Features of particle flows hydrodynamics in recirculating systems and pneumatic valves of CFB boilers

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Abstract. Circulating fluidized bed (CFB) technology is characterized by relatively high reactor velocity that exceed the transport velocity of medium-sized particles, and the presence of separators (mainly of the cyclone type) with a system for returning trapped particles to the reactor. These systems consist of downpipes equipped in the lower part with pneumatic valves for transporting material from the low (atmospheric) pressure zone in the cyclone to the high-pressure zone in the furnace. Pneumatic valves operate on the principle of fluidized valves, and the locking function of the valves is performed by a lifting section with an inclined return flow to the furnace (loop gate or J-valve) or a horizontal section connecting the drain to the furnace (L-valve). The report discusses the current state of development and methods for standpipe and pneumatic valve operation and calculating. A brief description of the test rigs and the results of our own studies of CFB returned system are given. The results of the work can be used for the design and commissioning of a circulating fluidized bed reactors, including systems with dual reactors for polygenerating systems and chemical looping.

1 Introduction

Circulating fluidized bed boilers (CFB) are widely used in the energy sector. The main advantages of CFB technology are well known, which are determined by the long residence time of particles in the reaction zone due to the high multiplicity of circulation, a stable and relatively low temperature in the furnace with a stage air supply, and the optimal temperature for sulfur oxides capture with limestone consumption. The key element of these boilers is the system for removing the material caught in the ash separators (cyclones) back to the furnace. These systems consist of downpipes equipped in the lower part with pneumatic valves for transporting material from the low (atmospheric) pressure zone in the cyclone to the high-pressure zone in the furnace. Pneumatic valves operate on the principle of fluidized valves, and the locking function of the valves is performed by a lifting section with an inclined return flow to the furnace (loop gate or J-valve) or a horizontal section connecting the drain to the furnace (L-valve). Typical scheme of CFB with J-valve is shown on Figure 1. Lowering movement
of the riser can occur in the mode moving down a dense bed or in the transitional regime, when a large leakage flow of the fluidizing agent (gas, steam, air) in the riser and possible mode bubbling fluidization, which can lead to leakage of air into the cyclone. The limits of the modes depend on the slip velocity (the sum of the material and gas velocities with different signs of the direction of movement). If the slip velocity positive (positive direction-down), then the movement occurs in a dense bed, if negative—in the transition mode, and at a high value of this velocity—in the bubbling bed mode. However, it is quite difficult to determine the slip velocity if you do not know the proportion of air entering the riser from the total air flow supplied to the lower part of the riser (air split). It is known that a significant part of the air goes with the flow of material to the lifting part of the gate or to the horizontal part of the L-valve [1 - 5].

![Figure 1. Stand alone CFB](image)

We performed an analysis of the influence of particle properties and regime parameters and proposed dependencies for the calculated estimation of the slip velocity based on the measured pressure gradient [5]. This paper provides a clarification of these dependencies. To estimate the proportion of gas entering the riser in the mode of movement in a dense bed, experimental studies were conducted using the method of gas tracers. The results of experiments are presented and compared with known data [2, 3, 4].

The test rig of [2] consisted of a riser, a cyclone, a standpipe and a loop seal. The riser had a cross-section area of 0.1x0.1 m² and a height of 4.5m. The cyclone was of high separation efficiency. The standpipe had a height of 3.0m and a diameter of 0.08m and connected the riser with a loop seal (J-valve). Particles real density was 2625 kg/m³, bulk voidage 0.50, minimum voidage 0.58, minimum fluidization velocity 0.09 m/s, Sauter diameter 0.36 mm. To study the gas flow behavior in the standpipe, high purity CO₂ gas was used as the tracer.

