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Hold-up characteristics of a novel gas–solid multistage fluidized bed reactor for control of hazardous gaseous effluents

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ABSTRACT

This paper illustrates the mechanism of gas–solid contacting in a multistage counter-current fluidized bed reactor attached with downcomers. Experimental hold-up study has been carried out extensively using the air and lime particles as gas and solid phases, respectively. The reactor was operated in continuous regime for two-phase system over a wide range of operating conditions including height of the downcomer to determine the solids hold-up at each stage of the reactor. The data was generated under steady and stable operation of the column. Based on the experimental data, empirical correlation is proposed for predicting the solids hold-up in the multistage fluidization. The results in this study assume importance from the standpoint of design and steady operation of multistage fluidized bed reactor with downcomer as an air pollution control system for control of gaseous emissions in the industries.

1. Introduction

The single-stage gas–solid fluidized bed reactors used for different operations such as drying, adsorption, calcinations, combustion of coal, etc. in the industries has some inherent drawbacks like high degree of internal mixing and the low contact efficiency between gas and solid phases. The staging of fluidized bed reactors by providing the horizontal baffles overcome these drawbacks associated with single-stage fluidization [1]. Sometimes provision of internal baffles improves the continuous co-current or counter-current flow of the gas–solid phases approaching plug-flow condition [2], in addition it establishes the temperature and concentration gradients along the length of the bed. Such arrangement in the fluidized bed controls the formation and growth of the bubbles. These multistage fluidized beds can be classified based on the transfer of the solids from one stage to next stage below namely (a) solids passing through perforated plates (b) solids passing through downcomers which are simply empty tubes allowing the transfer of solids from upper fluidized bed to lower one. In the former the diameter of perforated plate holes are large enough to allow simultaneous flow of solids and gas through them.

Enough literature is available with studies carried out on the multistaged fluidization without downcomer for gas–solid operations [3–6]. On the other hand, the literature of the stage-wise operation of the gas–solid reactor with downcomer is very scarce. These types of multistage fluidized bed reactor with downcomer have recently gained importance for recovery of gaseous pollutants as an air pollution control system in the industries. The limitation of a single-stage continuous fluidizer is wide distribution of residence time of solids and low efficiency of operation both with respect to gas and solid phase besides the chance of slugging in deep beds. The alternative way to obtain a narrow residence time distribution along with high efficiency of operation with respect to gas and solid phase is to use multistage fluidized bed systems. Although there are some commercial units utilizing the principles of stage-wise operation of a fluidized bed reactor as an air pollution control unit, the available data are insufficiently instructive. In these studies the objective is to examine the types of flow patterns prevalent in the gas–solid fluidized beds and determine the solids hold-up in the system over a wide range of operating variables under steady operation of the multistaged column. Finally, the variables affecting the solids hold-up has been grouped into various dimensionless parameters based on which a correlation is developed for predicting solids hold-up as functions of the operating variables.

2. Experimental

2.1. Flow properties of solids

The fluidization of solids depends on the properties of solids. The main properties affecting fluidization are solid particle size,

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Nomenclature

A	area of cross-section of the column or bed (m^2)
d_c	diameter of the cone (m)
d_p	diameter of the particle (m)
d_w	diameter of downcomer (m)
D	diameter of column (m)
G_f	mass velocity of air ($kg/m^2\,s$)
G_s	mass velocity of solids ($kg/m^2\,s$)
h_w	weir height (m)
s	acceleration due to gravity (m/s^2)
u_g	superficial velocity of air (m/s)
u_{mf}	superficial velocity of air at the onset of fluidization (m/s)
u_s	superficial solids velocity (m/s)
W	solids hold-up in each bed (kg)

Greek symbols

ε	bed porosity
ρ_g	density of air (kg/m^3)
ρ_s	bulk density of solid materials (kg/m^3)

