Chemical Engineering and Processing 49 (2010) 1095–1100

Contents lists available at ScienceDirect

Chemical Engineering and Processing: Process Intensification

journal homepage: www.elsevier.com/locate/cep

Minimum superficial fluid velocity in a gas–solid swirled fluidized bed

S. Harish Kumar [a,*], D.V.R. Murthy [b]

[a] Department of Chemical Engineering, Manipal Institute of Technology, Manipal University, Manipal 576104, India
[b] Department of Chemical Engineering, National Institute of Technology Karnataka, Surathkal, Srinivasnagar 575025, India

ARTICLE INFO

Article history:
Received 12 November 2009
Received in revised form 15 March 2010
Accepted 12 August 2010
Available online 20 August 2010

Keywords:
Minimum swirl velocity
Phase diagrams
Swirl fluidized bed

ABSTRACT

A swirl flow is achieved in a bed of solids by passing air through multiple fluid inlets, which are tangentially located at the base of a flat-based circular column. The minimum superficial velocities needed to achieve swirling of the bed are measured experimentally under varied conditions. An empirical correlation for the minimum swirl velocity has been proposed. The results indicate that a stable swirling regime operation of the bed is possible. There exists an upper limit of static bed depth beyond which stable swirling of entire bed is not possible. The minimum swirl velocities are found to be 1.2–1.3 times the minimum fluidization velocities predicted for conventional fluidized beds.

1. Introduction

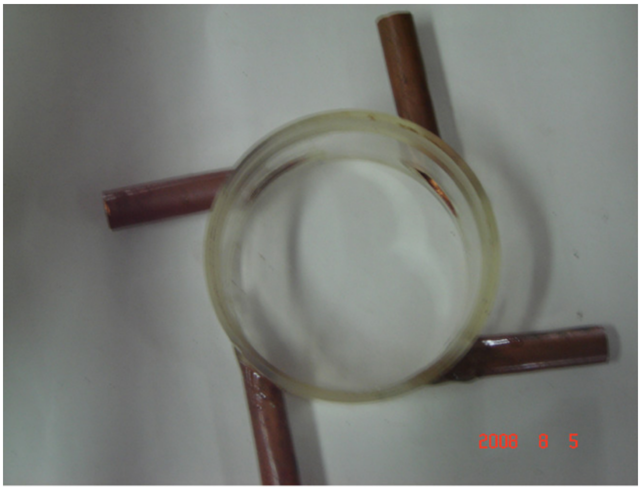
Fluidized beds are widely used in chemical process industries. Even though conventional fluidized bed systems are well developed, it is regarded that gas–solid fluidized beds with coarse particles have the problems of bubble formation, slugging and channeling. Several modifications in fluidized beds have been considered by various researchers to overcome those problems, achieve better contact between gas and solid phases such that the transfer rates are improved. Generating a swirl flow in a bed of solids is one such modification proposed and studied by a few investigators. Lee et al. [9] investigated coal combustion characteristics in a two-stage swirl fluidized bed combustor with lower stage (first stage) as a conventional fluidized bed and the upper stage (second stage) having a swirl generator; they reported that the presence of swirling in the second stage reduced the elutriation rates of fines and also improved the combustion efficiency by 20%. Chyang and Lin [2] used a multi horizontal nozzle distributor to achieve swirling-fluidizing pattern and reported that it produced a remarkable improvement in fluidization quality and elutriation reduction. A swirl fluidized bed featuring an annular bed and inclined injection of gas through distributor blades was studied by Sreenivasan and Raghavan [1]; they reported the hydrodynamic characteristics of swirl regime of operation, including their experimental results as well as an analytical model. Dolidovich [4] developed a swirled spouted bed to increase the efficiency of heat treatment of polyfractional and fine particles and to reduce their entrainment.

