The Influence of Particles Flow Rate on Separation and Pressure Drop in Cyclones and Down Flow Behavior in Recirculating Systems

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Abstract: Circulating fluidized bed (CFB) plants are widely used in energy, petrochemicals, aluminum production and other industrial sectors. The 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 recirculating system for returning trapped particles to the reactor. In fact, it is the cyclone capture efficiency that directly determines the multiplicity of circulation and the flow rate of circulating particles and their size characteristics. The dependence of the correction on the mass concentration of particles is proposed. A brief description of the test rigs and the results of our own studies of particle separation in cyclones are given. Experimental data on the effect of dust on the capture efficiency are presented in the form of a dependence of the relative entrainment of particles (1-efficiency) on the mass concentration of particles at the entrance to the cyclone. Recirculating 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. 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.

Key-Words: - circilated fluidized bed, hydrodynamics, solids separation efficiency pressure drop, recicrulating systems, down flow behavior, pneumatic valve

1 Introduction

The most important areas of modern research in the field of CFB technology as applied to coal combustion is a comprehensive study of the issues of the hydrodynamics of the furnace, the collection of particles in separators and their return to the furnace. Particulate capture efficiency is fundamental to increasing the flow rate of circulating particles and retaining small calcified particles in the circulating material. A high concentration in the above-bed space of the furnace leads to an increase in the heat transfer coefficient and ensures a uniform temperature distribution in furnace height, better efficiency of SO_2 capture, as well as a more efficient combustion. Not only is the overall capture efficiency important, but the fractional efficiency is also important. High fractional efficiency of small particles allows to significantly reducing the size of circulating particles, thereby increasing their concentration in the reactor and the conductive heat exchange to the furnace walls. Numerous studies have been carried out recently in China, which allowed creating own design of high-performance separators and CFB boilers [1]. A new cyclone design has been developed with an offset longitudinal outlet and a narrow, inclined inlet. The average size of the captured particles was less than 0.130 mm. As a result, the proportion of large particles in the bed became significantly less than in traditional CFB boilers. The proportion of small particles was more than half of the bed mass.

Cyclones pressure drop with high dust content is significantly reduced compared to a non-dusty stream. Data from our and other studies are reviewed and the dependence of the correction on the mass concentration of particles is proposed. The issues of increasing the fractional efficiency due to the installation of the exhaust pipe eccentrically to the cyclone axis are considered [2]. The cyclone capture efficiency was increased to more than 99.9 %. Such a high efficiency is associated with relatively large particles, narrow fractional dispersion and high concentration [3].

The key element of CFB boilers is the recirculation system. These systems consist of standpipe equipped in the lower part with pneumatic valve for transporting solids 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 to the furnace (L-valve). The downward movement can occur in the mode of a dense bed moving downward or in a transient mode, with a high flow rate of the fluidizing agent into the standpipe. A mode with bubbling fluidization is also possible, but it leads to the entry of air into the cyclone. The boundaries of the modes depend on the slip velocity (the sum of the speeds of the material and gas 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 standpipe from the total air flow supplied to the lower part (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 [4 - 8]. 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 [8]. 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 [5, 6, 7].

2 Effects of Solids Loading on Cyclone Performance and Down Flow Behavior

2.1 Cyclone performance

At low solid-to-gas mass loading ratios solid particles are clustered as thin strands on the cyclone walls and are transported in the downward direction in a spiral path, while at high mass loadings, a major portion of wall surface area is covered with a layer also known as a dense strand [15]. For the calculation of the solids separation in the gas cyclone the model of [16] has been taken as a basis very often. In this model the separation mechanism in the cyclone is divided into two parts. At first, immediately near the inlet that part of the solids loading, which exceeds a limiting value is forming a strand, which flows directly underflow. The remaining part of the solids is undergoing the separation in the vortex. Both mechanisms are described by semi-empirical correlations. The improvement of separation efficiency with the cyclone Reynolds number can be observed from the experimental data of [17] for a fixed value of particle kinematic residence time. Despite available experimental studies on the effect of solids loading on the separation efficiency, there is no consensus regarding the exact mechanism of this improvement [17]. Limiting value of the solids loading μg in different studies differ significantly – from 0.005 to 0.05.

