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A MODELING STUDY OF GAS STREAMING IN A DEEP FLUIDIZED BED OF GELDART GROUP A PARTICLES

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ABSTRACT

Gas streaming has been modeled in a deep fluidized bed of 5 m depth and 0.3 m inside diameter. The model results suggest that the lower pressure drop of the stream zone compared to the remainder of the bed is the reason for severe streaming flow in deep beds. The effects of different parameters such as bed depth, gas velocity and particle size on the severity of the streaming flow are also evaluated with the model.

INTRODUCTION

Several studies in the past decade have demonstrated that in sufficiently deep fluidized beds (i.e. beds approaching a depth of 1 m or greater) of Geldart Group A particles (1), gas bypassing may occur by increasing the superficial gas velocity (2-6). When this phenomenon occurs, the fluidizing gas bypasses the bed in the form of streams of gas, leaving a large fraction of the bed unfluidized or poorly fluidized. The concept of gas streaming was first reported in the literature by Wells (2). He performed several experiments in large scale units with up to 2.5 m diameter and 5 m bed depth and observed streaming flow under conditions that were expected to lead to operation in the bubbling regime. He attributed the streaming phenomenon to gas compression, caused by the pressure head of the deep solids bed over the distributor.

Karri et al. (3) investigated the formation of streaming flow in a column of 0.3 m inner diameter and 4.9 m height. They found that for all combinations of operating conditions investigated, the addition of a sufficient amount of fines to the bed of Geldart Group A particles was able to delay the streaming. In another work, Issangya et al. (4) used several pressure transducers mounted at various radial positions to detect the presence of streaming flow.

Recently, Karimipour and Pugsley (5) have performed a systematic study on the streaming flow in deep beds of FCC particles. They discussed the effects of streaming flow on the pressure fluctuations time series measured in the fluidized bed for different combinations of bed depth, gas velocity, particle size, and distributor design. They concluded that streaming flow does not appear suddenly, but emerges gradually in the bed by increasing the bed depth. They found that although changing parameters such as superficial gas velocity and/or fines content can reduce the severity of the streaming flow, streaming is the dominant phase for deep fluidized beds operating at gas velocities where a fully bubbling bed regime would normally be anticipated.

The only mathematical model to predict the onset of gas streaming is that of Wells (2). Wells (2) concluded that when the ratio of the density at minimum fluidization to the density of the emulsion phase becomes less than a critical value for a given bed depth, streaming occurs. The model of Wells (2) was tremendously valuable for improving the understanding of streaming, but it was not a direct function of operating conditions such as bed depth and gas velocity. The objective of the present work is to develop a simple phenomenological model for the streaming flow and to use the model to evaluate the effect of bed depth, gas velocity, and particle size.

MODEL DEVELOPMENT

Based on our finding from a separate experimental campaign (e.g. 5), the deep fluidized bed is divided into two adjacent regions in which the smaller region is occupied by the stream flow and the other region is assumed to be at minimum fluidization conditions. It is assumed that by increasing the superficial velocity the gas in excess of that required for minimum fluidization is directed into the stream zone. Also based on our experimental observations, the cross section of the gas stream is assumed to be circular and its diameter to be less than one fourth of the bed diameter. The stream therefore forms a vertical cylinder of constant diameter along the fluidized bed. A small lateral zone above the distributor is reported to be better fluidized (2) and gas and particles from other parts of the distributor find their way towards the stream and move upward through the stream. As such, particles can be assumed to move upward only in the stream and after discharging at the surface of the bed slowly return to the bottom through the non-streaming region.