Wilde and Broqueville [13] reported their experimental investigations of rotating fluidized bed in a static geometry. The fluidization gas was tangentially injected via multiple gas inlet slots in a cylindrical outer wall; this induced a rotating motion. Radial fluidization of the solids was created by introducing a radially inwards motion of gas towards a centrally positioned chimney, through which it exited finally. They reported that stable rotating fluidization was obtained when solids loading was sufficiently high. In a multi-stage vortex combustor for burning rice husk, Eiamsa-ard et al. [5] used four tangential air entry nozzles in a second stage. They reported that such an arrangement improved the mixing between gas and solids leading to better combustion. Kaewklum et al. [8] studied hydrodynamic regimes and characteristics of two types of conical swirling fluidized beds: one having tangential gas entry through four nozzles and second having an axial entry through an annular-spiral gas distributor. They reported four regimes in bed operation and empirical equations for predicting minimum fluidization velocity and bed pressure drop at minimum fluidization. A conical swirling fluidized bed combustor having an annular-spiral air distributor was used by Kaewklum and Kuprianov [7] to investigate hydrodynamics, combustion and emission characteristics. They reported four regimes of bed operation and very high combustion efficiency. Zhang [12] considered tangential gas entries in a numerical simulation study of a swirl fluidized bed reactor; however no experimental data has been reported by him.

The present study deals with understanding the hydrodynamic behavior of swirled fluidized bed wherein multiple fluid inlets are tangentially located at the base of a flat bottom circular column

* Corresponding author. Fax: +91 820 2571071.
E-mail addresses: harishmanipal@yahoo.co.in (S. Harish Kumar), dvrmvzm@gmail.com (D.V.R. Murthy).

0255-2701/$ – see front matter © 2010 Elsevier B.V. All rights reserved.
doi:10.1016/j.cep.2010.08.013



D_c =0.1m, D_i=0.012m, N=4.

D_c =0.08m, D_i=0.01m, N=3.

Fig. 1. Typical column base assemblies.

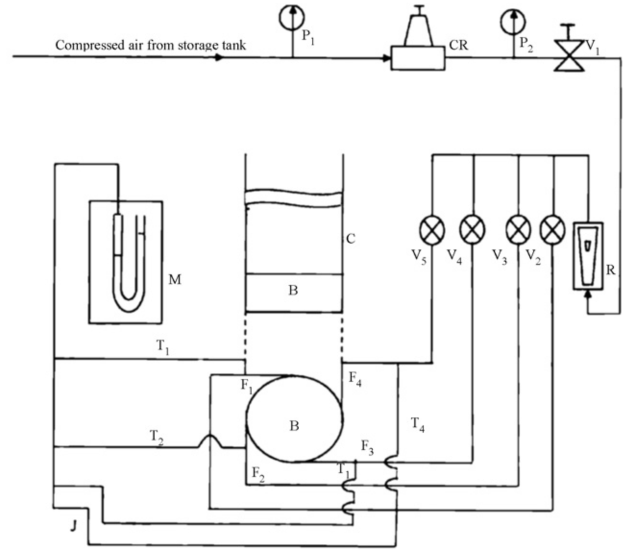
to achieve swirl flow inside the bed. This type of bed operation is distinctly different from the works reported earlier in the literature.

2. Experimental set-up and techniques

The experiments are carried out in transparent acrylic columns having the following:

- Column internal diameters: 0.08, 0.1 and 0.14 m.
- Fluid inlet internal diameters: 0.008, 0.01 and 0.012 m.
- Number of fluid inlets: 2, 3 and 4.

The flat base of each column is also made of transparent acrylic pipe to which tangential fluid inlets are assembled as shown in Fig. 1. The properties of bed materials used in the study are given in Table 1. Compressed air at 29.5 °C and 101.3 kPa (g) is supplied to the column to fluidize the solids. A schematic line diagram of the experimental set-up is shown in Fig. 2.



P_1, P_2 = Pressure gauges
CR = Constant pressure regulator with moisture trap
V_1 = Gate valve
R = Rotameter
V_2, V_3, V_4, V_5 = Needle valves
B = Base column with tangential fluid inlets.
C = Perspex column
J = Common junction for pressure taps
M= Manometer
F_1, F_2, F_3, F_4 = Tangential fluid inlets
T_1, T_2, T_3, T_4 = Pressure tapping

Fig. 2. Schematic line diagram of experimental set-up.