In [13] complex dependences are given for this quantity. The calculation results given in [18] showed that an increase in the average particle size leads to growth of μg. For large solids loading, the limit values are higher. With increasing cyclone size and temperature, the ultimate value of the solids loading increases.

2.2 Down Flow Behavior

In [5] was shown that with increasing aeration rate, the pressure drop of the L-valve horizontal section increases due to higher solid circulation flux (Gs). 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 [19], which is related to the pressure gradient and gas-solid slip velocity U_{Sl}. Other investigation of down flow behavior was made in [6]. The split of gas flow 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 to 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. The deviation bars in Fig.1 show the influence of voidage in the calculations of gas velocity and aeration split.

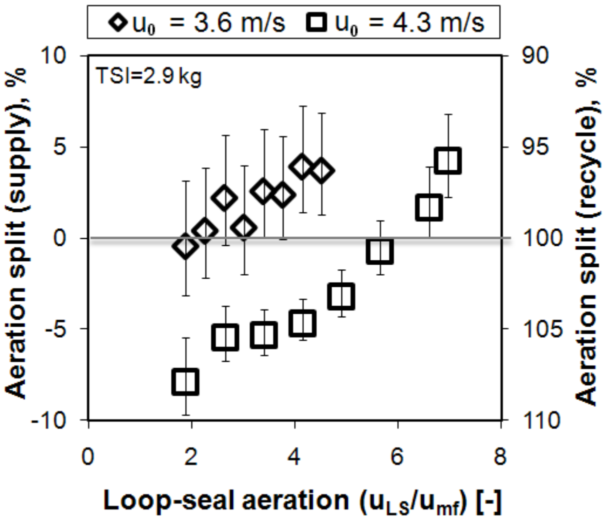

Figure 1. Effect of loop seal aeration on the aeration split, data [6]

In [20] a gas tracer (Helium) was used to study down flow behavior 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 was presented. Tracer gas was injected into the fluidization gas of the reactor with low concentration of 0.33 vol. %. 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, voidage expends from tapped bed voidage toward minimum fluidization. Linear correlation proposed by Knowlton and Hirsan [21] results in closest prediction.

3 Experimental setup and methods

A detailed description of the test rig and layout of reactors was given in [22]. 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 riser is installed shutoff 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 the riser with L-valve with 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.

The cyclone design is shown in Fig. 2. To increase the efficiency of capturing fine particles, the exhaust pipe of cyclone is installed eccentrically with a displacement from the inlet duct [23].

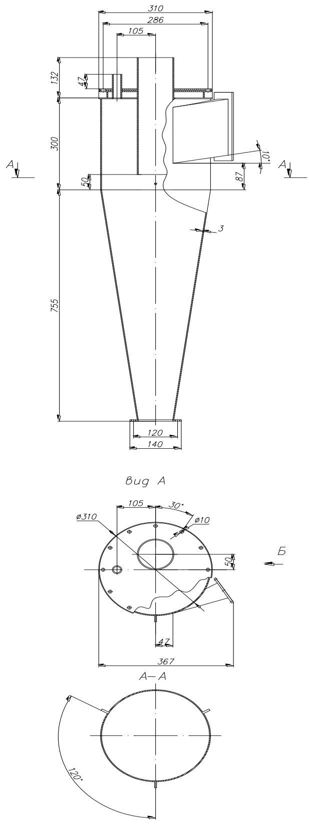

Figure 2. Cyclone design

During the research, solid flow rates in the standpipe riser 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. The pressure drop of cyclone was determined from the pressure differences between the upper part of CFB reactor and the exhaust pipe before entering fine filter, in addition, the pressure drop across sections of reactor - the cylindrical part of cyclone and cylindrical part of cyclone - exhaust pipe were measured. The amount of entrainment from cyclone was determined by the weight method. After 1 hour of operation, the accumulated material was unloaded and weighed. The samples of entrainment and circulating material were subjected to the determination of the fractional composition after each experiment. The small particles of entrainment were analyzed on a Fitch device.

