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Fluid Dynamics and Transport Phenomena

Solid concentration and velocity distributions in an annulus turbulent fluidized bed☆

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ABSTRACT

Solid concentration and particle velocity distributions in the transition section of a ϕ 200 mm turbulent fluidized bed (TFB) and a ϕ 200 mm annulus turbulent fluidized bed (A-TFB) with a ϕ 50 mm central standpipe were measured using a PV6D optical probe. It is concluded that in turbulent regime, the axial distribution of solid concentration in A-TFB was similar to that in TFB, but the former had a shorter transition section. The axial solid concentration distribution, probability density, and power spectral distributions revealed that the standpipe hindered the turbulence of gas–solid two-phase flow at a low superficial gas velocity. Consequently, the bottom flow of A-TFB approached the bubbling fluidization pattern. By contrast, the standpipe facilitated the turbulence at a high superficial gas velocity, thus making the bottom flow of A-TFB approach the fast fluidization pattern. Both the particle velocity and solid concentration distribution presented a unimodal distribution in A-TFB and TFB. However, the standpipe at a high gas velocity and in the transition or dilute phase section significantly affected the radial distribution of flow parameters, presenting a bimodal distribution with particle concentration higher near the internal and external walls and in downward flow. Conversely, particle concentration in the middle annulus area was lower, and particles flowed upward. This result indicated that the standpipe destroyed the core-annular structure of TFB in the transition and dilute phase sections at a high gas velocity and also improved the particle distribution of TFB. In conclusion, the standpipe improved the fluidization quality and flow homogeneity at high gas velocity and in the transition or dilute phase section, but caused opposite phenomena at low gas velocity and in the dense-phase section.

1. Introduction

Compared with liquid–solid fluidization, gas–solid fluidization is characteristic of uneven axial and radial distributions of velocity and solid concentration because of the great density difference between continuous and dispersed phases. This difference results in nonideal mixing and poor two-phase contact. To overcome this shortage, various methods including internals have been used to generate better gas–solid distribution in bed to accommodate the targeted chemical reactions. Internals can be divided into two categories based on their purposes: (1) high-resistance ones to break up bubbles and agglomeration, including horizontal internals (*e.g.* mesh, baffle, and paddle) [1–3], pagoda-shaped body, ridge-shaped body [4,5], and packing; and (2) low-resistance ones to change flow direction and distribution in bed, including ring, swage, tube bundle, bluff body, spiral flow pate, and wedge-shaped internal [6–10]. The former internals, which are usually used in bubbling fluidized bed, can inhibit bubble growth, facilitate bubble break-up, enhance mass transfer between bubbles and emulsion phase, and increase bubble residence time [11]. The latter internals, which are principally used in fast fluidization, can break the core-annular structure, inhibit a thick ring near the wall, increase the central solid concentration, and improve the gas–solid residence time distribution (RTD) [1].

The turbulent fluidized bed (TFB) widely used in MTO (methanol to olefins) [12,13] has two characteristics: (1) coexistence of bubble phase and emulsion phase like in bubbling fluidized bed, and (2) core-annular structure as in a riser. Although the bubbles update in a high frequency and mass transfer is efficient in bed, the gas and solid backmixing and the resultant broad RTD are big problems of the turbulent bed. Thus, internals are generally used to inhibit axial backmixing and facilitate radial gas–solid movement rather than bubble break-up, eliminate gas bypass, or facilitate gas exchange between bubble and emulsion phases [14]. Among many internals, vertical tube has very small pressure drop. Early research is principally focused on heat exchange and abrasion in fluidized bed, whereas few studies are focused on the effects of vertical tube on gas–solid flow. Meanwhile, recent works are concentrated on two equipments. First is integral circulating fluidized bed (o-ICFB) in fast fluidization regime [15–17], in which the S-type axial concentration profile and a dense–dilute–dense three-ring radial distribution are

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observed; second is a fluidized bed involving a vertical membrane tube for coupling of reaction and separation [18]. These studies involved fast fluidized bed or bubbling fluidized bed. Few works have considered the effect of a standpipe in TFB.

Our previous research on TFB [19–23] revealed that the transition section is the principal component of TFB reactor and that its flow parameter distribution determines reactor performance. However, few studies are involved with solid concentration and particle velocity distribution in the transition section. Therefore, we conducted systematic experiments and analyses on the axial and radial distributions of solid concentration and particle velocity in the transition section for both TFB and annulus-TFB (A-TFB) with a standpipe. The present study aims to explore the effect of standpipe internals on flow parameter distributions and lay foundations for developing this type of reactor.

2. Experimental

The experimental setup is a stainless-steel circulating fluidized bed (CFB) (Fig. 1). The main column (10) is 200 mm in diameter and 6500 mm in height, with a 300 mm (i.d.) expanded section on the top for solid recycle. The concomitant column (5) with a 300 mm i.d. is used to adjust the solid circulating flow rate. The gas distributor (above 14) used in the turbulent bed is a stainless-steel perforated plate (aperture ratio = 5%; aperture = 2 mm), on which multilayer fine stainless-steel screens were paved for material leakage prevention. Large and small flow meters were used to ensure accurate flow rate regulation of fluidizing gas. The third flow meter was used to adjust flow rate of aeration air in the concomitant fluidized bed (5). The exit gas from the top of the turbulent bed went through two-stage cyclones, in which most particles were recovered and the gas finally passed through a bag filter. To investigate the effect of standpipe internals on gas–solid flow pattern in turbulent regime, A-TFB was developed by inserting a 50 mm (i.d.) central standpipe in the traditional TFB. The fluidized particles used in this experiment are colorless glass beads with a true density of 2400 $kg \cdot m^{-3}$ and a packing porosity of 0.4. The Sauter mean diameter of the particles was 106 μm, as measured by a Malvern laser particle size analyzer (Mastersizer 2000). Thus, they were determined as Geldart A–B particles [1].