The column is filled with solids up to a required height and the bed is brought into fully swirled condition. Then fluid flow is gradually brought to zero. This operation is carried out three times. The flow is then gradually increased such that fountains of solids are formed near all the fluid inlets. By adjusting the valves in each of the lines connected to fluid inlets the fountain heights are brought to equal; as a result it has been assumed that the fluid rates into the inlets are equal. The total flow rate is gradually brought to zero. The static bed height is measured. The total flow rate is then gradually increased to conduct the experiment. Observations are noted for bed pressure drop at various fluid flow rates till the bed is transformed into an expanded swirled state. Later, the same observations are noted by gradually decreasing the flow to zero.

3. Results and discussions

3.1. Bed pressure drop–superficial velocity relationship

For a given set of conditions this relationship is obtained experimentally as shown in Fig. 3. For a given solid material, fluidized in

Table 1
Properties of bed materials.

Sl. no.	Solid material	Shape	Particle size, d_p (m)	Bulk density, ρ_b (kg/m^3)	Solid density, ρ_s (kg/m^3)
1.	Polmer1	Disc	2.25×10^{-3}	769	1110
2.	Polymer2	Spherical	3.75×10^{-3}	556	901
3.	Raagi seeds	Spherical	1.31×10^{-3}	869	1379
4.	Jowar seed	Spherical	2.81×10^{-3}	889	1351
5.	Wheat	Pear	3.17×10^{-3}	807	1352

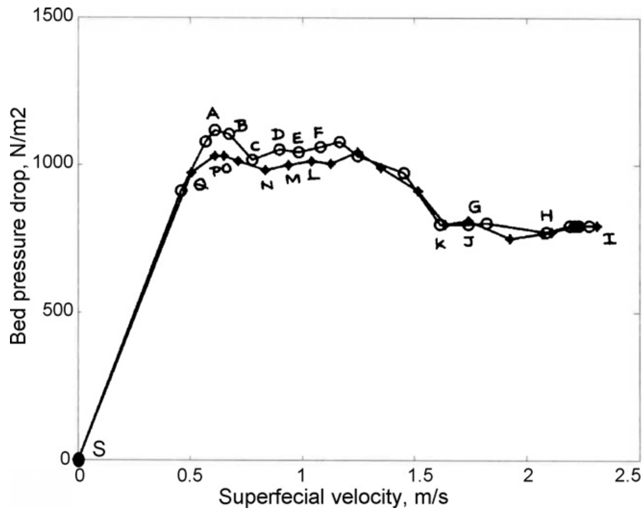


Fig. 3. Typical bed pressure drop–superficial fluid velocity relationships. (material: Raagi seeds, $H = 0.18$ m, $D_c = 0.1$ m, $D_i = 0.01$ m, $N = 3$). The various symbols represent: o increasing flow rate, * decreasing flow rate.

a given column having a given number of fluid inlets of a particular size, the bed pressure drop–superficial velocity relationships are obtained experimentally. These are obtained for increasing as well as decreasing fluid flow rates.

When a jet of fluid enters the bed tangentially through the fluid inlet, it has two components of velocities; the vertical component causes fluidization and the tangential component would be responsible for swirl motion of the bed material. At low flow rates of fluid, the solids remain in a packed state with the pressure drop raising as the flow rate is increased along SA. In this region it has been observed that small cavities are formed due to tangential component of velocity near the fluid inlets and close to the column wall. The vertical component of fluid velocity dies out at a certain height above the fluid inlet but within the static bed, as a result the bed remains in a packed state. As the flow rate is gradually increased, the tangential component of the fluid jet tries to push the particles away from the fluid inlet and makes the particles in the vicinity of the cavity densely packed. This results in increased pressure drop along SA. At point A, the vertical component of fluid velocity is good enough to cause the formation of fountain near the wall. It should be noted here that fountains of solids are formed in the vicinity of the fluid inlets and very close to column wall. However the solids in the middle portion of the bed still remain in a packed state. When the fountains are formed the bed pressure drop starts decreasing. Ideally it is expected that all the fountains near the fluid inlets should be forming at point A, since the flow rates through each of the fluid inlets are supposed to be equal. However, it has been observed that these fountains are formed one after the other at A, Band C as the fluid velocity is increased. This could be due to little imbalances in individual flow rates through the fluid inlets.

