Gas-particles Flow Transitions for High Density Powder

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Abstract—**The aim of this study is to get a better comprehension of gas-particles fluidization regimes. Experiments have been conducted in an experimental Circulating Fluidized Bed (CFB) with solid particles of high density (bronze) compared to those generally encountered in literature (sand, glass beads or FCC). Analysis of the standard deviation of pressure drop fluctuations normalized with pressure drop amplitude and visual observations have shown a choking limit and a bubbly-turbulent transition. We have shown that correlations of literature can well predict the choking limit and experimental specific velocities identified for turbulent transition, thus extending their application range.**

Index Terms—**fluidization, regime transition, choking, turbulent flow.**

I. Introduction

THE literature proposes numerous correlations associated with regime transitions, most of them developed for non-circulating fluidized beds and then applied to Circulating Fluidized Bed (CFB). However, in some cases it is apparent that the authors disagree concerning the types of regime on either side of identified transitions. This can be due to the disparity of the geometrical aspect of CFBs (with particular reference to riser section) from one author to another, although the powder used and the operating conditions are very similar.

Bai and co-authors [1] stressed the need for clear identification of the fluidization regimes in the riser of a CFB, to ensure better comprehension of the thermohydraulic context, and thus correctly design the loop. They emphasize the ambiguity of definitions proposed in the literature, both in qualitative and quantitative terms, and demonstrate a number of contradictions between theoretical predictions and experimental results.

Furthermore, the available correlations for CFBs have been developed for FCC type powders which density is around 2000 kg/m^3 and for which extrapolated studies concern at least densities less than 5000 kg/m^3 [2]. For example, among the recent compilation made by Yang & Leu [3] gathering 29 experimental cases, only 2 of them involve powder density higher than 5000kg/m^3 (less than 7% of all cases). As a consequence, extrapolation of fluidization regimes observations and results appears to be not so easy. Yet, high density powders such as bronze are of interest since bronze and air show the same density ratio than ashes and hot gas inside industrial CFB furnaces. High density powder can thus help to work with better hydrodynamic similitude.

We therefore decided to carry out experimental identification of regime transitions for those types of gas/particle flow with which it is concerned. Analysis of pressure drop fluctuations in the riser was conducted as done by others (see, for example, [4]-[6]), in order to identify the various fluidization regimes. Identified transition velocities have been confirmed by visual observations and compared with existing correlations which we consider to be the most appropriate for the studied flows. The transition velocities are namely the choking velocity for which the bed collapses, the velocity U_c related to the onset of transition between bubbly flow and turbulent flow and U_k related to the onset of established turbulent flow.

A good comprehension of gas-particles fluidization regimes will contribute to give us the means to appreciate the possibility of extrapolating experimental CFB correlations to industrial units as done for example in a previous work [7], which have much larger scale than laboratory facilities.

II. Experimental Facilities and Protocol

A. Experimental facilities

The experimental loop is a CFB which internal height of riser is 6.61m from the fluidization grid to the roof, and a square section measuring 0.175m x 0.175m (see Fig. 1), with transparent perspex walls. The exit riser is located on one side of the riser, at the top of it, with a 0.175m x 0.13m section, and the exit pipe forms a 90 degrees angle with the riser axis. The bed is supplied with air at ambient temperature by a booster pump, via a fluidization grid. Powder is introduced at the bottom of the riser, through side wall. Evacuation of the gas/particles suspension is made through the same side wall, at the top of the riser. The powder distributor system is a screw feeder (80R type, Parimix). Gas/particles separation is obtained by means of a centrifuge cyclone, which leads the solid phase to the distributor, via a solid flow rate measurement device (Endress & Hauser impact plate type flowmeter). The solid flow rate in the loop is thus known continuously. The riser is equipped with 12 differential pressure transducers (deltabar PMD 130, Endress & Hauser), which deliver a 4-20 mA signal proportional to the measured pressure. They can be connected as required to 50 pressure taps, 6 mm diameter drilled through the perspex wall, uniformaly arranged on the center line of the riser wall, and two absolute pressure transducers, located at the top and bottom of the riser. The taps are exempt of filters, which would prevent solid

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particles from getting inside the measurement line, but would also reduce the transducers sensitivity to dynamic pressure fluctuations. Pressure measurement lines have been optimized as suggested by Clark et al. [8] in order to guaranty the best response of the lines when measuring pressure drop fluctuations. This implies to purge systematically all connected pressure taps before measurements and limits the maximum acquisition duration time. Fluidization velocity inside the riser is determined from the measurement of air mass flow upstream the riser (AT 533 thermal mass flowmeter, Endress & Hauser), temperature and pressure in the riser.

