Fluidization characteristics of wide-size-distribution particles in the fluidized bed with heating plate

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Abstract. To study the effect of heating plate on fluidization characteristics, the mean bed voidage, the axial solid concentration and pressure fluctuation under different superficial gas velocity were investigated. Experiments were conducted in a fluidized bed of 1500mm in height and 215 mm internal diameter with a conic bed. The results show that the internal heating fluidized bed has a little higher pressure drop, a little lower bed voidage, but a more stable fluidized state. The increase of the coalescence and breakage frequency of the bubbles can explain that the heating plate breaks the big bubbles well.

Keywords: Heated Plate Flow, Fluidization Characteristics, Surface Gas.

1. Introduction

During the production of polycrystalline silicon through the method of the fluidized chemical vapor deposition, the big bubbles can not only reduce the heat conductivity coefficient, resulting in uneven distribution of temperature field [1], but also aggravate the formation of fines, and then decrease product yield and affect the product performance [2], so designing a new type of fluidized reactor that can prevent big bubbles is necessary. Improving the fluidization behavior of the fluidized bed, mass transfer and heat transfer through installing internals has become a research hotspot now [3]. As the internals break big bubbles well, the fluidized bed with internals can have larger coalescence and breakage frequency of the bubbles, more uniform distribution of concentration field, higher gas-solid contacting efficiency. Hedrick [4] designed the fluidized bed with baffles on which were disproportionately distributed openings to improve the distribution uniformity of the gas and the gas solid contact efficiency.

In this work, a new type of fluidized bed reactor for the production of polycrystalline silicon was put forward. There were 18 heating plates made by silicon in the fluidized bed, the heating plates could not only heat gases and solids in the bed, but also have an important role as the internal components to affect the fluidization of particles in the bed. Meanwhile, heater made by silicon can avoid that silicon deposit on the wall in the convention external reactor. The main objective of this work is experimentally illustrating the more stable fluidized state and the better effect of breaking big bubbles in the fluidized bed with internal heating plate.

2. Experimental equipment and methods

2.1 Experimental equipment and procedure

In order to investigate the fluidization characteristics in the fluidized bed internal heating plate with wide particle size distribution, an experimental device for this work was made, as shown in Figure 1. The experimental setup was carried out in an organic glass column of 0.215 m in diameter and 1.5 m in height. Air as the gas phase was introduced from the bottom of the conic bed, where a perforated-plate gas distributor of 1.3% of porosity provided a homogeneous gas distribution, and its flow rate was controlled by a rotor flow meter. Several air flow rates were used in the range of 10-45 m3/h STP, corresponding to superficial gas velocities ug in the range of 0 .0773-0.348 m/s. The conic bed could give a better fluidization to the large particles at the bottom. Fluidization was performed

with air at ambient temperature. After that, air came into the cyclone from the top, and the hopper collected the fines which were carried out.

A ruler has been placed along the glass column to determine the location of the bed surface. A pressure sensor located at 220 mm above the gas distributor, on the central axis of the bed connected to an online monitoring system allowed measuring the pressure fluctuations in the bed. The sampling frequency for dynamic pressure measurement was 100 Hz, with a sampling period of 60 s. The height of powders introduced in the column was fixed at 250 mm.

In this work, the heating plates were simulated by three groups of cuboid organic glass plates, located at 200 mm, 400 mm, 600 mm, and adjacent groups were installed perpendicular to one another. Each one was composed of six plates, 100 mm wide and 3 mm thick, and the length from the center to the outside were 180 mm, 155 mm, 100 mm, showing the layout symmetry axis on both sides.