Similar to the acceleration zone of a circulating fluidized bed riser (6, 7), the stream can be modeled by a force balance over a single particle inside the stream:

$$\rho_p V_p \frac{d\upsilon_p}{dt} = \frac{1}{2}\rho_g \left(\frac{u_{st}}{\varepsilon_g} - \upsilon_p\right)^2 A_p C_D - (\rho_p - \rho_g) V_p g \qquad (1)$$

Assuming the particles as spheres of constant diameter, and incorporating Eq. 2 from the derivative theory, the force balance equation can be re-written as Eq. 3:

$$\frac{d\upsilon_p}{dt} = \upsilon_p \frac{d\upsilon_p}{dz} \qquad (2)$$

$$\frac{d\upsilon_p}{dz} = \frac{3\rho_g C_D}{4 d_p \rho_p \upsilon_p}\left(\frac{u_{st}}{\varepsilon_g} - \upsilon_p\right)^2 - \frac{g}{\rho_p \upsilon_p}(\rho_p - \rho_g) \qquad (3)$$

We have estimated the drag coefficient, C_D, in Eq. 3 based on the correlation of Mostoufi and Chaouki (8). The porosity in these equations is calculated from the solids mass balance equation as follows:

$$G_p = \rho_p (1 - \varepsilon_g) \upsilon_p \qquad (4)$$

The initial value of the particle velocity at the bottom of the stream is obtained from the solids mass balance. Thus, Eq. 3 will be solved subject to the following initial condition:

$$\upsilon_p \Big|_{z=0} = \frac{G_p}{\rho_p (1 - \varepsilon_{mf})} \qquad (5)$$

Once the axial profile of particle velocity in the stream is determined from Eq. 3, the corresponding solids holdup can be calculated from

$$\varepsilon_p = 1 - \varepsilon_g \tag{6}$$

The axial profile of the pressure drop along the stream can be determined from the momentum balance over the stream. The momentum balance could be expressed as follows:

$$\frac{dp}{dz} = \left(\frac{dp}{dz}\right)_{head} + \left(\frac{dp}{dz}\right)_{acceleration} + \left(\frac{dp}{dz}\right)_{friction} \tag{7}$$

where

$$\left(\frac{dp}{dz}\right)_{head} = \rho_p \varepsilon_p g + \rho_g \varepsilon_g g \tag{8}$$

$$\left(\frac{dp}{dz}\right)_{acceleration} = \rho_p \varepsilon_p \upsilon_p \frac{d\upsilon_p}{dz} + \rho_g \varepsilon_g \frac{u_{st}}{\varepsilon_g} \frac{d}{dz}\left(\frac{u_{st}}{\varepsilon_g}\right) \tag{9}$$

The pressure drop caused by friction includes two sources, i.e., gas-wall and particle-wall frictions:

$$\left(\frac{dp}{dz}\right)_{friction} = \left(\frac{dp}{dz}\right)_{gas-wall} + \left(\frac{dp}{dz}\right)_{particl-wall} \tag{10}$$

These pressure losses are defined by the Fanning equation as

$$\left(\frac{dp}{dz}\right)_{gas-wall} = f_g \frac{1}{2d_{st}} \rho_g \frac{u_{st}^2}{\varepsilon_g} \tag{11}$$

$$\left(\frac{dp}{dz}\right)_{particle-wall} = f_p \frac{1}{2d_{st}} \rho_p \varepsilon_p \upsilon_p^2 \tag{12}$$

Since gas-wall and particle-wall frictions form a minor portion of the overall pressure drop, type of the friction factor does not have a major effect on the results. Here, the gas-wall friction factor, f_g, has been calculated from the Blasius formula (9)

$$f_p = \frac{0.316}{Re_g^{0.25}}, \qquad Re_g \le 10^5 \tag{13}$$

and the particle-wall friction factor has been estimated using the correlation of Kanno and Saito (10)