As the flow rate is gradually increased, the fountain heights continue to increase in the region CDEF. When the flow rate is gradually increased beyond F, it has been found that the solid fountains become inclined and exhibit oscillatory behavior and then gradually transforming the bed into partially swirled state. The pressure drop has been found to be decreasing during this transition. This may be due to the reason that the tangential component of the fluid velocity could be increasing whereas the vertical component of the fluid velocity decreasing. Beyond point G, the bed is transformed into fully swirled state. In this state the vertical component of fluid velocity may not exist and the bed motion is completely due to the tangential component of fluid velocity alone. In the region G to H, the bed is in complete swirled state, with bed pressure drop remaining almost constant. When the fluid flow rate is gradually increased beyond H, the bed is transformed into an expanded swirled state till the point I. The pressure drop in the expanded swirled state has been found to be slightly higher than the earlier swirled state.

From the expanded swirled state, the flow rate is gradually decreased to bring back the bed into stable swirling in the region HK. At point H the minimum superficial velocity for expanded swirling (V_4) has been identified which is the minimum fluid velocity at which the bed changes from expanded swirling to stable swirling. Upon further decreasing the flow rate gradually, the bed is transformed into a partially swirled state from K to L, leading to gradual increase in bed pressure drop. So it is considered that the point K may be taken as the point of minimum superficial velocity of fluid for swirling (V_3). When the flow rate is decreased gradually from L to O, the formation of solid fountains has been observed. Small reduction in flow rate at O causes one of the fountains to collapse leading to decrease in bed pressure drop. Here the minimum superficial velocity for fountains region (V_2) is identified. Upon further decrease in the flow rate gradually, in the region OPQ, the fountains are collapsed one by one and the bed is transformed to a complete packed state beyond point Q. The minimum superficial velocity for fountain formation (V_1) is identified corresponding to point Q. In region QS the bed remains in packed state with pressure drop gradually decreasing.

It may be noted here that the paths taken for increasing and decreasing flow rates are different. This may be due to the bed hysterisis, which is typically exhibited by the beds of particulate solids. Hence the hydrodynamic parameters of interest, namely minimum superficial velocity for swirling (minimum swirl velocity) and the bed pressure drop during stable swirling (swirl bed pressure drop) are found out from the measurements of decreasing flow rate. These are found to be reproducible by conducting a number of trials in each experimental run. It is to be noted that there are no problems of bubble formation, segregation and channeling observed in the stable operation of the swirled beds.

3.2. Phase diagrams

As the fluid flow rate is increased the bed transforms from a packed state to a swirled state; these transitions are mapped in the form of phase diagrams using the bed pressure drop–superficial fluid velocity data obtained experimentally. The data needed to generate these diagrams for various conditions are given elsewhere [6] and typical plots are shown in Figs. 4 and 5. Regions 1, 2 and 3 do not have much practical relevance. Steady swirling with complete mixing of solids occurs in region 4 with bed pressure drop remaining fairly constant. In region 5 also the bed would be in a completely swirled state, but with more bed expansion. So, one can think of operating the bed either in region 4 or 5 and the choice of operation depends on a given application. For instance in an operation like drying of heat sensitive solids, it may be useful to operate the bed in region 5, where the gas residence time is lower, but with higher gas consumption. In region 4 the gas consumption would be lesser but the contact time between the phases would be more.

3.3. Maximum swirlable bed depth

For a given column configuration and solids, it has been observed that stable and satisfactory swirling is not possible beyond a particular static bed depth. When experiments have been conducted beyond this bed depth the following has been observed.

The solids in the lower portion of the bed nearer to fluid inlets are found to be in swirling motion completely, whereas in the upper portion of the bed swirl action has given way to bubbling flow, slugging or aggregative fluidization; stable swirling of entire bed has not been possible at any gas flow rate. With an increase in gas