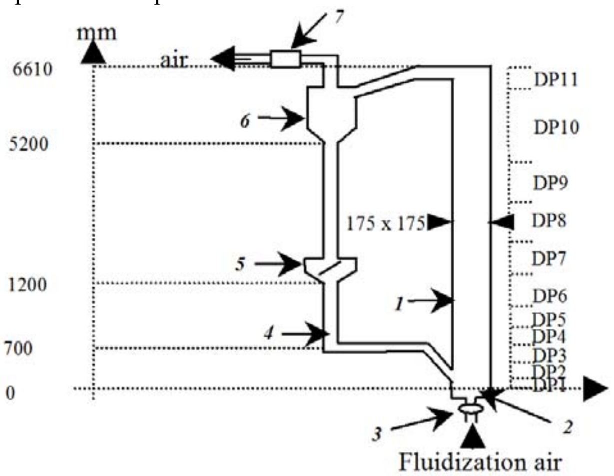

Fig. 1: Experimental device. 1: riser. 2: fluidization grid. 3: fluidization flowmeter. 4: solid screw feeder. 5: solid mass flowmeter. 6: separator. 7:filter.

Parameters of operating conditions are given in Table I. Bronze powder is used, which characteristics are also given in Table I. The particles are fluidized with air at ambient temperature. Bronze and air show the same density ratio than ashes and hot gas inside industrial CFB furnaces.

B. Experimental protocol

The differential pressure signals are acquired at 100Hz during 45 to 120 seconds, both for the riser segments defined between pressure taps, and for the total riser pressure drop. Acquisition duration has been adjusted regarding on the period of the pressure drop fluctuations observed.

TABLE I
EXPERIMENTAL CONDITIONS

Symbol	Value / Type	Description	SI
A	0.175 x 0.175	riser section	m^2
H	6.61	riser height	m
CuSn5	bronze	particle material	none
d_p	43.10^{-6}	mean particle diameter	m
Φ	1	sphericity	none
ρ_s	8790	particles density	$kg.m^{-3}$
G_s	0.14 to 6	solid mass flow	t/h
	1.25 to 54.42	solid mass flow rate	$kg.m^{-2}.s^{-1}$
U_0	1 to 9	superficial gas velocity	m/s
μ	$1.8\ 10^{-5}$	gas viscosity	Pa.s

A mean value and a standard deviation are calculated for each of these signals. The ratio of standard deviation on mean value (called normalized standard deviation and denoted NSD) is used for the flow transition analysis as suggested in previous works [9], [10] and then again by Johnsson et al. [11]. It is based on the method suggested first by Yerushalmi et al. [12] analyzing pressure drop fluctuations. NSD reflects variations in the amplitude of measured pressure fluctuations. By plotting evolution of this quantity versus fluidization velocity, it is possible to identify fluidization regime transitions (for example, see [3], [6], [13]-[15]).

Besides, visual observations were performed through the perspex walls of the riser to identify a change of the two-phase flow behavior with changing U_0. These observations were obtained in order to confirm fluidization regime transitions detected with NSD analysis.

Two types of test program were proposed. The first one involved constant solid flow rate in the loop, with a variable fluidization velocity, while the second one was associated with a constant pressure drop in the riser (namely a constant effective solid inventory in the riser) when the fluidization velocity varies.

III. RESULTS & DISCUSSION

A. Experimental results

Measurements of NSD at constant pressure drop in the riser did not show any detectable fluidization regime transitions. In all cases, we observed a growth of NSD with increasing fluidization velocity U_0.

