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Regimes of large-scale fluidized beds for solid fuel conversion

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A R T I C L E I N F O

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A B S T R A C T

The description of the fluidization behavior of Group B particles is ambiguous at high velocities in wide vessels like those in commercial solid-fuel converters, such as boiler furnaces and gasifiers. Bubbling, turbulent, and fast-fluidization regimes are analyzed here, especially related to fuel converters. It is claimed that the bottom bed in fluidized-bed boilers is operated in a bubbling regime of fluidization even at high gas velocities accompanied by a tall splash zone, as long as there is sufficient bed material in the system. It is also shown that the upper part of conventional circulating fluidized-bed boilers operates in a dilute-flow mode. In denser flows, a comparison between transport velocity and terminal velocity indicates that clustering flow is much stronger for Group A particles than for Group B, and hence, fast fluidization should be less developed for Group B than for Group A particles. The influence of temperature is also shown.

1. Introduction

Many of the investigations on fluidization regimes have been carried out in narrow vessels using Group A particles with the focus on catalytic reactors and pneumatic transport systems. The information from these studies has influenced also opinions on fluidization in wide vessels with Group B particles, such as in commercial fluidized-bed boilers and gasifiers for solid fuels, which often are said to operate under fast-fluidization conditions, and sometimes to contain the features of turbulent fluidization. In the present work these matters are analyzed and questioned.

2. Bed particles and equipment

2.1. Differences between solid-fuel conversion equipment (FCE) and test rigs for regime studies

- The bed particles. FCE uses Group B particles (although size distributions are present and segregation takes place). In regime studies Group A particles are preferred, even though studies involving Group B particles are carried out in relation to pneumatic transport when choking is of interest.
- The width of the fluidization vessel. In FCE the vessels are much wider than the maximum bubble size and slugging does not occur, while in narrow test rigs slugging often is passed on the way to turbulent fluidization, especially with Group B particles.
- The temperature. FCE is operated at 800–900 °C, whereas test rigs for regime studies are run under ambient conditions.
- The particle flow-rates are usually much lower in FCE than in catalytic reactors and pneumatic-conveying systems.

2.2. Differences between Group A and B particles

- Group A particle beds expand during an increase in velocity from minimum fluidization before bubbling starts. With Group B, bubbling starts directly at minimum fluidization velocity when the velocity is increased.
- Group A beds continue to expand more than Group B beds during an increase in fluidization velocity.
- Group A has a maximum bubble size, whereas Group B bubbles are limited only by the bed height or by the width of the fluidization vessel.
- During a rise in the fluidization velocity, Group A particles enter into the turbulent regime of fluidization directly if the vessel is larger than the maximum bubble size, or after slugging in narrower vessels, whereas Group B particles always enter in a slugging mode before the transition to turbulent fluidization in the narrow test rigs where observations have been made. However, the existence of a turbulent regime has been questioned [1]. In some fluidization diagrams, the turbulent fluidization has been located only to Group A particles [2,3], whereas in other diagrams [4] the turbulent regime extends also to Group B particles.
- A transport velocity, marking the onset of substantial circulation, can be determined both for Group A and B particles, but the definition of fast fluidization following the transport velocity is not clear. Bi et al. [5] identified fast fluidization to be limited by classical and

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accumulative choking, and this concept has been supported by others [6], who also refer to the similarity of their ideas with those of the well-known work of Takeuchi et al. [7]. In a recent summary of data on regimes [3], the region of fast fluidization identified by [5,6] is present only for Group A particles, whereas for Group B particles classical and accumulative choking coincide, implying that the fast fluidization region would be absent for such particles. Instead, Ref. [3] puts the upper limit of fast fluidization at the “minimum pressure velocity”, a limit which previously has been firmly discarded [8]. Besides, there is no such minimum in wide vessels owing to negligible wall friction.

2.3. General

Obviously, there are problems of interpretation on the applicability of the regime concepts, and direct observation in fuel-conversion equipment using Group B particles might give more clarity on their mode of operation. Below an example from a wide vessel fluidized with Group B particles will be provided.

3. Tests and conditions

3.1. The tests

The fluidization situation at high velocities and coarse particles will be discussed using results that have been published from the 12 MW circulating fluidized bed (CFB) boiler at Chalmers University [9]. The boiler has a height of 14 m and a cross section of about 1.5 × 1.5 m. The walls are not tapered in the bottom part, but the furnace is otherwise similar to those of CFB boilers and gasifiers. The boiler was operated with coal at a temperature of 850 °C, using only primary air during the tests concerned (all air was introduced through the bubble-cap nozzle-plate at the bottom). The furnace (the “riser” or “reactor”) is equipped with >50 pressure taps along the height, with the densest spacing in the bottom region to allow accurate measurements of bed height. The bed material was sand of a mean size of 0.32 mm, mixed with a small quantity of ashes. Two types of test were carried out to determine the velocity at maximum pressure fluctuation, u_c, which has been called transition to turbulent fluidization. The first set of tests was run with a constant mass of bed material in the system, M = constant. In the other set of tests, bed material was added to maintain a constant pressure drop, Δp_{ref} = constant, across the lowest 1.6 m of the riser. In all cases, the fluidization velocity was changed from a low value up to 6 m/s under otherwise almost constant conditions. (This could be achieved in this boiler by various means of temperature control, such as an external heat exchanger, air preheat, and flue-gas recirculation).