Solid particle velocity and concentration in turbulent bed were measured with a PV6D optical probe purchased from the Institute of Process Engineering, Chinese Academy of Sciences. Two 1 × 1 mm fiber arrays are situated on the probe top, and each optical fiber in the array can send and receive optical signals. The calibrated interval between the two arrays is 2.25 mm. The stainless-steel sleeve of the probe is 382 mm long and 4 mm in external diameter. The probe collects gas–solid flow pattern information in the bed through multi-measurement ports. In the TFB experiment, 11 radial measurement points were selected in each measurement port according to the relative radial position $\phi = r / R_2 = 0$–1. In the A-TFB experiment, 8 radial measurement points were selected according to $\phi = (r - R_1) / (R_2 - R_1) = 0$–1 ($r$ is the radial distance to the bed axis, R_1 is the radius of the internal, and R_2 is the radius of the turbulent bed). The measurement parameters after experiment optimization were as follows: sampling frequency, 5 kHz; data length, 131072; and 5 measurements at each point.

3. Results and Discussion

3.1. Cross-sectional average solid concentration

The vertical distribution of cross-sectional average solid concentration, which was integrated from local solid concentration, in A-TFB under different operation conditions is shown in Fig. 2. The superficial gas velocities all located in the regime of turbulent fluidization. Meanwhile, the static bed heights were not too high to cause plug flow, also not too low to result in incomplete transition section. The bed can be divided into three parts according to the steepness of the cross-sectional average solid concentration: dense-phase section at bottom, sharp transition section in the middle, and dilute-phase section on top. The axial part above 2 m has the nearly unchanged solid concentration and belongs to the dilute-phase section. The solid concentration at the lower

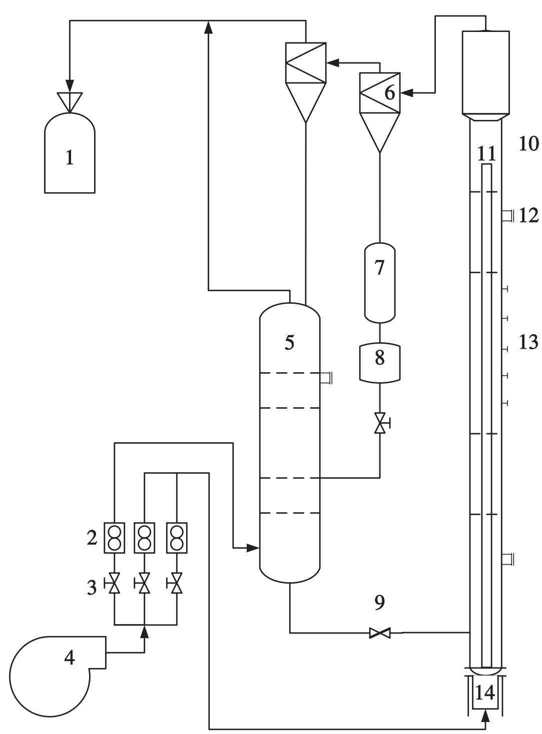

Fig. 1. Schematic of experimental setup. 1—bag filter; 2—gas flow meter; 3—stop valve; 4—blower; 5—fluidized bed; 6—cyclone; 7—expanded section; 8—store tank; 9—butterfly valve; 10—turbulent bed; 11—internal; 12—feed port; 13—measurement port; 14—buffer tank.

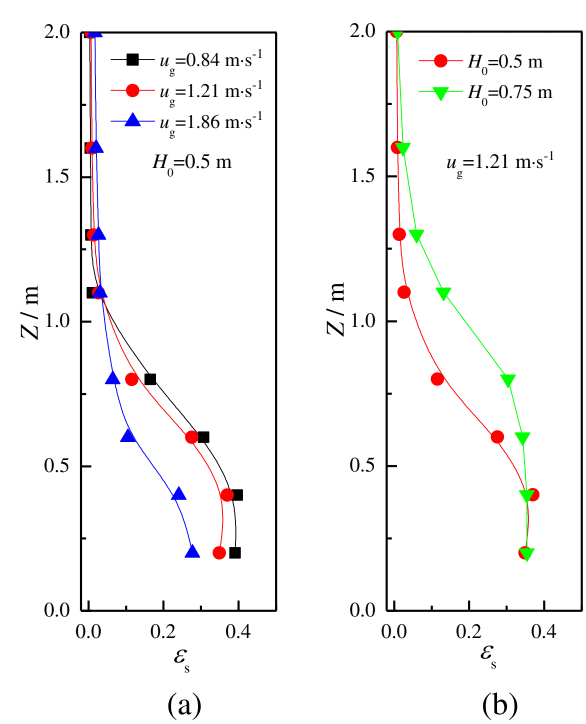

Fig. 2. Axial distribution of cross-sectional average solid concentration of A-TFB under different conditions ($H = 6.5$ m, $D_2 = 0.2$ m, $D_1 = 0.05$ m, (a) $u_g = 0.84, 1.21, 1.86$ $m \cdot s^{-1}$, $H_0 = 0.5$ m; (b) $u_g = 1.21$ $m \cdot s^{-1}$, $H_0 = 0.5, 0.75$ m).