Properties of sand particles used in the study: real particles density 2550 - 2620 kg/m^3, bulk density 1570 - 1520 kg/m^3, vibration bed (tapped) density 1690 – 1670 kg/m^3, Sauter diameter 0.323 and 0.17 mm. 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 CO_2 gas was used as the tracer. CO_2 was injected into the system at Point A, and CO_2 concentration at Tap 1, 2 and 3 (located on the line of standpipe) were simultaneously measured by a CO_2 detecting system with 3 channels, each equipped with a sampling probe and a CO_2 sensor.

4 Experimental results and discussion.

4.1 Separation and Pressure Drop in Cyclones

The capture efficiency in cyclone proved to be very high - more than 99.9%. Therefore, it is better to analyze the influence of factors on the full efficiency of the cyclone through the amount of entrainment from the cyclone. In Figure 3 shows the dependence of relative entrainment on mass concentration. Mass concentration, that is, the ratio of the flow rate of solid particles to the total flow rate of solid particles and gas, in our opinion, is the best criterion in comparison with mass loadings. This approach is often used in the analysis of data on two-phase flows. At μ = 0 it is a gas flow, at μ = 1 it is a heavy phase flow. In [24], the calculations of fractional capture efficiency were performed for 4 models. The first two models are based on force balance and the following two on a combination of force balance and residence analysis. For most options, calculation model [25] gives maximum efficiency values. Approximately 1.5 times smaller is the diameter of particles captured by 50% compared with other models.

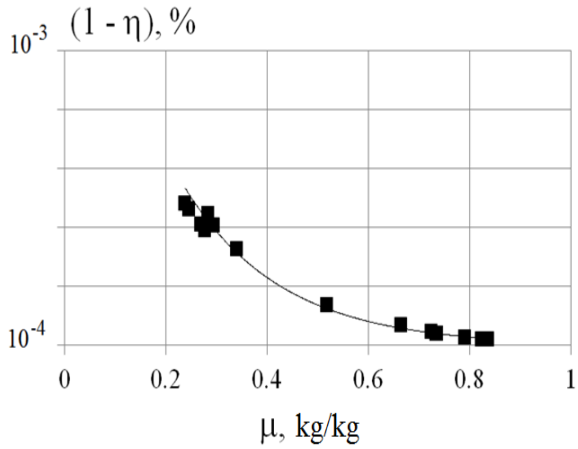

Figure 3. The mass concentration of a stream against relative entrainment of particles (1-efficiency)

Fractional efficiency of the cyclone is determined by [26]. This method is based on the three-zone model proposed in [27] and got its development in [28] for high-temperature cyclones. In this model, three zones are considered: the entrance zone, the down flow and the lifting flow. Turbulent mixing is determined by the profile of radial concentrations in each zone, taking into account the exchange of particles. The analysis includes the real geometry of the cyclone, takes into account the exchange of particles between zones 2 and 3 and the distribution of the relaxation time. In zones 1 and 2, gas and particles move downward, and in zone 3, the gas velocity vector is directed predominantly upward. The proposed in [27] system of equations for each zone under the appropriate assumptions and transformations performed by the authors leads to the following set of dependencies for the calculation of fractional efficiency:

$$\eta_i = 1 - (\boldsymbol{k}_0 - \sqrt{\boldsymbol{k}_1^2 + \boldsymbol{k}_2}) \cdot \exp(-\boldsymbol{f}(\boldsymbol{d})) \qquad (1),$$

where k_0, k_1 and k_2 – depend on the ratio of the diameters of the cyclones and the exhaust pipe, as well as on the function of the particle diameter.

The function of the particle diameter $f(d)$ was determined on the basis of generalization of data on high-temperature cyclones and own experimental data. As a result of processing this data, the following dependence was obtained:

$$\boldsymbol{f(d)} = 0.565 \cdot \left(\frac{\boldsymbol{h}_e - \frac{\boldsymbol{a}}{2}}{l} \right)^{0.44} \cdot \left(\frac{t_g + 273}{293} \right)^{0.3} \cdot \left(\frac{\boldsymbol{d}_i}{\boldsymbol{d}_p} \right)^{1.04} \qquad (2),$$

where h_e is the height of the exhaust pipe, m; t_g – gas temperature in the cyclone, °C; d_i and d_p – current and calculated particle size, m.