Subscripts

c	calculated/predicted
e	experimental
f,g	fluid (air)
mf	minimum fluidization
o	superficial
p	distributor/plate
s	solids
w	wall

density, porosity. The solid phase such as calcium oxide (lime) has been selectively chosen for fluidization to determine the hold-up of solid phase. The adsorption of sulfur dioxide gas onto lime have been investigated to control hazardous gaseous pollutants as second part of the experiment. The calcium oxide (CaO), has been used for control of gaseous pollutants in the single-stage fluidization by many researchers [7,8]. The methods followed to determines the properties are: (i) particle size (d_p), it was determined by standard sieve analysis in B.S. sieves to the investigated size (0.353–0.50 mm) and kept in desiccators to avoid being deliquesced. The only sharp cut fractions were used in the experimentation. The average of the diameters of the sieves through which the solids passed and the sieve on which it retained was taken as diameter of the particles. (ii) Density (ρ_s): it was determined by kerosene to ensure that the wettabillity of the lime by the liquid does not affect the results. The density of the solids was found to be 2040 kg/m^3. (iii) Porosity (ε_s): the porosity or void fraction of a solid material of a definite size is the ratio of the void volume to the volume of the bed gives the porosity of the bed. To ensure accuracy in fluidization experiments, the cylinder diameter was taken same as the diameter of the column. The porosity was recorded to be 0.48. (iv) Minimum fluidization velocity (u_{mf}): the minimum fluidization velocity was determined experimentally in a single-stage fluidizer at room temperature and atmospheric pressure. Fig. 1 shows the effect of superficial gas velocity on the bed pressure drop. With increase in gas velocity, the pressure drop due to solids increased and reached at a point where pressure drop due to solids remained constant as 125 N/m^2. Further there was no change in pressure drop with increase in gas velocity up to 0.283 m/s. Under this fluidized bed conditions, pressure drop is approximately proportional to the apparent weight of the lime particles. The minimum fluidization point was defined as the intersection of the extrapolated fixed bed characteristic with the line of constant bed pressure of the fluidized bed [9], which revealed the minimum fluidization velocity of the lime particle (average d_p = 426 μm) as 0.112 m/s.

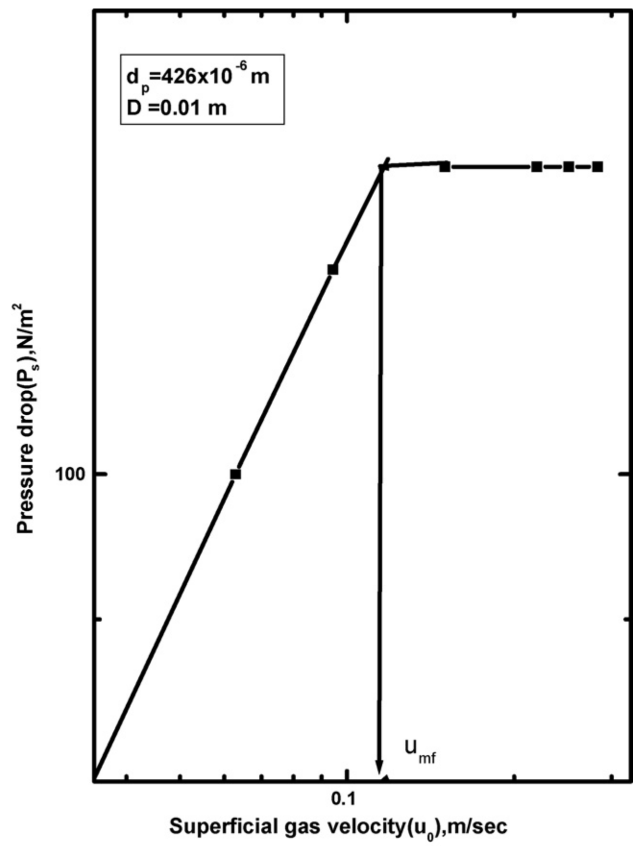

Fig. 1. Effect of superficial gas velocity (U_0) on solids pressure drop (P_s) in single-stage fluidizer.

2.2. Experimental set-up and procedure

Fig. 2 is the schematic of the multistage fluidized bed reactor developed and used in this study. The configuration of this staged gas–solid fluidized bed reactor is similar to that of the sieve trays distillation column. The reactor consists of a three stage fluidization column (FB_1, FB_2 and FB_3) having provision of solid feeding from the top and air supplying system from the bottom along with other auxiliary equipments used for experimentation. Each stage of the column was constructed of perspex cylinder of 0.10 m internal diameter and 0.305 m long. The stainless steel plates of 0.002 m thick each (G_1, G_2 and G_3) were used as internal baffles between two stages and each plate was drilled with perforations of 0.002 m diameter on a triangular pitch having 8.56% total grid openings [10–11]. The perforations were made accurately avoiding possible burrs and protrusions during drilling process. The grid plates were covered with fine wire mesh (100 mesh size) to prevent the solids from the falling through the openings. Each section was provided with a downcomer of perspex cylinder of 0.025 m internal diameter (D_1, D_2, D_3 and D_4) and the downcomers were fitted to the gas distributors by special threading arrangement having the provision for adjusting the weir height as desired. The downcomers were further fitted with a cone at the exit end in order to reduce the up flow of the gas through the downcomer and consequently,