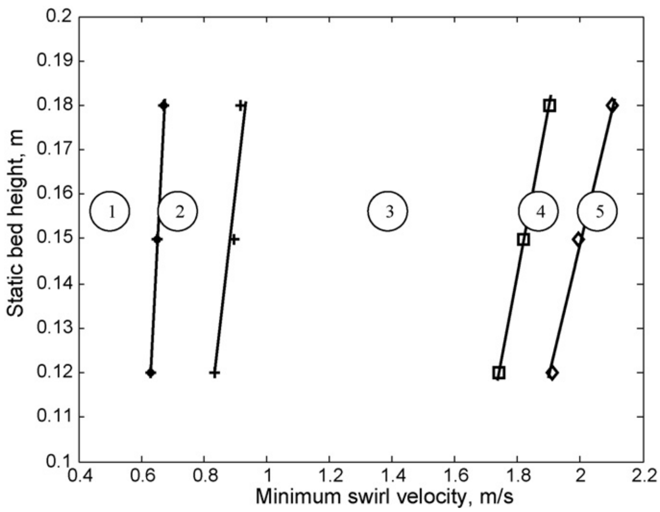


Fig. 4. Phase diagram: (material: Ragi seeds; D_c = 0.1 m, D_i = 0.01 m, N = 3). The various symbols represent: *V_1; + V_2; □ V_3; ◊ V_4. (1. static bed, 2. formation of fountains, 3. partial swirling, 4. stable swirling, 5. expanded swirling).

flow the bed instability has been found to be increasing further. Such an unstable bed behavior may be due to a possible mechanism as explained below.

For a given column diameter, fluid inlet diameter and number of fluid inlets, when a given bed of solids are in complete swirled state, the fluid jet will have only the tangential component of velocity when it just emerges from the fluid inlet; as the fluid moves up taking a spiral path along the column wall, entire bed of solids are kept in a swirled state, but the tangential component of jet velocity continues to decrease along its flow path. At a certain vertical height in the column the tangential component of velocity may die out completely and all the fluid moves up vertically. However, the average value of tangential component may be sufficient to keep the entire bed in a completely swirled state. So, for a given column configuration there has to be a maximum weight of solids which may be supported by the average tangential component of fluid velocity to keep the entire bed of solids in a complete swirled state. This upper limit of bed weight can be expressed as a static bed height called the maximum swirlable bed depth (MSwBD). Beyond this upper limit,

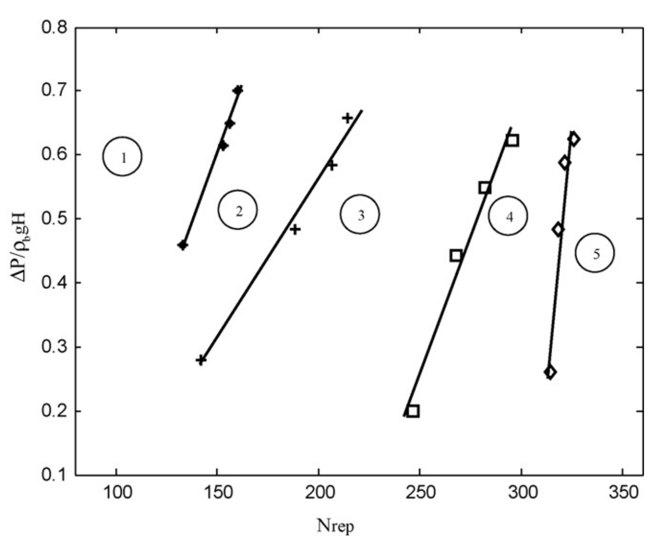


Fig. 5. Phase diagram: (material: polymer1; D_c = 0.1 m, D_i =0.01 m, N = 3). The various symbols represent: * ΔP_1, N_{rep1}; + ΔP_2, N_{rep2}; □ ΔP_3, N_{rep3}; ◊ ΔP_4, 1N_{rep4}. (1. static bed, 2. formation of fountains, 3. partial swirling, 4. stable swirling, 5. expanded swirling).