In [2] was shown that with increasing aeration rate, the pressure drop of the horizontal section increases due to higher solid circulation flux (Gs). The loop seal characteristics are closely related to the flow state in the standpipe, which affects the actual gas passing through the loop seal. When Gs increases with increasing aeration rate (Q), at fixed velocity in the riser (U) and bed inventory (TSI), the mass of solid and thereby solid height in the standpipe decrease because more solids accumulate in the riser. At the same time, the pressure gradient in the standpipe increases. This is a special feature of transient packed bed flow [6], which is related to the pressure gradient and gas-solid slip velocity U_{sl}. 
According to the experimental measurement of voidage by laser fiber and gas flow rate measured by gas tracer, the slip velocity $U_{Sl}$ can be calculated by the equation $U_{Sl} = \frac{G_S}{\rho_p(1-\varepsilon)} + U_g/\varepsilon$ [6]. With increasing aeration rate, particle velocity $U_S$, increases due to higher $G_S$. At the same time, the upward gas flow rate in the standpipe, $Q_V$, also increases. Therefore, the slip velocity will increase with increasing aeration rate. Although the solid height decreases, the solid seal can provide a pressure head because of the increasing pressure gradient and slip velocity $U_{Sl}$. With increasing aeration rate, the upward gas flow, $Q_V$, keeps on increasing and the voidage gradually approaches to the minimum fluidization voidage. With very high aeration rates, upward flowing bubbles can be visually observed and the flow reaches the bubbling state.

The test facility [3] had riser diameter of 0.07 m and height 4 m. Cross section of standpipe was 35*50 mm. The aeration nozzles, located only in the supply section of the loop seal are inserted sideways at angle of 45°, to prevent particles clogging the nozzles. The fluidization gas was air and particles are Ilmenite mineral particles with a particle size distribution of 100-200 μm and a mean particle size of 143 μm. The particle density ($\rho_p$) is 4400 kg/m³ and the Geldart classification of particles is class B. The minimum fluidization velocity $u_{mf}$ was equal of 0.029 m/s. The parameters varied are the lo of 0.003 to 0.28 m/s and riser velocity 3 to 5 through stopping loop seal aeration and measure be height with the standpipe for a given period of time.

The flow travelling through the particle column in the standpipe (gas velocity) can be calculated by gas flow rate, cross section and voidage in the standpipe. If it is positive than the loop seal aeration flow is split between the loop seal supply chamber-standpipe and the recycle chamber-riser. This split is defined as the fraction of total volumetric flow of the loop seal aeration entering supply side or the recycle side of the loop seal. The aeration split increases with increase in relative loop seal aeration and is calculated from - 8% to +6%. Up to 6% of the aeration gas is entering the standpipe. This concludes that remaining 94-100% of the aeration gas is entering the recycle chamber. In case of negative gas velocity additional gas is entering the recycle side of the loop seal. If the calculated aeration in the recycle side is in the range of 2-6 minimal fluidization and enough to keep the recycle side of the loop seal fluidized. Therefore the loop seal worked well even without the aeration in the recycle chamber (lifting part). The deviation bars in Fig.2 and Fig.3 show the influence of voidage in the calculations of gas velocity and aeration split.

![Figure 2](image-url)
The lower deviation shows the value at a voidage of 0.48 and the upper deviation shows the value of voidage 0.54 close to minimum voidage. As observed the voidage can affect the results significantly. The assumption of a constant pressure gradient holds true to a limited extent. Therefore, to find out the exact gas flow pattern would implicate the use of tracer gases. Johansson et al [7] reported aeration split values of 2 to 7 % using tracer gases. In that work, for a separate downcomer of the same facility gas velocity values of -0.05 m/s to + 0.1 m/s were recorded.

![Figure 3](image_url)

**Figure 3.** Effect of loop seal aeration on the aeration split data [3]

A cold prototype of chemical looping system (CLC) was presented in [4]. Particles real density was 2650 kg/m³, bulk voidage 0.46, minimum voidage 0.514, minimum fluidization velocity 0.068 m/s, Sauter diameter 0.321 mm. In [4] a gas tracer (Helium) was used to study solid and gas flow in an L-valve. Effect of solid flow rate and pressure drop variation on the quantity of gas in the vertical section of the L-valve is presented. Results were then used to calculate the voidage of the moving solid bed in the L-valve vertical section. Tracer gas was injected into the fluidization gas of the reactor with low concentration of 0.33 vol. %. Low inlet concentration of tracer gas was used to ensure that the tracer gas does not modify significantly properties of the fluidization gas (air). Helium was then detected in the cyclone gas exit. The packed bed solid flow regime was reached when the pressure drop across the standpipe was negative, corresponding to positive slip velocities. The slip velocity is positive in this regime, as gas downward flow is faster than solids velocity. As slip velocity increases in value, voidage of solid bed expends from tapped bed voidage toward bed voidage at the minimum fluidization condition. Linear correlation proposed by Knowlton and Hirsan [8] results in closest prediction.