The calculated particle size is determined by the formula:

$$d_p = \sqrt{\frac{9 \cdot \mu_g \cdot a \cdot b}{\pi \cdot H_c \cdot \rho_s \cdot U_e}} \qquad (3),$$

where H_c – height of the cylindrical part of the cyclone,m; U_e – gas velocity at the cyclone inlet, m/s; a – the height of inlet duct, m; b – the width of inlet duct, m.

It should be noted that dependence 2 was obtained by processing numerous data on fractional efficiency, including for high dust conditions, as well as for industrial cyclones. Thus, the calculation to some extent takes into account the influence of solids loading. A traditional approach is used to determine the overall capture efficiency:

$$\eta = \frac{1}{2}\left[1+\varphi(x)\right]\cdot 100 \qquad (4),$$

where $\varphi(x)$ is the probability integral (tab. value) at

$$x = \frac{lg\,\frac{d_m}{d_{50}}}{\sqrt{lg^2\,\sigma\eta + lg^2\,\sigma_ч}} \qquad (5),$$

where d_m is the average particle diameter at the inlet, m; $lg\,\sigma_ч$ - standard deviation in the distribution function of particles at the entrance to the cyclone.

The values d_m и $lg\,\sigma_ч$ are determined by the construction in normal-logarithmic coordinates from the known fractional composition of particles at the entrance to the cyclone.

Comparison of calculated and experimental data on the total capture efficiency showed that the calculated values are always lower than the experimental ones. Noticeable differences are observed when m is greater than 0.5. The difference is only 0.004 - 0.005%. Thus, we can conclude that for cyclones with very high efficiency, the inclusion of the mass concentration at the inlet lies within the possible errors of both calculation and experiment.

It is known that the pressure drop of a cyclone decreases with increasing mass concentration. The presence of solids in the cyclone, forming particle strands that descend to the collection device (dipleg or hopper) produces this effect in cyclones. For low-to-medium inlet solid loadings and constant geometry, cyclone non-dimensional pressure drop (Euler number) decreases monotonically with solids concentration. Nevertheless, the curve has a minimum (a critical solid inlet loading) and for very high solid loadings (such as in CFB) the opposite process has usually been detected (Fig. 4). The cyclone pressure drop can be calculated by following folmula:

$$\Delta P_c = \psi_c \cdot \xi_c \cdot \frac{U_c^2 \rho_g}{2} \qquad (6),$$

where ψ_c is correction for the influence of inlet flow concentration, ξ_c is the cyclone resistance coefficient, reduced to the velocity in it (U_c).

To calculate the correction for solid concentration, several dependencies are proposed. The most representative form of dependence is proposed in [29]. This form gives a correction value of 1 for a pure gas flow, with a minimum correction at mass concentrations (as solid flow rate to summ of solid and gas flow rates - μ) of about 0.3 and a further increase in its value with increasing concentration. In a paper [30], a series of dependencies are analyzed to calculate the correction for mass concentrations. Experimental data [30] showed the absence of influence of gas velocity and particle sizes. Figure 4 shows our data and data [30]. They are summarized by a formula:

$$\psi_c = \frac{1}{25{,}8 \cdot \mu^{1{,}71} + 1} + 0{,}72 \cdot \mu \qquad (7),$$

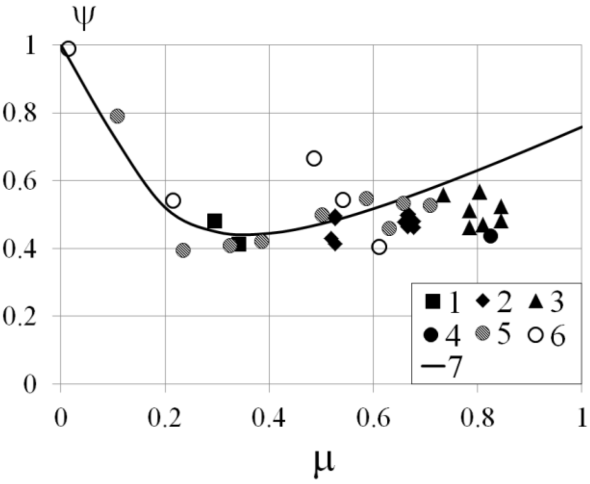

Figure 4. The correction on dust content of the flow against mass concentration of particles