NSD at constant mass flow rate G_s plotted versus U_0 were fitted with curves which present a maximum value for a specific fluidization velocity U_c, as shown on Fig. 2. For the five values of G_s investigated, these specific velocities have been identified as fluidization regime transition when compared with visual observations. Fig. 3 a and b illustrate the change of flow behavior for G_s = 5,2 t/h (47 $kg.m^{-2}.s^{-1}$) : at U_0 less than the transition velocity U_c, we can observe, through the bottom bed walls, the presence of clusters impinging the walls (white patterns on Fig. 3a) and of numerous trails spreading on the walls. Clusters and trails have no well defined displacement: some go up, others down, and it is clear that numbers of them have transverse velocity. These observations let us think of the existence of a splash zone over a dense bed, where the two-phase flow is the most turbulent. At U_0 greater than U_c, we no longer observe clusters impinging the walls (Fig. 3b). Only few trails can be seen. Furthermore, it is important to notice that, for the whole range of G_s and U_0 investigated, the bottom bed voidage ε was closed to 0,95 and visual observations have shown, close to the perspex walls and all along the riser height, the existence of a boundary layer where particles were flowing down to the bottom bed. These observations show that pneumatic transport has not been reached since such a fluidization regime is characterized by a dilute bed with particles all going upflow.

At constant mass flow rate G_s and for the five values of G_s investigated, we found a minimum fluidization velocity below which mass flow rate could not be maintained. When the minimum fluidization velocity was reached from upper values, the CFB flow, which was the most dense and turbulent we could observed, suddenly broke down: the bed collapsed. No solid recirculation was possible (except with increasing U_0) and the riser presented two distinct regions. The first one, in the bottom of the riser, was a dense bed which could be described as a bubbling bed with very few air bubbles. The second one, localized over the first one, was so dilute that it was hardly possible to detect solid particles entrained to the top of the riser. Thus, solid mass flow rate was zero. These results were completed by investigation of G_s down to 0.14 t/h (1.25 kg.m^{-2}.s-^{1}) in order to have a better description of this limit.

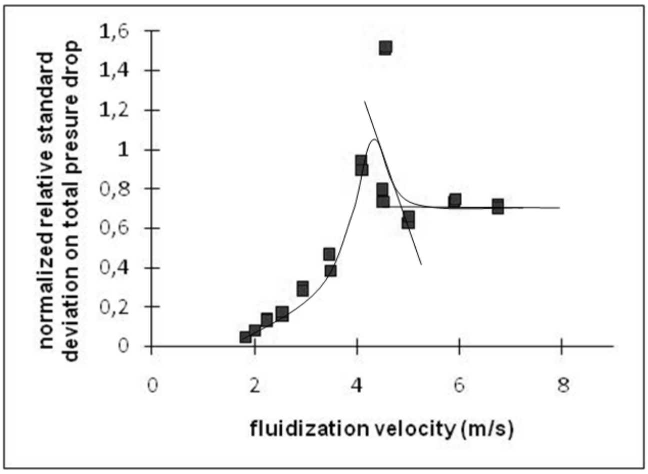

Fig. 2. Normalized standard deviation for total pressure drop in the riser versus fluidization velocity for G_s = 1 t/h (9.07 kg.m^{-2}.s^{-1}). Fitted curve and lines allow to identify transition velocities.

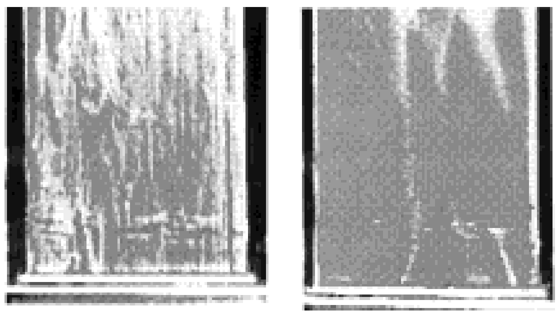

Fig. 3.a. U_0 = 3 m/s < U_c Fig. 3.b. U_0 = 7 m/s > U_c

Fig. 3.a & 3. b. Pictures a and b showing the change of flow behavior at the bottom of the riser for Gs = 5,2 t/h (47 kg.m^{-2}.s^{-1}).

In conclusion, two fluidization regime transitions have been observed at constant mass flow operating conditions : a bubbly-turbulent transition and, at lower fluidization velocity, a breakdown limit in terms of fluidization velocity called choking velocity U_{ch} at which sudden change from CFB to bubbling/fixed bed was observed, and under which constant mass flow rate could not be maintained. These transitions are discussed and analyzed in further sections.

The present results give average values for the characteristic velocities of the flow as following:
U_c = 4.5 m/s and U_k = 5 m/s.
As shown on Fig. 2, U_c is determined as the extreme of the NSD curve and U_k is the intersection of the extended lines of the NSD attenuation and plateau parts. As noticed by others [3], values of U_k are not easy to obtain and may sometimes not be identified at all.