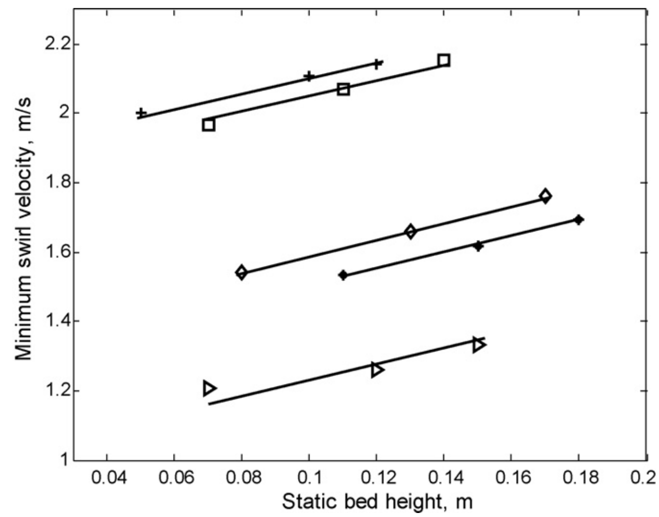


Fig. 6. Influence of static bed height on minimum swirl velocity. The various symbols represent: + polymer 2; D_c = 0.08 m, D_i = 0.01 m, N = 3; □ wheat; D_c = 0.1 m, D_i = 0.012 m, N = 4; ◊ Jowar seeds; D_c = 0.1 m, D_i = 0.01 m, N = 4; * polymer 1; D_c = 0.1 m, D_i = 0.01 m, N = 3; Δ Raagi seeds; D_c = 0.1 m, D_i = 0.008 m, N = 4.

a given set of fluid jets cannot achieve stable swirling of entire bed of a given solids in a given column, but the bed exhibits instability at all fluid velocities as observed in this work and reported above.

This upper limit may depend on the column diameter, fluid inlet diameter, number of fluid inlets, fluid and solid properties. However, these upper limits are not quantified in this work. It may be noted that such an upper limit for stable operation does not exist for conventional fluidized beds. However, it exists for single spouted beds [10] and multiple spouted beds [11].

3.4. Minimum swirl velocity

The point K in Fig. 3 is considered to be the point of minimum swirl velocity and hence defined as the "minimum superficial velocity of the fluid at which the bed changes from stable swirling to partial swirling upon the gradual decrease in the flow rate".

This velocity is found to be dependent on static bed height, fluid inlet size, number of fluid inlets, column diameter and particle properties.

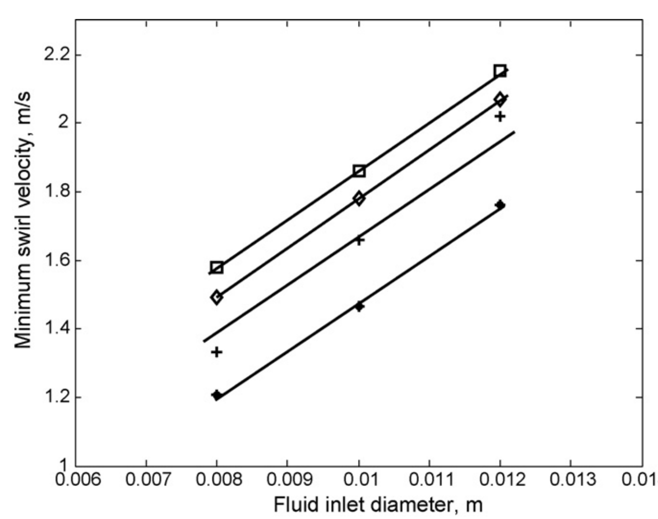


Fig. 7. Influence of fluid inlet size on minimum swirl velocity. The various symbols represent: □ wheat, D_c = 0.1 m, N = 4, H = 0.14 m; ◊ wheat, D_c = 0.1 m, N = 4, H = 0.11 m; + Raagi seeds, D_c = 0.1 m, N = 4, H = 0.15 m; *Raagi seeds, D_c = 0.1 m, N = 4, H = 0.07 m.