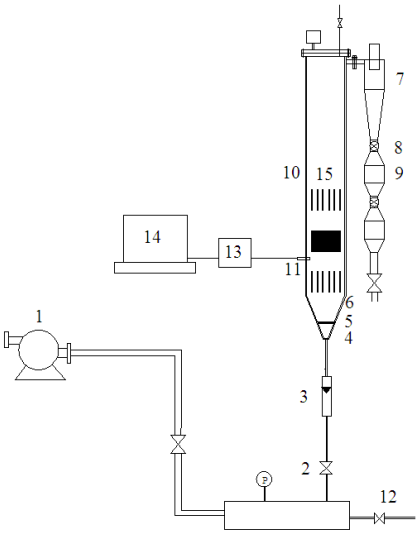

Figure 1. Experimental flow chart

1-fan; 2,8,12-ball valve; 3-flow meter; 4-air chamber; 5-distributor; 6-conic bed; 7-cyclone; 9-hopper; 10-fluidized bed; 11-pressure sensor; 12-atmospheric valve; 13-A/D; 14-computer; 15-heating plate

2.2 Materials

In the process of polycrystalline silicon production, CVD can form a wide particle size distribution. So two groups of polydisperse particles belonging to Geldart B [5] were made composed of four components by particle size as follows: a) 45-80 μm, b) 106-113 μm, c) 250-380 μm and d) 380-500 μm. The weight ratios of the above four components in Set 1-2 were 0:30:30:40, 20:20:20:40, respectively. The particle sizes and properties are summarized in Table 1.

Table 1. The properties and size distribution of glass beads

Glass bead	Particle proportion by weight (%)				Average Size (μm)	True density (kg/m^3) (kg/m^3)	Porosity
	45-80μm	106-113μm	250-380μm	380-500μm			
Set 1	0	30	30	40	203.9	2600	0.366
Set 2	20	20	20	40	125.9	2600	0.331

2.3 Experimental method

The mean bed voidage of fluidized bed is an important parameter of the axial solid concentration as a whole, and it can indirectly explain the particle distribution in the dense zone. The mean bed voidage of the fluidized bed can be calculated by taking Formula 1 [5]:

$$1-\bar{\varepsilon}=\frac{V_s\rho_b}{V\rho_s} \quad (1)$$

In bubbling bed and turbulent bed, the axial pressure drop of the fluidized bed can be calculated by taking Formula 2 [6]:

$$\Delta P=\left[\rho_g(1-s)+\rho_p s\right] gh \tag{2}$$

As $\rho_P \gg \rho_g$, $\rho_g(1-s)$ in the Formula 1 can be negligible. After introducing $\rho_B = \rho_P s$ into it, we can get the axial solid concentration as Formula 3:

$$\rho_B=\frac{\Delta P}{g\,h} \tag{3}$$

According to Formula 3, as long as the pressure drop between the two points along the bed is measured, the solid concentration in the center can be got. As radial position of the pressure tap can also influence the pressure drop, the method can not be effectively used in the conic bed. But we can calculate the approximate solid concentration between 0 and 200mm through the voidage of the dense phase because the bed between the two points is the only section completely in the dense phase zone in this work.

When calculated the power spectral density, we redefined pressure signal to show the power spectrum of pressure fluctuations more clearly by:

$$x(t)=p(t)-\overline{p(t)}[7] \tag{4}$$

After that, the signal was calculated by Fourier transform. Then we estimated the true power spectrum by:

$$G(f)=\frac{1}{N}|F(f)|^2[8] \tag{5}$$

3. Results and discussion

3.1 Bed pressure drop

The bed pressure drop can indicate the resistance of bed, which would provide the important basis for industrialization and engineering design. The pressure drop reached a constant value once the particles were fluidized, no matter with heating plates or not in the fluidized bed. But with heating plate mean a larger pressure drop, as is shown in Figure 2, which was because that the short circuit caused by the bubble flow was destroyed by internal heating plate, and the drag force would increase due to friction between the plate and particles.