In a device operated with solid fuels, in addition to the impact from attrition, the ashes inevitably influence the composition of the bed. In separate tests under similar conditions [10], segregation of particle sizes was observed in the vertical direction: the average particle size at the exit was around 0.25 mm for an average bottom-bed particle size of 0.32 mm. In a previous test [11] it had been noted that the difference between the average sizes in the bottom and the top of the furnace decreased for rising fluidization velocities. Here, the range of the average particle size of 0.20 to 0.32 mm will be considered. [It should be noted that many CFB boilers operate with smaller particle sizes at the exit, resulting in circulating material sizes of 0.1–0.2 mm, compared to the present size in the top of the furnace of 0.25 mm. However, these particles all belong to Group B].

3.2. Temperatures

Almost all measurements on fluidization regimes have been carried out at room temperature. The relationships obtained to describe the regimes and their limits are usually of the type $Re = f(Ar)$. The Reynolds number is $Re = ud_p\rho_g/\mu$, u is velocity m/s, d_p particle size m, ρ_g gas density kg/m^3, and μ dynamic viscosity Ns/m^2. The Archimedes number is $Ar = d_p^3\rho_g g(\rho_p - \rho_g) / \mu^2$ where ρ_p is solids density kg/m^3 and g m/s^2 gravity. A temperature (T, K) dependence arises through the gas density and the dynamic viscosity. The gas density is

$$\rho_g = \rho_{g0} T_0 / T \tag{1}$$

From Sutherland's formula [12], expressed for simplicity for air and verified by comparison with data in the temperature range of interest, the temperature dependence of the dynamic viscosity can be expressed

$$\mu = \mu_0 \frac{T_0 + 120}{T + 120}\left(\frac{T}{T_0}\right)^{3/2} \tag{2}$$

Index 0 denotes ambient temperature, $T_0 = 290$ K, $\rho_{g0} = 1.23$ kg/m^3 and $\mu_0 = 18 \cdot 10^{-6}$ Ns/m^2.

Hence

$$\mathrm{Re} = \mathrm{Re}\frac{T + 120}{T_0 + 120}\left(\frac{T_0}{T}\right)^{5/2} \tag{3}$$

and

$$Ar = Ar_0\left[\frac{T + 120}{T_0 + 120}\right]^2\left(\frac{T_0}{T}\right)^4 \tag{4}$$

The terminal velocity (index t) can be expressed by a simplification of Haider and Levenspiel's expression for spherical particles [13]

$$\mathrm{Re}_t = 1/\left(18/Ar + 0.591/Ar^{0.5}\right) \tag{5}$$

which gives the temperature dependence of the terminal velocity of about $T^{-1/2}$ when $Ar \to 0$ and about $T^{1/2}$ when $Ar \to \infty$. The Ar of actual interest is in the order of 100 to 10,000.

4. Results

The standard deviation of the pressure fluctuations, registered while increasing the fluidization velocity, is shown in Fig. 1. The purpose of the tests was to identify u_c, the onset velocity of the transition to turbulent fluidization, as described in Ref. [14]. When the amount of bed material was constant in the system, a clear peak was determined, as shown in Fig. 1. However, this peak is an artifact, found in many systems, resulting from emptying of the bottom bed at high velocity, and redistribution of the bed material in the riser and in the system. On the other hand, if bed material is added to the bottom bed to maintain the pressure drop constant across the lowest 1.6 m of the furnace, there is no transition; no maximum is observed at least not in the range of velocities concerned. [The gas velocity was limited to the values seen in Figs. 1 and 2 by the rising pressure drop in the nozzles of the air distributor as a function of velocity, arising as a consequence of the unusual mode of operation].

The gradual reduction of the height of the dense, bubbling bed during a rise of the fluidization velocity is illustrated in Fig. 2 for the two cases of operation, constant pressure drop and constant bed content. In both cases there is a gradual transfer of bed material from the bottom bed to the riser and to the rest of the system as the fluidization velocity increases, despite recirculation of bed material, and the height of the dense bubbling bed decreases. In no case a change in the fluidization regime of the bottom zone has been observed as long as there is a bottom bed.

One could ask: what is the state of fluidization before the bed is consumed by elutriation? There are two observations indicating that the bed is bubbling as long as there is sufficient bed material. One observation is that the frequency character of the pressure fluctuations is maintained while the velocity increases, although the amplitude changes