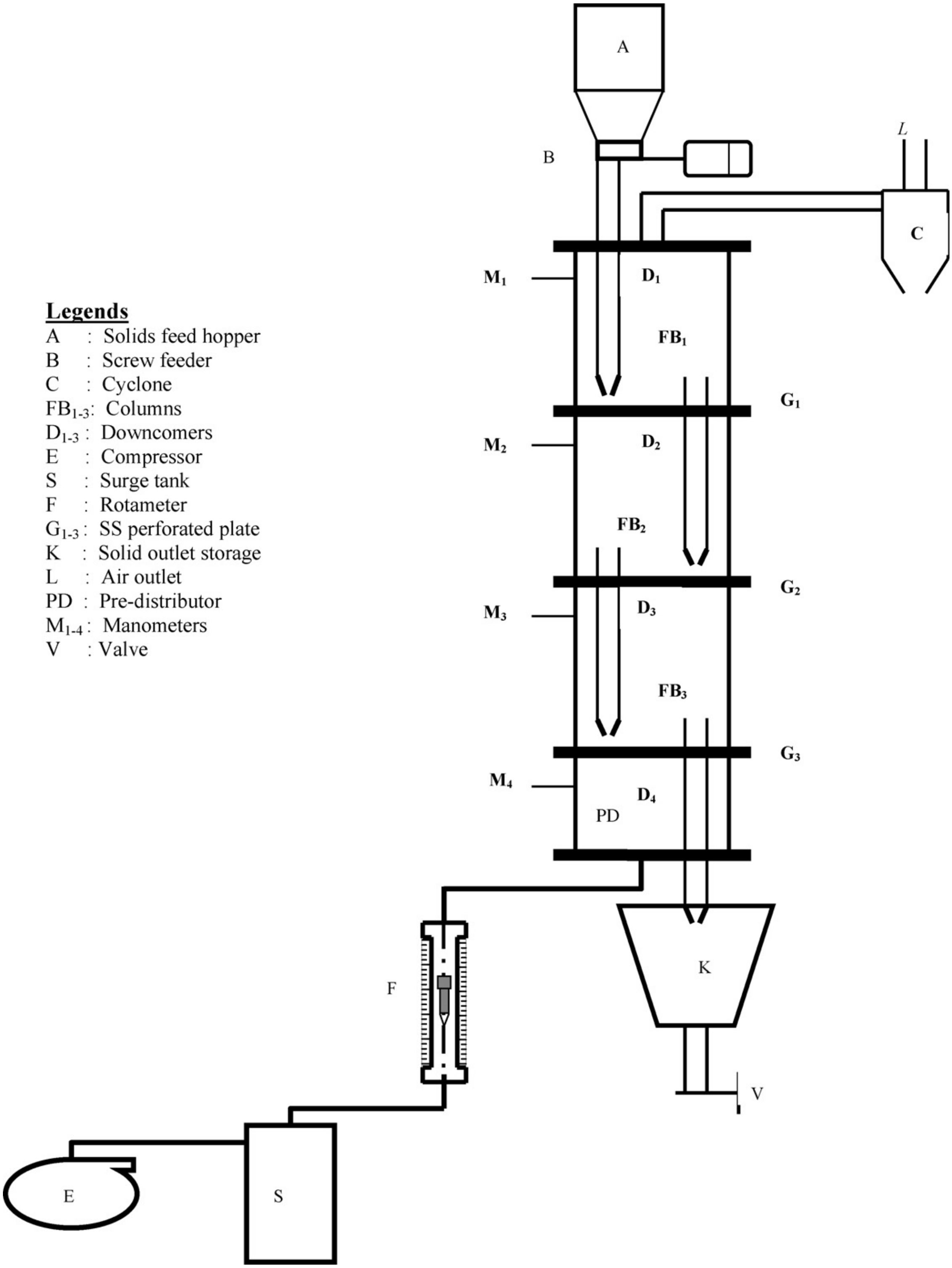

Fig. 2. Schematic diagram of the experimental set-up of a three-stage counter-current fluidized bed reactor.