$$f_p = \frac{0.057}{2\upsilon_p}\left(g d_{st}\right)^{1/2} \tag{14}$$

The wall in our case corresponds to the “wall” of the cylindrical stream in the bed. In order to solve these equations, the solid circulation rate (G_p) is needed as an input. Since the system is not a real circulating fluidized bed, a pseudo-circulating rate may be calculated from the correlations proposed for the internally circulating fluidized bed. An internally circulating fluidized bed resembles the current case in that both of the systems involve flow of gas and solids between a fluidized bed at minimum fluidization conditions and a dilute bed (a riser in an internally circulating fluidized bed and a stream in the current case). The net rate of the particle exchange between two zones along the fluidized bed is considered to be trivial. The correlation of Jeon et al. (11) has been used for this purpose:

$$\Delta P_{or} = 5.327 \times 10^3 \left(\frac{u_{st}}{u_{mf}}\right)^{0.795} \left(\frac{d_p}{d_{or}}\right)^{0.728} \tag{15}$$

$$G_p = C_{dis} \frac{S_{or}}{S_{st}} \sqrt{2\rho_p (1-\varepsilon_{mf}) \Delta P_{or}} \quad (16)$$

In the above equations, the orifice refers to that point at the bottom of the bed that allows for the exchange of gas and particles between the stream and non-streaming zones.

For the pressure drop through the none-streaming zone which is considered to be at minimum fluidization conditions, the pressure drop is assumed to be due to the mass of the particle bed:

$$\frac{dp}{dz} = \rho_p g (1-\varepsilon_{mf}) \quad (17)$$

RESULTS AND DISCUSSIONS

The model predictions of pressure drop along the fluidized bed for a bed depth of 5 m are provided in Fig. 1. As can be seen in the figure, the model predicts a lower pressure drop immediately above the distributor for the non-stream zone compared to the case of the stream zone. Therefore, streams do not form in this region. However, the stream pressure drop decreases dramatically with increasing distance from the distributor, which makes the streams a preferable pathway for the gas. The higher pressure drop of the stream immediately above the distributor is due to the much higher flow of gas and particles in the stream compared to the non-stream zone. Similar trend of pressure drop has been reported for the bottom of FCC risers (7, 8). As illustrated in the figure, as the upper surface of the bed is approached, the difference between the pressure drop of the streaming and non-streaming zones decreases. The result of this would be that preferential flow of gas through the stream would be diminished, allowing gas to diffuse into other parts of the bed and provide more uniform fluidization at upper regions. This is consistent with visual observations from experiments, which showed improved fluidization at the upper regions of the bed.