B. Analysis & discussion

Observations and measurements performed in the present experimental program showed fundamental changes in the gas - particles flow behavior while U_0 was modified at constant G_s. For the high values of fluidization velocity U_0 investigated, and for constant mass flow rate G_s, a maximum of turbulence in the riser appears to correspond to a specific velocity U_c. Then, when decreasing U_0, a minimum fluidization velocity is reached, below which only bubbling or fixed bed is observed, consequently to a sudden breakdown of the flow. Those transition limits are discussed and compared with existing correlations in the following sections.

Transition from captive regime to turbulent regime

Transition from captive to fast transport regime involves turbulent regime, as observed by many authors [1], [6], [14]-[19]: with the increase in U_0, the fluidized bed in the riser changes from a dense condition to a bubbly condition, and turbulence inside the riser grows consequently. This increase in turbulence is accompanied by variations of density fluctuation amplitude. It has been verified [19] that these density fluctuations can be quantified by riser wall pressure measurement, comparing the signals delivered by wall pressure transducers with those delivered by optical probes. Thus the maximum standard deviation of total pressure drop in the riser corresponds to a maximum level of turbulence in the riser, associated with a specific velocity. This description agrees with results of the present work. The upper bound that we denoted U_c is related to the transition from bubbly to turbulent regime.

Above U_c, as U_0 is increased, turbulence then smoothly decreases while fluidization regime tends to fast transport type and can tend to pneumatic transport. The established turbulent regime is achieved for a characteristic velocity U_k.

Below U_c, another specific velocity is reached when decreasing U_0. This corresponds to a sudden breakdown of the bed with no possible solid recirculation. This specific velocity has been called by many authors "choking velocity" (see [20]-[22] for example). Bi et al. [18] have proposed a review of choking description, pointing out that some choking occurrence can be induced by equipment-limited operating modes, while others are due to the suspension characteristics. It has been verified, in the present work, that the choking transition obtained was not of the first type by comparing the pressure balance in the CFB loop and upstream the fluidization grid. The present choking transition is of the second type, the same as observed by Louge et al. [21].

Many correlations have been proposed for estimation of U_c and U_k, in the case of powders with ρ_s < 5000 kg/m3, fluidized in a pipe or a vessel without any solid recirculation. Most of them, expressed in terms of Archimede number Ar and Reynolds number Re, have been reviewed by Arnaldos et al. [23]. It is obvious that such correlations cannot fully describe CFB transition regimes as a main parameter as solid mass flow rate is ignored.

Yang, G. et al. [24], and, Bai et al. [1] have proposed correlations for specific transition regime velocities. Both of them express their relations according to operational parameters such as solid mass flow rate, riser dimensions, particles and gas properties, for operating conditions such as those indicated in Table I.

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For the onset of transition between bubbly flow and turbulent flow, Yang & Leu [3] have suggested a correlation for a wide range of density particle values, expressed in terms of Reynolds and Archimede numbers:

$$Re_c = 0.837\ .(Ar)^{0.487} \quad (1)$$

where

$$Ar = \rho_s\, g\, (\rho_s - \rho_g)\, d_p^3 / \mu^2 \quad (2)$$

$$Re_c = \rho\, g\, U_c\, d_p / \mu \quad (3)$$

When plotting our data on Fig. 4 with others results of studies [25]-[39] gathered in [3], the correlation coefficient of Yang & Leu's correlation value changes from 0.820 to 0.817. This allows to claim the validation of this correlation for a wide range of particle density extended to high density particles.