widening the stable operating range. The bottom-opening diameter of the cone was 0.012 m for flow of solids to the next stage [12]. Pressure tapings were provided just below the grid plate and the near the air out let and four manometers were provided to measure the pressure drop at every stage as well as the total pressure drop. A compressor (E) of 5.0 kW was used to supply the air as fluidizing gas and its flow rate was measured using a calibrated rotameter (F). A gas distributor of 0.150 m long was provided at the bottom of column for uniform distribution of gas to the column. A conical hopper was attached at the bottom of column for storage of solids coming out from the bottom stage through the downcomer.

The gas leaving the column from the top stage was passed through a 0.15 m diameter standard cyclone (C) and then into the exhaust system. A cloth bag was attached at the bottom of the cyclone to collect the fines, if any, carried over.

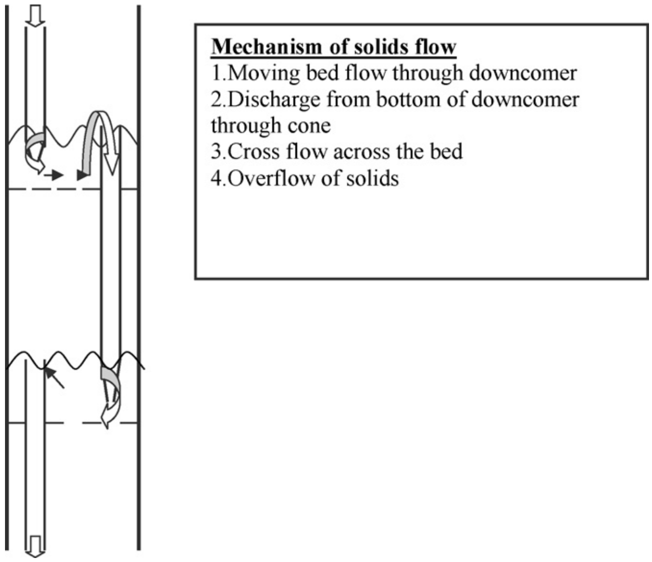

Fig. 3. Schematic of solids flow from upper stage to next stage.

A perspex hopper of 0.150 m internal diameter and 0.40 m long called as feeding funnel was used to hold the lime and attached to the screw feeder .The solids from the screw feeder was fed through a Perspex tube of 0.012 m internal diameter to the first stage downcomer of the reactor. The screw feeder was fitted to a motor of 0.25 kW and the speed of the motor was controlled by a variable rheostat. A compressor was used to provide the air through a pre-distributor underneath the last stage. The gas flow rate was adjusted by a rotameter. Necessary precautions were made to ensure that no air from outside intruded into the column during operation.

The solids were fed into the column at the rate from 0.035 to 0.142 kg/m^2 s with gas flow rate varying from 0.32 to 0.59 kg/m^2 s. The weir height of the downcomer were kept at 30, 50 and 70 mm and the gap between the downcomer bottom and the grid plate were kept 15, 25 and 35 mm, respectively. The pressure drops across each stage and across the entire column were recorded. In the experiment, for a given apparatus conditions, the effect of gas flow rate and solid flow rate on solids hold-up have been observed.

Fig. 3 is the pictorial representation of flow conditions progressively existing on the stage, beginning with a fixed bed of lime particle with gas flowing upward and then gradually changing over to the continuous operation in the column with gas and solid flowing counter currently at constant flow rates. The solids were first fed into the feeding funnel of the screw feeder and the screw feeder fed the solids at the desirable rate through the feeding downcomer to the top stage. At very low gas flow rate (less than minimum fluidization velocity), solids filled the downcomer and piled over the distributor plate of the top stage just below the downcomer. Airflow rate was then increased from bottom of the column. As the gas rate was increased in small increments, the solids dispersed and started distributing in the top stage due to the difference in the solids gradient developed across the stage from the left (location of the upper downspout) to the right of the stage, resulting in the cross-flow of the fluidized solids on the stage. As the flow rate of gas was gradually increased, the fluidized particles flowed over the weir to the next lower stage, as shown in Fig. 3. The solids were then transferred from stage to stage fluidizing in each stage and finally transferred to the solids outlet storage. The operation of the column was usually smooth and steady when all the stages had sufficient material.