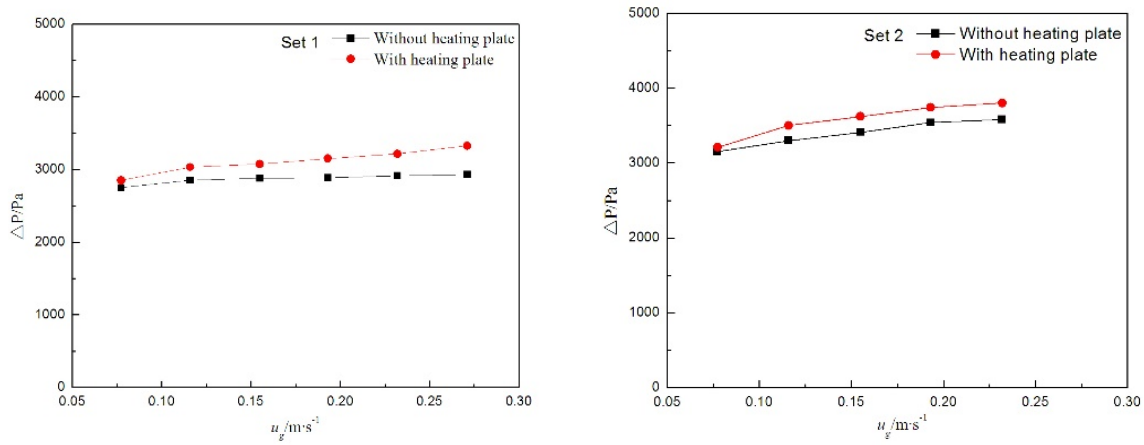

Figure 2. Comparison of the pressure drop under different superficial gas velocities

3.2 Average bed voidage

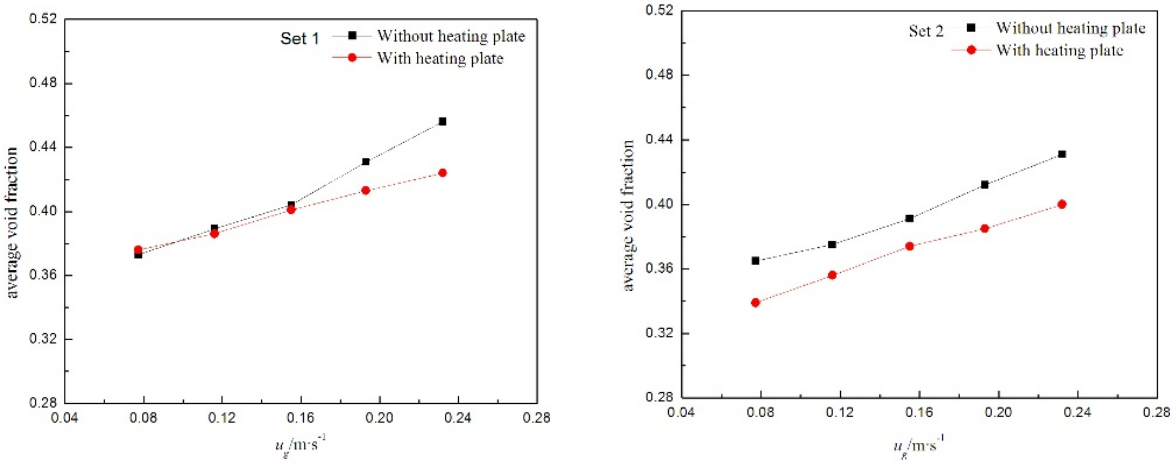

Figure 3. Comparison of the average bed voidage under different superficial gas velocities

The internal heating plate would affect the characteristics of the gas-solid two phases flow, thus led to the change of the average bed voidage. In the presence of the internal heating plate, the average bed voidage was smaller during the course of fluidization, so the appropriate size and location were necessary. But the increasing trend with superficial gas velocity was more smooth, which mean a stable fluidized state.

3.3 Axial solid concentration

The solid concentration of different bed section in the axial direction were measured and calculated for Set 2. The major difference mainly appeared in 200-400 mm and 400-600 mm, that the fluidized bed with the internal heating plate had a higher solid concentration in 200-400 mm and a lower solid concentration in 400-600 mm. This was because the internal heating plate suppressed the big bubbles formation well, thus the number of particles splashed to the higher levels due to breaking of big bubbles at the interface decreased. And then the fluidized state became smooth.