1 – Velocity in CFB reactor: U = 3.06 - 3.1 m/s; 2 - U = 3.55 - 3.8 m/s; 3 - U = 4.29 - 4.42 m/s; 4 - U = 4.7 m/s; 5 - data [30] (d_p = 0.08 -0.111 mm); 6 - data [30] (d_p = 0.111 - 0.5 mm); 7 - formula (7)

4.2 Down Flow Behavior in Recirculating Systems

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:

$$U_{sl} = U_s + U_g = \frac{G_s}{F\rho(1-\varepsilon)} + \frac{G_g}{F\rho_g \varepsilon} \qquad (8)$$

The expense of a material is measured, and interconnection between voidage, slip velocity and pressure gradient is defined by Ergan's [19] equation:

$$\left(\frac{\Delta P}{L}\right) = \frac{150\mu_g}{d_p^2} \cdot U_{sl} \cdot \left(\frac{1-\varepsilon}{\varepsilon}\right)^2 + \frac{1{,}75\rho_g}{d_p} \cdot U_{sl}^2 \left(\frac{1-\varepsilon}{\varepsilon}\right)$$

$$\frac{\Delta P}{L} = a_1 \cdot U_{sl} \cdot \left(\frac{1-\varepsilon}{\varepsilon}\right)^2 + b_1 \cdot U_{sl}^2 \left(\frac{1-\varepsilon}{\varepsilon}\right) \qquad (9)$$

For voidage in the standpipe, Knowlton [21] offered a simple linear dependence:

$$\varepsilon = \varepsilon_v + (\varepsilon_{mf} - \varepsilon_v) \frac{U_{se}}{U_{mf}/\varepsilon_{mf}} \qquad (10)$$

Dependences (8) and (9) 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:

$$U_{sl} = \frac{-a_1\left(\frac{1-\varepsilon_v}{\varepsilon_v}\right)\cdot\left(\frac{1-\varepsilon_{mf}}{\varepsilon_{mf}}\right) + \sqrt{a_1^2\left(\frac{1-\varepsilon_v}{\varepsilon_v}\right)^2\cdot\left(\frac{1-\varepsilon_{mf}}{\varepsilon_{mf}}\right)^2 + 4b_1\left(\frac{1-\varepsilon_v}{\varepsilon_v}\right)\cdot\left(\frac{1-\varepsilon_{mf}}{\varepsilon_{mf}}\right)\left(\frac{\Delta P}{L}\right)}}{2b_1\left(\frac{1-\varepsilon_v}{\varepsilon_v}\right)\cdot\left(\frac{1-\varepsilon_{mf}}{\varepsilon_{mf}}\right)}$$

(11)

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 [26], 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. The known pressure drop in the lifting part of the valve was determined porosity. To estimate the flow rate of gas entering the valve, the air contained in the initial particle flow was also taken into account. Fig. 5 shows the processing of the experimental data of our studies using a J-valve (particles average diameter of 0.8 mm, true density of 1420 kg/m ,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.

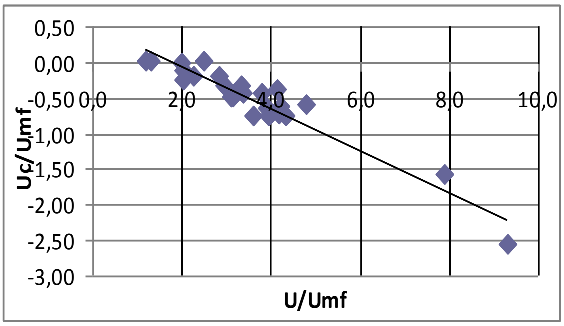

Figure 5. Dependence of the relative slip velocity in the standpipe on the relative velocity of the total air flow

The regime in a dense bed in the standpipe 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 U_{mf}, 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 noted in [21, 26, 31]. A master curve of bed height to standpipe relative fluidization velocity represented on Fig. 6.