When the system was operated with solids, it was observed that all the stages of the reactor were identical in their operation as well as performance. The system was allowed to reach equilibrium so that the inlet and outlet solids flow rate were equal. The gas and solid flow rate was then cut-off simultaneously and the solids of all stages were weighed and no discernible difference in the solids hold-up across the stage was noticed from stage to stage (with variation less than 2%). In view of identical performance and equal solids hold-up, the variations of solids hold-up across a single-stage under various flow conditions have been presented in Fig. 4.

3. Results and discussion

3.1. Pressure drop in the multistaged fluidized bed reactor (with solid flow)

While operating the system with solids, it was observed that all the stages of the reactor were identical in their operation as well as performance and the pressure drops across each stage and across the entire column were recorded. No discernible difference in the pressure drop across the stage was noticed from stage to stage. In view of identical performance, the pressure drop due to solids across each stage has been obtained from the difference between the pressure drop with and without solids. The variations of pressure drop across a single-stage under various flow conditions have been presented in Fig. 4.

Fig. 4 describes the pressure drop due to solids measured across each stage varying the gas, solid flow rates and weir heights. It is seen from the figures that the pressure drop due to solids, ΔP_s, decreases with increase in the gas flow rate and increases with increase in the solids flow rate. It is also seen from the figures that the pressure drop due to solids, ΔP_s, decreases with increase in the gas flow rate and increases with increase in the solids flow rate. Similar observations were reported by [5,13]. The minimum pressure drops occurred in the column at high gas flow rate (0.58 kg/m^2 s) corresponding to minimum solid flow rate (0.071 kg/m^2 s) are 67.1, 112.3 and 164.3 N/m^2 at 0.03, 0.05 and 0.07 m weir height, respectively. The maximum pressure drops occurred in the column at low gas flow rate (0.32 kg/m^2 s) corresponding to maximum solid flow rate (0.141 kg/m^2 s) are 98.4, 139.6 and 185.1 N/m^2 at 0.03, 0.05 and 0.07 m weir height, respectively. It appears that the reason may be that an increase in gas rate increases the porosity of the bed in the system, resulting in decrease in the solids concentration and hence the pressure drop across the stage, as the height of the fluidized bed in the system corresponds to downcomer weir height.

The steady state hold-up for all the stages was studied as a function of the solids flow rate with the gas flow rate, weir height as parameters. Fig. 5 describes the effect of mass velocity of gas on solids hold-up measured across each stage varying the solid flow rates from 16.67 to 66.8 g/min and weir heights from 0.03 to 0.05 m in the three-stage reactor operated under steady-state conditions, with the lime particles and gas flowing counter currently. It is seen from the figures that the solids hold-up, W, decreases with increase in the gas flow rate and increases with increase in weir height. It is seen from the figures that the solids hold-up, W, increases with increase in the solids flow rate [5,13]. The minimum solids hold-up obtained in the column at high gas flow rate (0.58 kg/m^2 s) corresponding to minimum solid flow rate (0.035 kg/m^2 s) are 0.057 and 0.096 kg at 0.03, 0.05 m weir height, respectively and minimum solids hold-up obtained in the column at high gas flow rate (0.58 kg/m^2 s) corresponding to minimum solid flow rate (0.071 kg/m^2 s) are 0.139 kg at 0.07 m weir