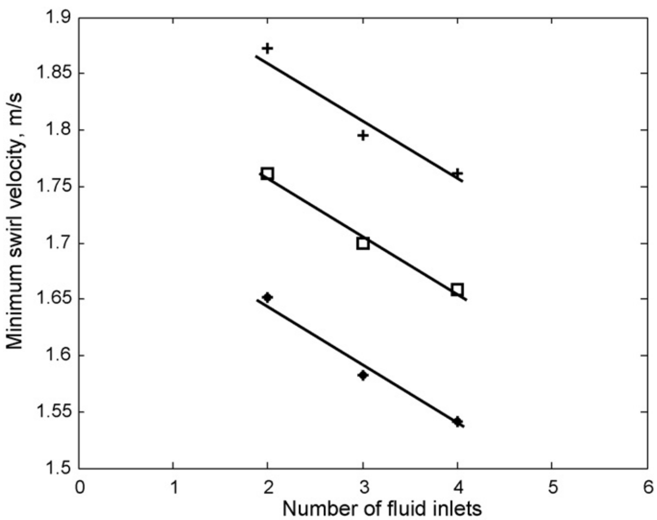


Fig. 8. Influence of number of fluid inlets on minimum swirl velocity; (material: Jowar seeds; $D_c = 0.1$ m, $D_i = 0.01$m); the various symbols represent: * $H = 0.08$ m, □ $H = 0.13$ m, + $H = 0.17$ m.

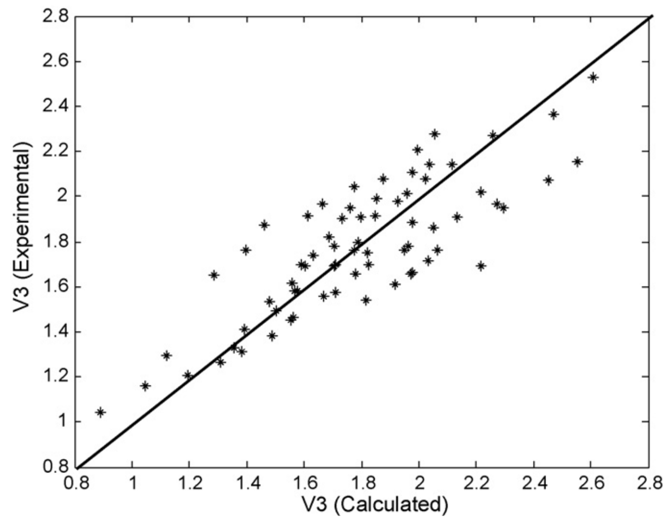


Fig. 10. Comparison of experimental and calculated minimum swirl velocities.

As the static bed height increases, the quantity (bed weight) of solids increases, which obviously requires more fluid quantity and energy to swirl fluidize the solids, and hence increased fluid velocities as shown in Fig. 6. For a given set of conditions, as the fluid inlet size increases, the fluid jet velocity decreases and the momentum of gas jet also decreases. So, higher flow rates would be required to achieve swirling with larger diameter fluid inlets as shown in Fig. 7.

As the number of fluid inlets (N) increases, the solids quantity in each of N segments of the bed decreases and the observed total fluid rate is found to be decreasing. A reduction in flow rate leads to reduction in velocity, which in turn results in higher static pressure at column bottom. Hence, because of the combined effect of reduced quantity of solids in each segment and increased static pressure at column bottom, as N increases, the transition to swirl bed can occur at lower superficial velocities as shown in Fig. 8.

For a given set of conditions, the quantity of solids (bed weight) increases and fluid superficial velocity decreases as the column diameter increases. A reduction in fluid velocity leads to an increase in static pressure at the column bottom. Hence, though the bed weight increases, an increase in static pressure at the column bottom helps in transforming the static bed into a swirled bed at a lower superficial velocity as observed in Fig. 9.

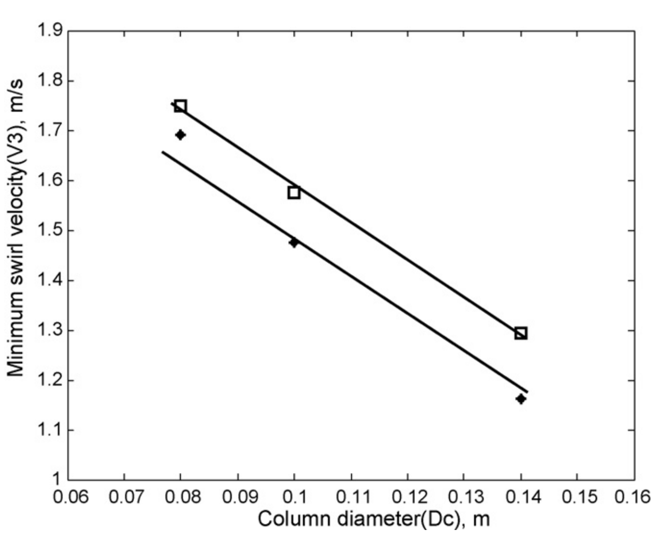


Fig. 9. Influence of column diameter on minimum swirl velocity; (material: Raagi seeds; $D_i = 0.1$ m, $N = 4$); the various symbols represent: * $H = 0.07$ m, □ $H = 0.11$ m.