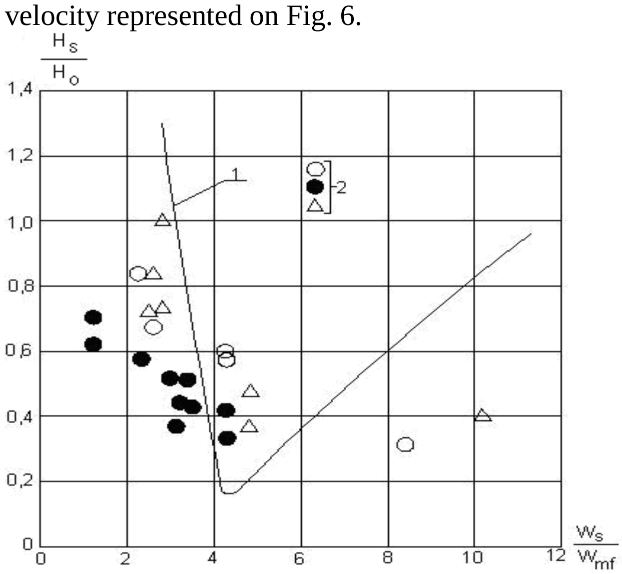

Figure 6. Relative standpipe bed material height level dependence of relative air velocity (air blown to the standpipe). 1 – experimental data [4], 2 – VTI experimental data [26, 31]

There is also a similar curve [4]), 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 velocity. Consider the data from Fig. 5, 6 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 CO_2 tracer generally confirmed the above provisions. 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 [5, 7], 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 (Fig. 7).

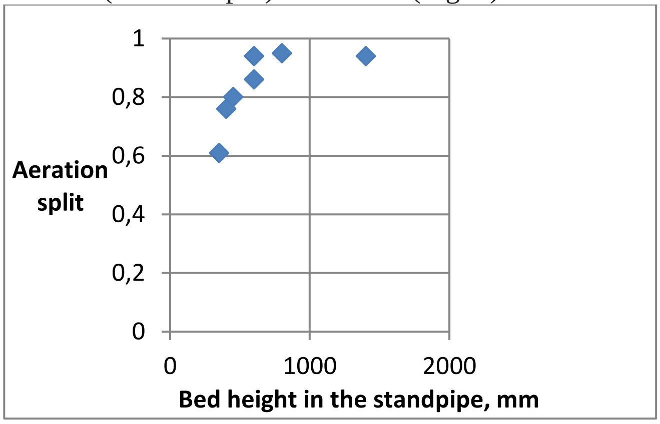

Figure 7. Effect of bed height in the standpipe on aeration split

At low bed levels, a fairly large portion of the supply air moves up into the standpipe.

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. Within recommended modes throughput of the standpipe has reached 1800 t/h·m^2 (there was no opportunity to reach higher values because of cyclone overload). For standpipe diameter calculation there should be used recommended value of down flow bed material velocity ≈ 0.1 m/s.

5. Conclusions

The effect of mass loading at the inlet of the cyclone on the fractional capture efficiency is considered. According to experimental results, this effect is especially large at a mass concentration of more than 0.5. The calculated values of the total capture efficiency are very large, which is associated with both a long residence time and a narrow range of particle sizes with their high average diameter. Comparison of calculated and experimental data on the total capture efficiency showed that the calculated values are always lower than the experimental ones.

The presence of solids in the cyclone, forming particle strands that descend to the collection device (dipleg or hopper) produces this effect in cyclones. To calculate the correction for solid concentration, new dependence is proposed.

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. 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.

A significant effect of the bed level in the standpipe on the air fraction (aeration split) was found. At low bed levels, a fairly large portion of the supply air moves up into the standpipe. 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 down flow mode and at high velocities air bubbles move upwards, counter flow fluidized particles.

Within recommended modes throughput of the standpipe has reached 1800 t/hm2 (there was no opportunity to reach higher values because of cyclone overload). For standpipe diameter calculation there should be used recommended value of down flow bed material velocity near 0.1 m/s.

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