2 Experimental setup and methods

To study the hydrodynamics of interconnected CFB and FB reactors the experimental setup was constructed. The main elements are CFB and fluidized bed reactor (FB) reactors with associated overflow system. A detailed description of the test rig and layout of reactors was given in [9]. CFB reactor is a vertical column with cross-section 0.2 × 0.3 m and 5.4 m height, to the top of column the inlet cyclone duct is attached. The air is discharged from the cyclone to the settling chamber, at the top
of which installed the removable filter. To the conical part of cyclone attached the riser with cross-section 0.1 × 0.1 m. In the middle part of the standpipe is installed shut-off rotary valve, which is used to determine the flow rate of material through the circulation loop. The riser is connected to the upper loop seal. The design of the loop seal allows releasing one part of the material directly to the CFB reactor, and the other part to the lower part of FB reactor through other standpipe with L-valve (44 × 94 mm cross-section and 420 mm length in horizontal part). FB reactor has a lower section with 0.28 × 0.2 m cross-section and a height of 0.5 m, a transition cone section and an upper section of 0.4 × 0.4 m cross-section and 1.5 m height. It is connected to pipe with loop seal placed in the conical part of reactor and providing feed of the material to lower section of CFB reactor.

Properties of sand particles used in the study: real particles density 2550 - 2620 kg/m³, bulk density 1570 - 1520 kg/m³, vibration bed (tapped) density 1690 – 1670 kg/m³, Sauter diameter 0.323 and 0.17 mm.

During the research, solid flow rates in the standpipe under cyclone were measured with a cut-off valve. The flow rates of all air flows were measured using pre-calibrated flow-rate orifice plate and rotameters. Studies were conducted in the range of fluidization numbers from 1 to 4. For large fluidization numbers, gas tracers were not used, and the slip velocity under the conditions of the loop valve was determined from data on the pressure gradient. To study the gas flow behavior in the standpipe, high purity CO₂ gas was used as the tracer. CO₂ was injected into the system at Point A, and CO₂ concentration at Tap 1, 2 and 3 (located on the line of standpipe) were simultaneously measured by a CO₂ detecting system with 3 channels, each equipped with a sampling probe and a CO₂ sensor.

3 Experimental results and discussion

There are two regimes of down flow moving bed: packed bed and transition packed bed flow depends on negative or positive slip velocity. If the slip velocity is negative, flow is in packed bed mode with voidage equal to vibrated bed. If the slip velocity is positive, transitional mode exists in which voidage increases with slip velocity. Slip velocity connects with superficial solids and gas velocities and voidage in the standpipe as:

\[ U_{sl} = U_s + U_g = \frac{G_s}{F \rho_s (1 - \varepsilon)} + \frac{G_g}{F \rho_g \varepsilon} \] (1)

It is known that the considerable part of air goes with a material stream to the upward part of the pneumatic valve or to horizontal part of the L-valve. The expense of a material is measured, and interconnection between voidage, slip velocity and pressure gradient is defined by Ergan’s [6] equation (2):

\[ \left( \frac{\Delta P}{L} \right) = \frac{150 \mu_g}{d_p^2} \cdot U_{sl} \cdot \left( 1 - \frac{\varepsilon}{e} \right)^2 + \frac{1.75 \rho_g}{d_p} \cdot U_{sl}^2 \left( 1 - \frac{\varepsilon}{e} \right) \] (2)

\[ \frac{\Delta P}{L} = a_{1} \cdot U_{sl} \cdot \left( 1 - \frac{\varepsilon}{e} \right)^2 + b_{1} \cdot U_{sl}^2 \left( 1 - \frac{\varepsilon}{e} \right) \]

For voidage in the standpipe, Knowlton [8] offered a simple linear dependence:

\[ \varepsilon = \varepsilon_v + (\varepsilon_{mf} - \varepsilon_v) \frac{U_{se}}{U_{mf}} \] (3)

Dependences (2) and (3) lead to the cubic equation for slip velocity versus pressure gradient. Calculation for mean square value of voidage can be more precisely executed. Then to the equation, it can be reduced to square, and the slip velocity will be equal:
Knowing the slip velocity at the measured pressure gradient it is easy to calculate the voidage in the pipe and the share of air upward rate, without using a method of gas tracers.