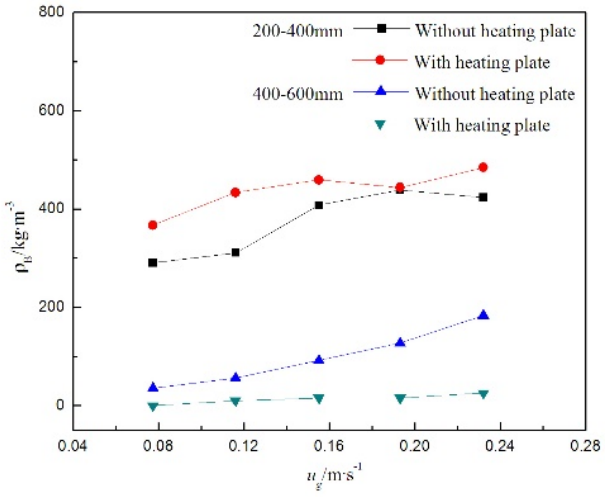

Figure 4. Comparison of the axial solid concentration under different superficial gas velocities for Set 2

3.4 Pressure fluctuation

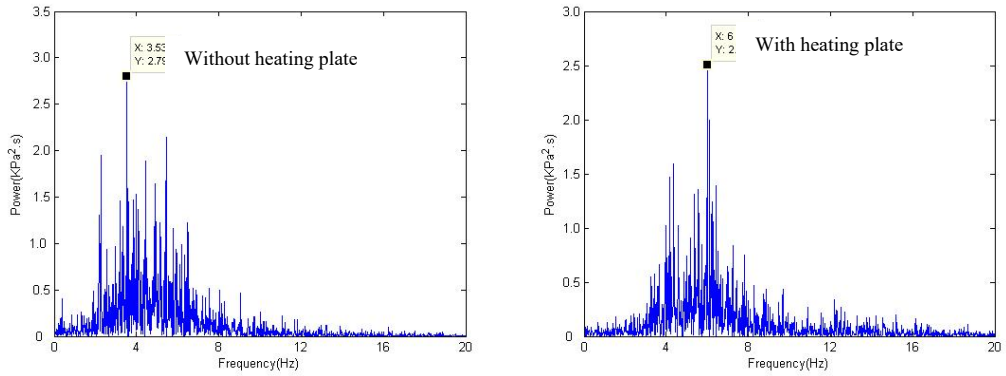

Figure 5. Power spectra from FFT at 0.15m/s for Set 1

To study the effect of the internal heating plate on the the coalescence and breakage of bubbles further, we calculated the pressure pulsation power spectrum for Set 1. Figure 5 showed that at 0.15m/s, the fluidized bed with internal heating plate had a larger main frequency value, increasing from 3.533 to 6.000Hz, and a smaller peak value, decreasing from 2.795 to 2.504 $Kpa^2 \cdot s$. The larger main frequency value and smaller peak value showed the increase of the coalescence and breakage frequency of the bubbles and the decrease of the bubble size, which could explain that the heating plate breaks the big bubbles well.

4. Conclusions

(1) The fluidized bed with internal heating plate had a little higher pressure drop, a little lower bed voidage, but a more stable fluidized state.

(2) The coalescence and breakage frequency of the bubbles in the fluidized bed with internal heating plate increased and the bubble size decreased, and the internal heating plate broke big bubbles well.

Nomenclature

$\bar{\varepsilon}$= mean bed voidage
V_s=static volume,m^3
V=fluidized volume,m^3
ρ_b=bulk density, $kg \cdot s^{-1}$
ρ_s=true density, $kg \cdot s^{-1}$
u_{mf}=minimum fluidization velocity, $m \cdot s^{-1}$
u_g=superficial gas velocity, $m \cdot s^{-1}$
u_t=carry-over rate, $m \cdot s^{-1}$
H=axial height, m
p(t)=acquired data, Kpa
f=sampling frequency, Hz

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