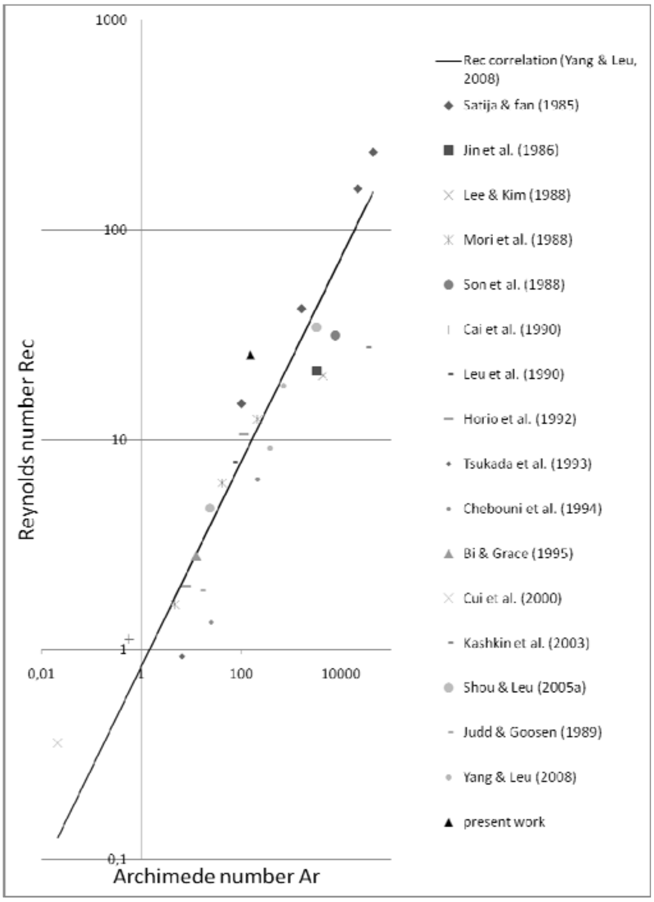

Fig. 4. Onset of transition between bubbly flow and turbulent flow: comparison between experimental velocity values of U_c (including air/bronze flow for the present work) with Yang & Leu's correlation [3]. Data extracted from [3],[25]-[39]. Correlation coefficient is 0.817.

For the onset turbulent flow, Yang & Leu [2008] have suggested a correlation for a wide range of density particle values, expressed in terms of Reynolds and Archimede numbers:

$$Re_k = 1.174\ .(Ar)^{0.481} \quad (4)$$

where

$$Re_k = \rho\, g\, U_k\, d_p / \mu \quad (5)$$

When plotting our data on Fig. 5 with others results of studies gathered in [3], the correlation coefficient of Yang & Leu's correlation value changes from 0.890 to 0.857. This allows to claim the validation of this correlation for a wide range of particle density extended to high density particles.

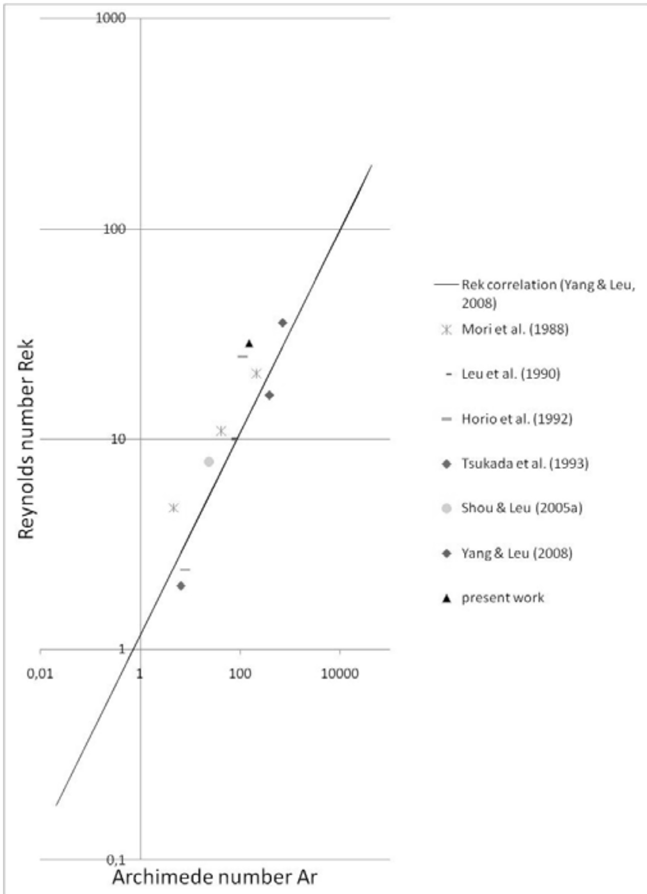

Fig. 5. Onset of turbulent flow: comparison between experimental velocity values of U_k (including air/bronze flow for the present work) with Yang & Leu's correlation [3]. Data extracted from [3], [28], [31]-[33], [38]. Correlation coefficient is 0.857.