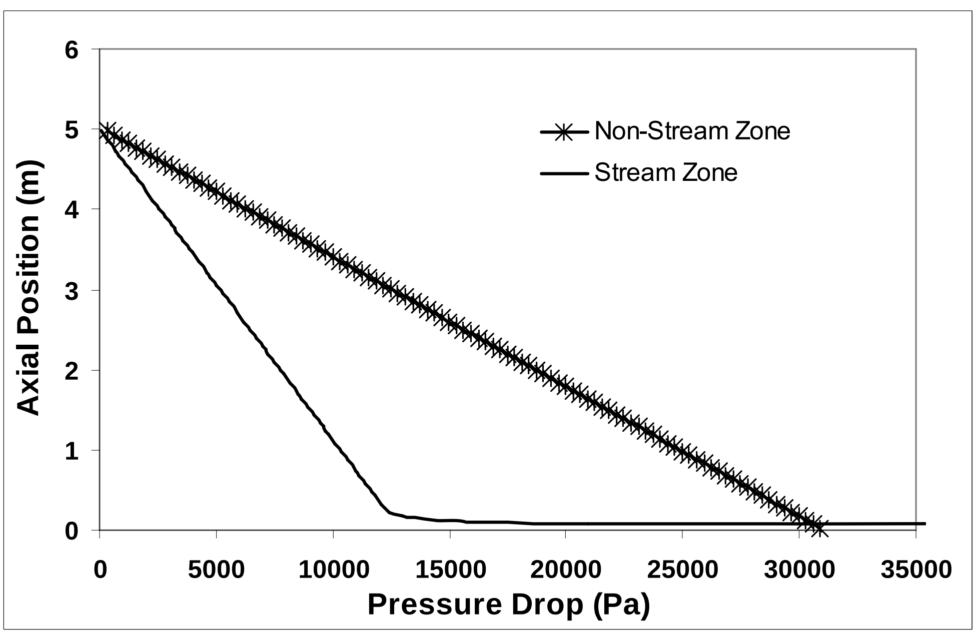

Figure 1. Axial profile of the pressure drop in the fluidized bed, Bed depth = 5 m, Superficial gas velocity = 0.2 m/s, Particle diameter = 84 microns

Effect of Bed Depth

Fig. 2 illustrates the differences between the pressure drops of stream and non-stream pathways at the bottom of the fluidized bed for different bed depths. As can be seen, the difference in the pressure drops of the two zones, which is considered to be the motivation for the formation and stability of the streams, increases with increasing bed depth. Experimentally we found that the onset of streaming flow occurred gradually in the fluidized bed as bed depth was increased. According to the model results, this can be attributed to the gradual increase of the difference in pressure drop between the streaming and non-streaming zones. This difference is probably low enough in shallow beds that the gas is able to fluidize all of the cross section and prevents the formation or permanence of streaming flow.

Effect of Gas Velocity

Fig. 3 provides the axial profile of the pressure drop in the fluidized bed for different superficial gas velocities. As Fig. 3 illustrates, two changes occur in the fluidized bed by increasing the gas velocity. Firstly, the difference between the pressure drops of the streaming and non-streaming zones decreases and secondly, the region expands above the distributor where streaming is not preferred or present. The positive influence of increasing the gas velocity on diminishing the streaming flow has been emphasized in all of the previous experimental works in the literature (2-6). As the figure indicates, at gas velocities higher than 1 m/s streaming flow is not preferred anywhere in the fluidized bed and uniform fluidization would be possible throughout the bed.

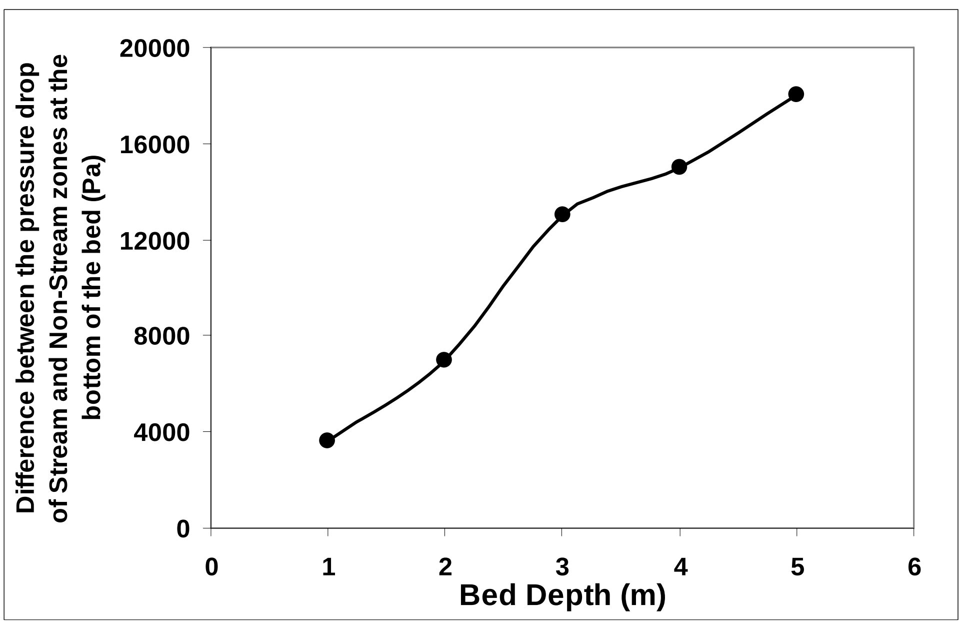

Figure 2. Difference between the pressure drop of Stream and Non-Stream zones at the bottom of the bed for different bed depths, Superficial gas velocity = 0.2 m/s, Particle diameter = 84 microns

