacitance probe which measures local voidage fluctuations. Transition to turbulent fluidization is said to occur when there is a transition from large amplitude, low frequency fluctuations to a lower amplitude, higher frequency signal.

As indicated in Table 3, method (c) is significantly more popular than the other three methods. However, the pressure fluctuation method is subject to a total lack of standardization in how the measurements are taken, whether absolute or differential pressure is recorded, whether or not the resulting signal is normalized, and what calculated parameter is plotted vs. superficial velocity. Details are provided by Brereton and Grace (1992) and summarized in Table 4. The key areas of difference are as follows:

(a) Measurement instrument: Most investigators have employed sensitive pressure transducers, but some have used manometers and others have not specified the instrument used.

(b) Absolute or differential pressure: Most investigators have used a differential pressure signal, but many have determined absolute pressures, and some have not bothered to specify whether their pressures were differential or absolute. When differential pressures have been taken, the height interval over which the signal has been taken has varied widely.

(c) Dimensional or dimensionless: About half of those who have used pressure fluctuation measurements to determine the onset of turbulent fluidization have not normalized the fluctuations in any way. The others have made the measurements dimensionless by dividing by some mean quantity, usually the time-mean absolute or differential pressure.

(d) Derived quantity: A number of different quantities have been derived from the fluctuating pressure signal by different investigators. Measures of the fluctuation intensity have included the r.m.s. fluctuation about the mean, mean absolute amplitude about the mean, peak-to-peak difference and the difference between the 10% and 90% levels on a cumulative probability curve.

Adding to the lack of consistency, some workers have taken U_c while others have utilized U_k to characterize the regime transition. Moreover, as discussed by Rhodes and Geldart (1986), when differential pressure measurements were taken, both taps may not always have been below the bed surface.

When one considers all of these inconsistencies, it is not surprising to find wide variations in the values reported by different laboratories. There is clearly no reason why parameters obtained in such different manners should produce consistent results or decipherable trends. Moreover, if differential pressures are measured, it is possible to have an artefact in the slug flow regime where, as shown by Kehoe and Davidson (1973), the pressure at a point rises and falls with a regular, approximately triangular waveform. The measured differential pressure fluctuation will then depend on whether the spacing between the two points of measurement is close to an integral multiple of the slug-to-slug spacing.

This is clearly another situation where fluidization workers should adopt some standardization in the interests of being able to interpret and compare each other's data. To this end, we have recommended (Brereton and Grace, 1992):

(a) that differential pressure measurements be taken;

(b) that the lowest tap be within two column diameters of the gas distributor;

(c) that the highest tap always be below the expanded bed level;

(d) that a sensitive pressure transducer capable of rapid response be employed;

(e) that the signals be normalized by dividing by the time mean differential pressure;

(f) that U_c, which is less subject to interpretation, than U_k be taken as corresponding to the onset of turbulent fluidization;

(g) that appropriate care be taken to avoid too large a dead volume or unequal volumes on the two sides of the differential pressure transducer;

(h) that all details of the measurements, including the heights at which the pressure taps are located, be fully reported.

CONCLUDING REMARKS

There seems to be an almost fatalistic attitude in the fluidization community that measurements are inherently difficult and subject to wide scatter, and that it is enough to make measurements and report them in the literature. Little attention is paid to reconciling methodology. There are few standard methods, no agreed nor standard systems. There is commonly minimal characterization of the particulate materials and incomplete description of the equipment used.

If Agricola really was practicing fluidization in the 16th century, would he find that we have progressed as far as we might? Even if we adopt the early 1940's or the modern beginning of fluidization, with 50 years behind us, is the fluidization community advancing the field as quickly as we might if we were willing to accept some greater measure of standardization in experimental techniques and in characterizing and reporting particle and experimental properties? The field has advanced and continues to advance, but we would go further if tests like the collapse test and measurement of regime transitions could proceed along agreed lines. This is one challenge, among many, for the next 50 years.

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NOTATION

B	gap thickness, m
d_p	mean particle diameter, m
d_{pi}	characteristic particle diameter for i^{th} size fraction, m
L	height of gap, m
p	gauge pressure, Pa
t	time, s
t_1	time at which fluidizing gas flow is interrupted, s
t_2, t_3	subsequent times, $t_3>t_2>t_1$, s
U_c	superficial gas velocity corresponding to peak pressure fluctuation intensity, m/s
U_k	superficial gas velocity where pressure fluctuation intensity levels off, m/s
U_{mf}	minimum fluidization velocity, m/s

U_{tf}	void throughflow velocity, m/s
x_i	mass fraction of particles in i^{th} size interval
z	height above distributor, m
ρ_p	particle density, kg/m^3

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