Initially, in our experiments on the setup described in [10], the gas tracer method was not used, so it was important to correctly process these modes when air was supplied only under the loop standpipe (shutter riser). The known pressure drop in the lifting part of the valve was determined porosity in the regime of a bubbling fluidized bed. To estimate the flow rate of gas entering the valve, the air contained in the initial particle flow was also taken into account. Figure 2 shows the processing of the experimental data of our studies using a J-valve with an average diameter of about 0.8 mm and a true density of 1420 kg/m$^3$ (minimum fluidization velocity of 0.175 m/s) as a function of the calculated slip velocity (positive direction - down) from the relative velocity of air supply to the loop.

From these data it follows that the regime of movement in a dense bed in the riser occurs when air is supplied up to about two fluidization velocity, the transition mode corresponds to the air flow at a velocity of 2 - 5 minimal fluidization, then the movement mode in a fully fluidized state sets in. These results explain the fact that the nature of the dependence of the relative bed level in the standpipe on the relative air velocity (Figure 5) noted in [8, 9, 10].
Figure 5. Relative standpipe bed material height level dependence of relative air velocity (air blown to the standpipe). 1 – experimental data [1], 2 – VTI experimental data [10]

There is a master curve of bed height to standpipe relative fluidization air velocity represented on Figure 5. There is also a similar curve (T. Knowlton, 1988 [1]), which says that if an air velocity is about 5 velocities of minimum fluidization equal, there will be reached a minimum fluidization bed height level. If fluidization number is less than 4 there will be increased relative bed material level with decreasing of velocity. If fluidization number is more than 6, then this material level will be increasing with gaining velocity.

Consider the data from Figure 5 it becomes clear that it is necessary to maintain the bed level in the standpipe and watch it not to exceed 0.4 of its height while limitation of air blown to the standpipe at level of 3-6 fluidization number.

The results of studies using CO₂ tracer generally confirmed the above provisions. They were carried out at various inventory and air velocities in the riser. The inventory in the riser often changed due to changes in the standpipe bed level. This provided a range of particle velocity in the standpipe from 0.018 to 0.1 m/s. The slip velocity ranged from 0.01 to 0.09 m/s, and the fluidization numbers varied within relatively narrow limits (2.6 - 3.6). As in [2, 4], most of the air (more than 90%) comes to the lifting part of the valve and then to the riser. However, a significant effect of the bed level in the standpipe on the air fraction (aeration split) was found (Figure 6). At low bed levels, a fairly large portion of the supply air moves up into the standpipe.
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Figure 6. Effect of bed height in the standpipe on aeration split

It is necessary to pay attention to the fact that in all the considered papers, the fluidization number 4 is critical for the operation of the valves. This indicates the fact of flow modes change, at low velocities – there is a dense bubbling downflow mode and at high velocities air bubbles move upwards, counter flow fluidized particles. Within recommended modes throughput of the standpipe has reached 1800 t/h·m² (there was no opportunity to reach higher values because of cyclone overload). For standpipe diameter calculation there should be used recommended value of downflow bed material velocity ≈ 0.1 m/s.

Conclusions

To estimate the proportion of gas entering the riser in the mode of movement in a dense bed, experimental studies were conducted using the method of gas tracers. The results of experiments are presented and compared with known data [2, 3, 4].

Lowering movement of the riser can occur in the mode moving down a dense bed or in the transitional regime, when a large leakage flow of the fluidizing agent (gas, steam, air) in the riser and possible mode bubbling fluidization, which can lead to leakage of air into the cyclone. The limits of the modes depend on the slip velocity (the sum of the material and gas velocities with different signs of the direction of movement). Dependencies are given to determine slip velocity.

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Acknowledgments

The article was prepared in the form of a generalization of the work carried out in connection with the centenary of the founding of the All-Russia Thermal Engineering Institute (VTI).
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