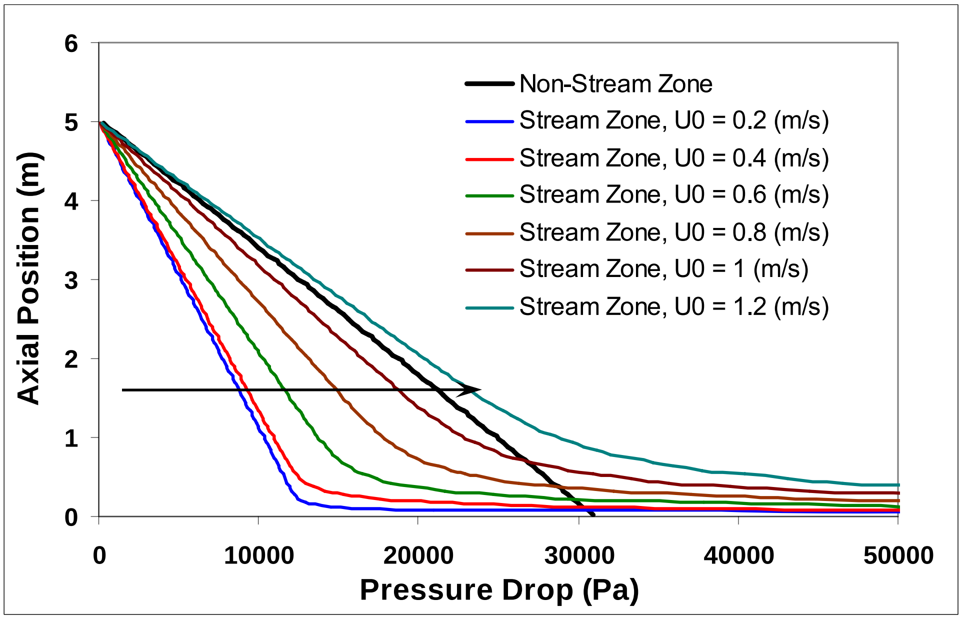

Figure 3. Axial profile of the pressure drop in the fluidized bed for different superficial gas velocities, Bed depth = 5 m, Particle diameter = 84 microns

Effect of Particle Size

Fig. 4 illustrates the axial profile of the pressure drop in the fluidized bed for different particle sizes and a constant particle density of 1400 kg/m^3. As can be seen, the pressure drop in the stream increases by increasing the particle size. Thus, its preference as an alternative pathway with lower pressure drop for gas decreases gradually. According to the literature, streaming flow has only been reported for Geldart Group A particles; it does not appear to exist for coarser Geldart B particles (2-6). Thus, as the model predicts, the fluidized bed of these particles display uniform fluidization. The results show that the model is able to predict this directional effect of increasing particle size.

Effect of Stream Size

The effect of the size of stream zone (i.e. stream diameter) on the axial profile of pressure drop has also been investigated (results not shown due to space constraints). Our model predicts that decreasing the stream size from 1/4 to 1/8 of the bed diameter reduces the preference of streaming as an alternative pathway for gas flow.

CONCLUSIONS

In the present work, gas streaming flow has been modeled in a deep fluidized bed of 5 m bed depth and 0.3 m diameter. The trend of the model predictions have been qualitatively compared and validated with the experimental findings. The model is based on the assumption that the stream already exists in the bed. The initiation of streaming flow has been discussed in our previous work (6). According to that work, the potential for streaming always exists in a fluidized bed. The results of the present work suggest that what causes a severe streaming flow with increasing bed depth is probably the gradual increase of the difference between pressure drop of two zones: that smaller portion of the bed where streaming becomes preferred and the remainder of the bed at minimum fluidization. Our model results show that increasing the bed depth favors the streaming flow, while increasing the gas velocity increases the uniformity of the bed and decreases the streaming severity. Streaming flow was found to be less severe for larger particle sizes. All of these findings are in conformity with experimental investigations reported previously in the literature.