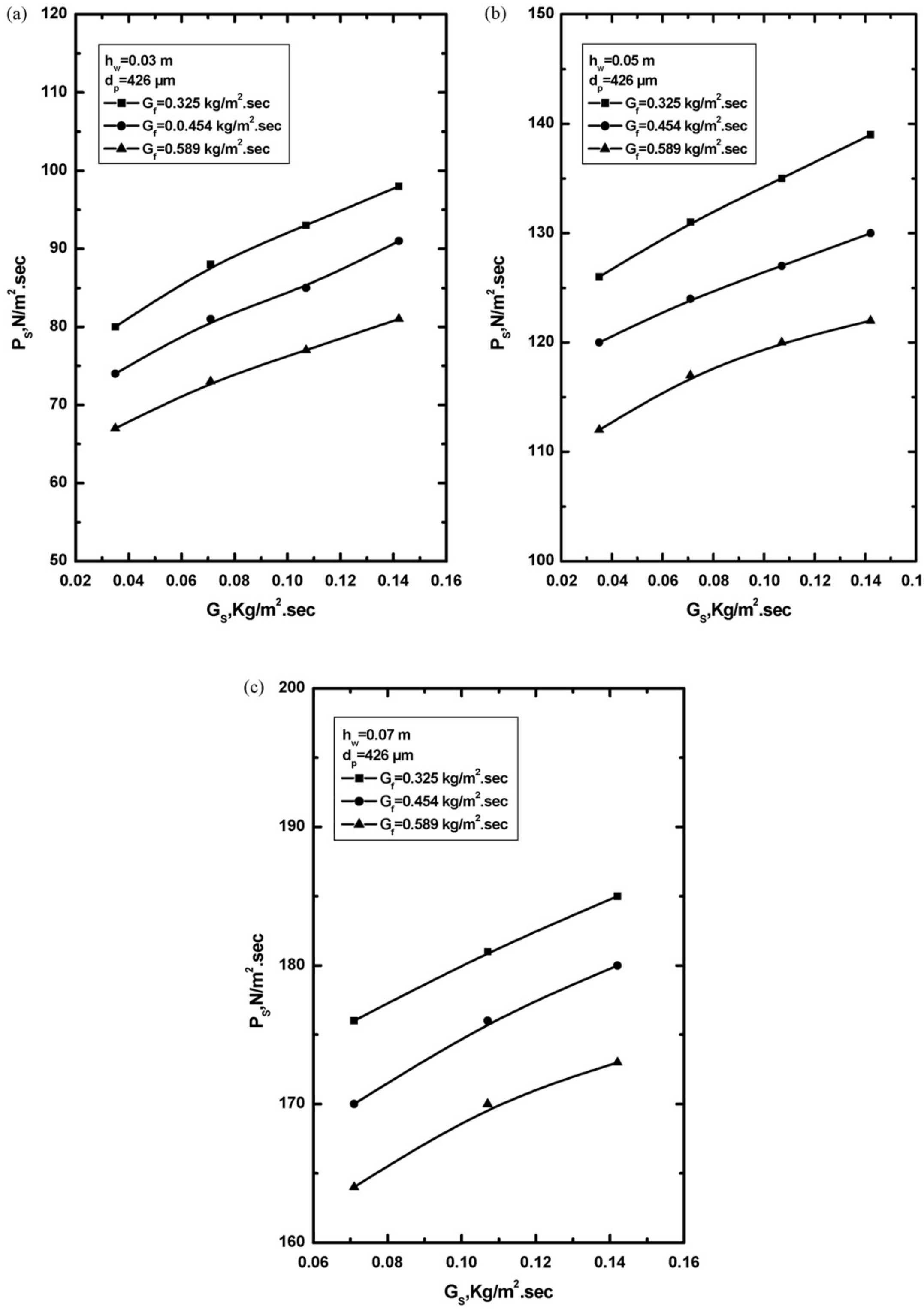

Fig. 4. Effect of mass velocity of solids (G_s) on pressure drop (P_s) (a) 30 mm weir height, (b) 50 mm weir height, and (c) 70 mm weir height.

height. The maximum solids hold-up obtained in the column at low gas flow rate (0.32 kg/m^2 s) corresponding to maximum solid flow rate (0.141 kg/m^2 s) are 0.083, 0.117 and 0.156 kg at 0.03, 0.05 and 0.07 m weir height, respectively. The reason may be that an increase in gas rate increases the porosity of the bed in the system, resulting in decrease in the solids concentration and hence the solids hold-up at the stage. Increasing the weir height increases the solids concentration resulting in increasing the solids hold-up.