Correlation for choking transition

For choking transitions, the review of Bi et al. [18] suggests that correlations of Yang [20] and Bi et al. [22] correspond to the type of choking encountered in the present work.

Yang 's correlation [20] is :

$$\frac{U_{ch}}{\varepsilon_{ch}} = U_t + \left\{ \frac{2\, g\, D \left(\varepsilon_{ch}^{-4.7} - 1\right) \rho_s^{2.2}}{6.81\ 10^5\, \rho_g^{2.2}} \right\}^{0.5} \quad (6)$$

with

$$G_s = \left(U_{ch} - U_t\right)\left(1 - \varepsilon_{ch}\right)\rho_s \quad (7)$$

and Bi 's correlation [22] is:

$$\frac{U_{ch}}{\sqrt{g\, d_p}} = 21.6 \left\{ \frac{G_s}{\rho_g U_{ch}} \right\}^{0.542} Ar^{0.105} \quad (8)$$

Louge et al. [21] has found out that Yang 's correlation (6) could be successfully applied to risers of large scale. Fig. 6 shows how data from the present work are bounded by these

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correlations (6) and (8). We find here conclusions of Bi et al. [18] who noticed the over-estimation of correlation (4), while (5) under-estimates the experimental data. The prediction of these correlations can be considered satisfactory.

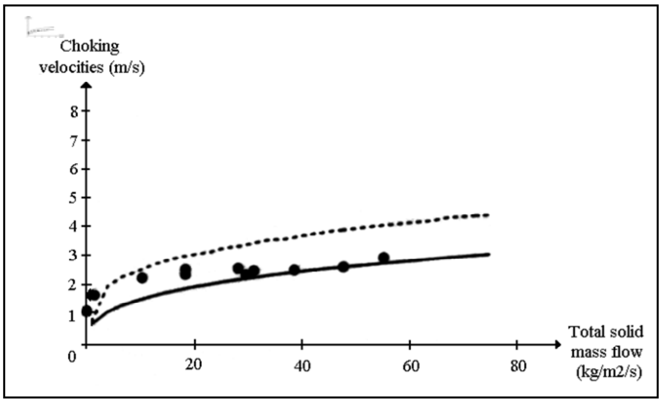

Fig. 6. Comparison of choking limits from Yang [20] and Bi et al. [22] correlations (6) and (8) with experimental choking data of the present work.

IV. Conclusion

Experiments have been conducted in a CFB with solid particles of high density (bronze) compared to those encountered in literature (sand, glass beads or FFC). Bronze was chosen to give with fluidization air the same density ratio than that of industrial CFB suspensions. Investigations have been performed for a large range of solid mass flow rate and fluidization velocity. Analysis of the standard deviation of pressure drop fluctuations normalized with pressure drop amplitude and visual observations have shown that i) a choking velocity is found at low fluidization velocity, associated with a sudden breakdown of the flow and no possible solid recirculation, ii) a bubbly-turbulent transition at higher fluidization velocity is identified, characterized by the maximum of normalized standard deviation of pressure drop fluctuations, iii) the onset of established turbulent flow is characterized.

Concerning the choking velocity, our data have been successfully bounded by existing correlation developed by Yang [20] and Bi et al. [22] as suggested by Louge et al. [21].

Concerning turbulent transition, application of the Yang & Leu's correlations [3] has been successfully extended to high density particles using the present experimental results.

Further investigations are planned for refined characterization of high density particle suspensions with X-ray measurements.

Symbols & Units

Symbol	Quantity	SI [a]
Ar	Archimede number	none
Re_c	Reynolds number for U_c	none
Re_k	Reynolds number for U_k	none
A	riser section	m^2
d_p	mean particle diameter	m
d_s	mean Sauter particle diameter	m
D	riser internal diameter $D = (4\ A/\pi)^{1/2}$	m
g	gravitational acceleration	m/s^2
G_s	solid mass flow	kg/s
	or solid mass flow rate	$kg.m^{-2}\ s^{-1}$
H	riser internal height	m
U_a	superficial gas velocity, fluidization velocity	m/s
$U_{c,k,t}$	characteristic velocity	m/s
U_{ch}	choking velocity	m/s
ε	porosity	none
Φ	sphericity	none
μ	gas viscosity	Pa.s
$\rho_{g,s}$	gas and solid density	kg/m^3

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