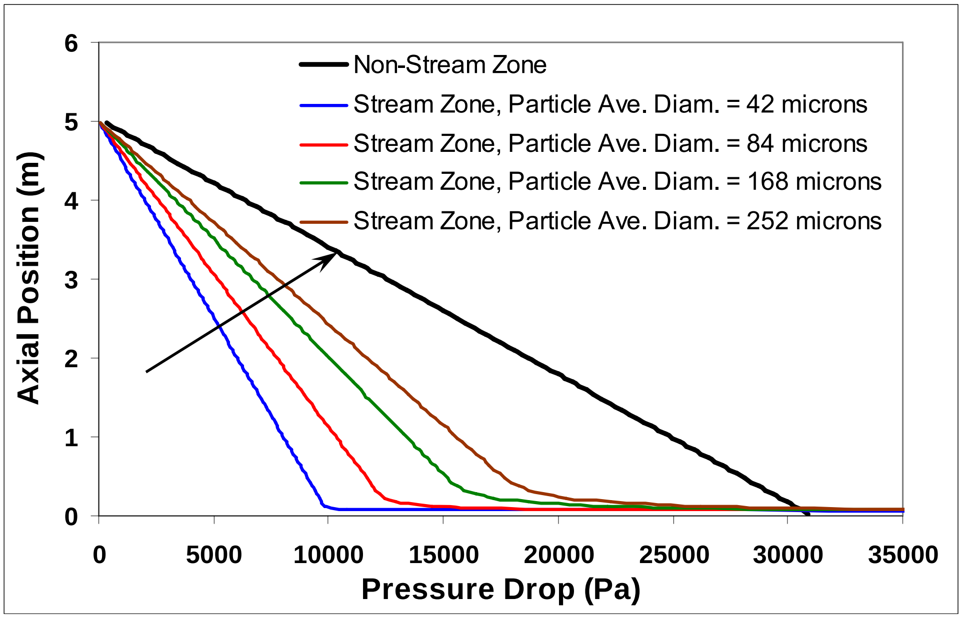

Figure 4. Axial profile of the pressure drop in the fluidized bed for different particle sizes, Bed depth = 5 m, Superficial gas velocity = 0.2 m/s

NOTATION

A_p	cross-sectional area of particle (m^2)
Ar	Archimedes number ($d_p^3\rho_g(\rho_p-\rho_g)g/\mu^2$)
C_{dis}	gas discharge coefficient
C_D	effective drag coefficient
d_p	particle diameter (m)
d_{st}	stream diameter (m)
D	fluidized bed diameter (m)
f	drag coefficient correction factor
f_p	solid-wall friction factor
f_g	gas-wall friction factor
g	acceleration of gravity (m/s^2)
G_p	solids flux (kg/m^2s)
p	pressure (Pa)
ΔP_{or}	orifice pressure drop (Pa)
Re_g	gas Reynolds number ($D\ U_0\ \rho_g/\mu_g$)
S_{or}	orifices cross sectional area (m^2)
S_{st}	stream cross sectional area (m^2)
t	time (s)
u_{mf}	minimum fluidization velocity (m/s)
u_{st}	gas velocity in stream (m/s)
v_p	particle velocity (m/s)

V_p particle volume (m^3)
z fluidized bed height above distributor (m)

Greek Letters
ε_g gas voidage
ε_p gas voidage
ε_{mf} voidage at minimum fluidization
ρ_g gas density (kg/m^3)
ρ_p particle density (kg/m^3)
μ gas viscosity (Pa·s)

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