4. Correlation of the experimental data

An attempt has been made to correlate the solids hold-up with variables of the system. The operating variables

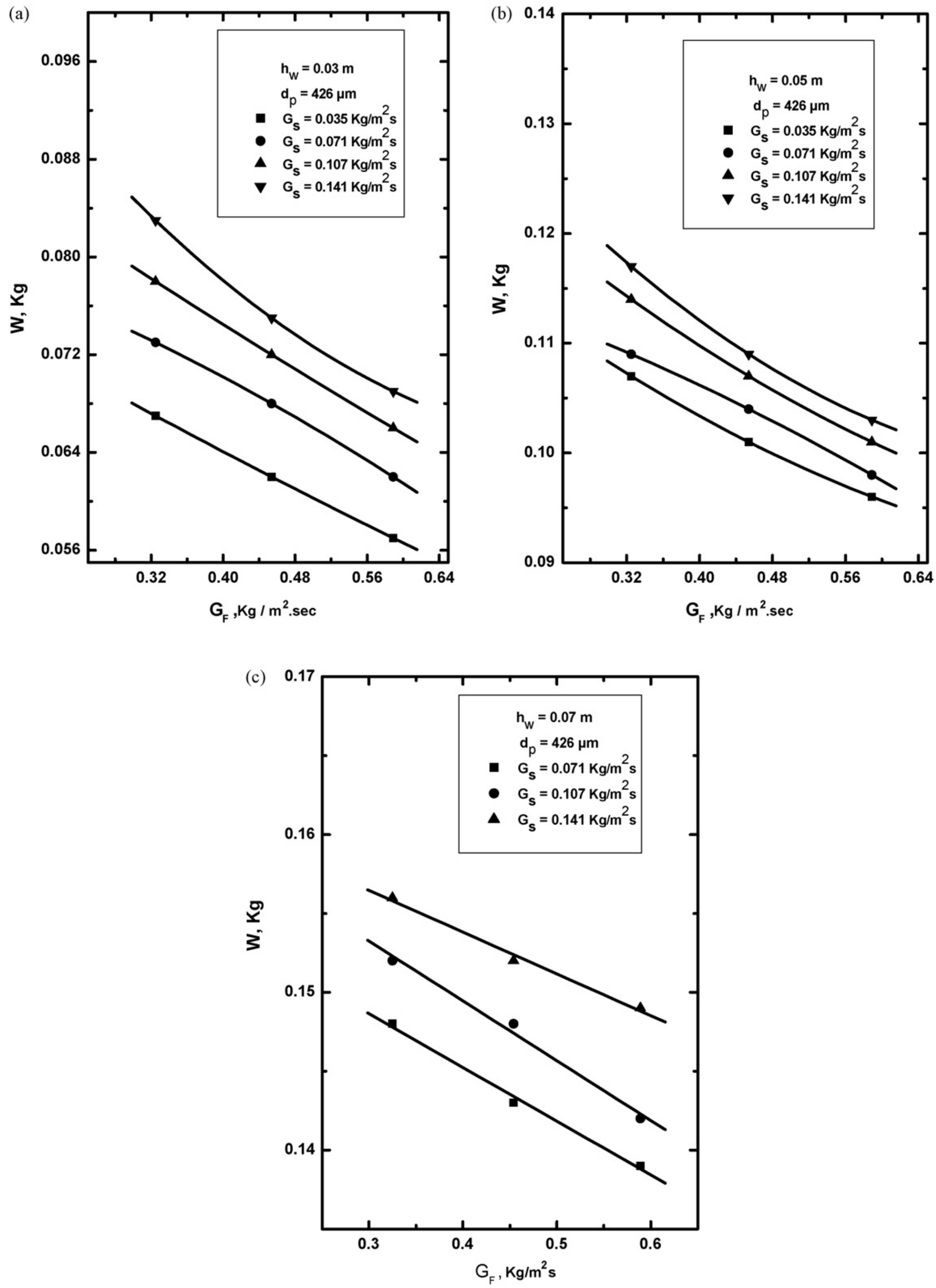

Fig. 5. Effect of mass velocity of gas (G_f) on solids hold-up (W) (a) 30 mm weir height, (b) 50 mm weir height, and (c) 70 mm weir height.

grouped into dimensionless numbers by employing Buckinghum's π theorem and multiple linear regression analysis has been made in order to establish the functional relationship between the solids hold-up and the dimensionless groups.

Conceivable variables on which the friction factor in the present system may depend are:

(a) Flow properties—gas velocity (u_g) and solids velocity (u_s);
(b) geometrical properties—cross-sectional area of reactor/bed (A), diameter of reactor (D), diameter of the downcomer (d_w), weir height (h_w);
(c) physical properties—namely the density of gas (ρ_g), density of solid (ρ_s), diameter of particles(d_p) and gravitational constant (g).

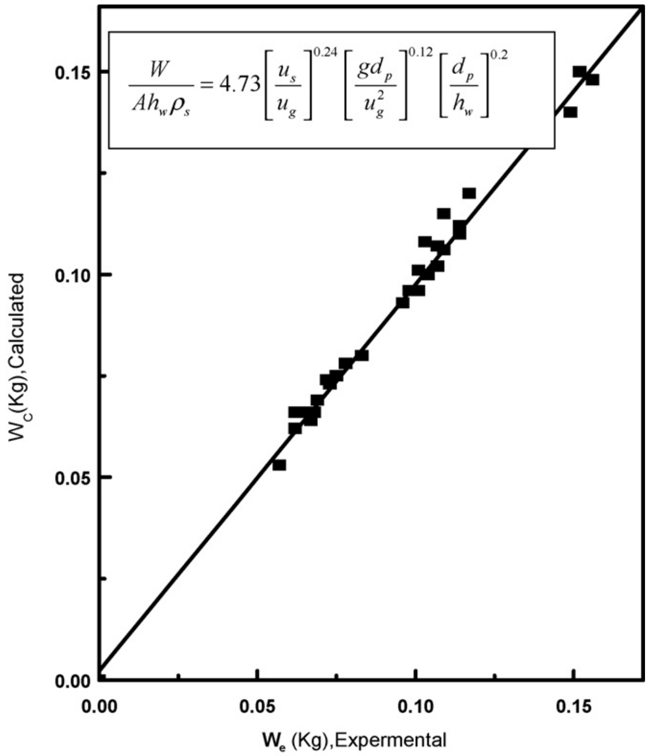

Fig. 6. Correlation between predicted hold-up (W_c) and actual hold-up (W_e) in a multistage fluidized bed.

The dimensionless analysis is $$\frac{W_c}{\rho_s} = k\left[\frac{u_s}{u_g}\right]^a\left[\frac{gd_p}{g_g^2}\right]^b\left[\frac{d_p}{h_w}\right]^c \qquad (1)$$

In order to establish the functional relationship between solids hold-up and the various dimensional groups in Eq. (1), multiple linear regression analysis has been performed to evaluate the constants and coefficients of the equation. It can be seen that the following equation, which yield the minimum percentage error, present the best possible correlation:

$$\frac{W_c}{Ah_w\rho_s} = 4.73\left[\frac{u_s}{u_g}\right]^{0.24}\left[\frac{gd_p}{u_g^2}\right]^{0.12}\left[\frac{d_p}{h_w}\right]^{0.2} \qquad (2)$$

Eq. (2) was obtained by multiple least square regression method. The coefficient of correlation was of the order of 0.9965 and standard deviation of the experimental data from regression analysis was found to be too small. To check the consistency of the experimental data, the values of solids hold-up calculated (W_c) using Eq. (2) have been plotted against the experimental values (W_e) in Fig. 6, where the solid line represents the regression equation and points are experimental values. The percentage deviation between the experimental data and those predicted by Eq. (2) has been found to be small. Thus, the empirical correlation satisfies the experimental data of the present system satisfactorily.

5. Conclusions

The solids hold-up of a gas–solid fluidized bed reactor was studied in this work. The unique feature of this reactor was the multistage system. The system was designed to operate continuously with entry of solids and exit of solids from each stage through the downcomer. Detailed studies on the solids hold-up have been carried out for design and characterization of the reactor. During steady operation, fluidized particles move across the stage to the next stage through the downcomer, as gas flows upward through the distributors followed by voidage of the solid particles. The maximum solids hold-up occurred at each stage of the column at low gas flow rate corresponding to maximum solid flow rate. This justified the present multistage system, where the solids hold-up achieved is three times higher than the single-stage system for the same superficial gas and solid velocities. The empirical correlation has been developed to estimate the solids hold-up on the stage during the continuous operation of the column. Experimental results are in excellent agreement with the correlations. These data presented in this study assume significance from the perspective of design and stable operation of staged fluidized bed reactors.

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