3.5. Empirical correlation

Based on 71 experimental data, the minimum swirl velocity has been correlated in terms of modified Dean's number and the following equation is proposed.

$$\mathrm{De} = 22.29\left(\frac{H}{D_c}\right)^{0.1679}\left(\frac{D_i}{D_c}\right)^{1.2080}(\mathrm{Ar})^{0.3919}$$

for $90 < \mathrm{De} < 680$; $0.2 < H/D_c < 1.9$; $0.07 < D_i/D_c < 0.13$; $1.6 \times 10^5 < \mathrm{Ar} < 24 \times 10^5$.

The equation is formulated using regression analysis with the help of MAT LAB. The square root mean deviation between the calculated and experimental values of De is found to be 11.10% and the correlation coefficient is 0.96. Using the above equation the minimum swirl velocities are computed and compared with experimental values; Fig. 10 shows the comparison.

3.6. Comparison between conventional fluidized bed (CFB) and swirled fluidized bed (SFB)

A CFB is typically used when the solids are smaller in size, preferably <1 mm. When the particles are coarser and fluid happens to be a gas, problems like bubble formation, gas channeling, solids segregation and slugging of bed are encountered in the operation of a CFB. These problems will lead to poor quality fluidization, improper mixing of phases and poor transfer rates. These problems related to poor quality fluidization are not observed in SFBs studied in this work, where coarse solids are fluidized using air as fluidizing medium. In a CFB, the minimum fluidization velocity depends on particle and fluid properties only, but not on static bed height and column diameter; whereas in a SFB all these parameters influence the minimum swirl velocity. Also, there is no upper limit of static bed height in the operation of a CFB, whereas there exists a MSwBD in SFB.

For a given industrial application it is necessary for us to find the minimum fluid flow rate required to keep a given bed of solids in a fluidized state; larger this rate more the fluid consumption hence more would be the operating costs. So, it may be useful to compare minimum fluid velocities needed in CFB and SFB.

Table 2
Experimental results of minimum swirl velocity.

Sl. no.	Solid material	D_c (m)	D_i (m)	N	H (m)	V_3 (m/s)[a]	V_f (m/s)[b]
1.	Jowar seeds	0.1	0.01	2	0.08	1.652	1.130
2.	Polymer 1	0.08	0.01	2	0.10	1.912	0.864
3.	Raagi seeds	0.08	0.01	2	0.09	2.077	0.630
4.	Polymer 1	0.1	0.01	3	0.11	1.535	0.864
5.	Raagi seeds	0.1	0.01	3	0.15	1.821	0.630
6.	Wheat	0.1	0.01	3	0.14	1.765	1.228
7.	Polymer 2	0.08	0.01	3	0.12	2.276	1.088
8.	Jowar seeds	0.08	0.01	3	0.11	2.273	1.130
9.	Polymer 1	0.1	0.01	4	0.11	1.782	0.864
10.	Raagi seeds	0.1	0.01	4	0.15	1.659	0.630
11.	Jowar seeds	0.08	0.01	4	0.08	2.109	1.130
12.	Wheat	0.1	0.008	4	0.15	1.578	1.228
13.	Raagi seeds	0.1	0.008	4	0.15	1.332	0.630
14.	Raagi seeds	0.1	0.012	4	0.12	1.906	0.630
15.	Raagi seeds	0.08	0.01	4	0.11	1.949	0.630
16.	Raagi seeds	0.14	0.01	4	0.11	1.293	0.630

[a] Experimental – present work.
[b] Calculated – using Chitester et al. equation [3].

For the solids used in this work, the minimum fluidization velocities are computed using the following equation proposed by Chitester et al. [3]

$$Re_{p.mf} = [(28.7)^2 + 0.0494(\mathrm{Ar})]^{0.5} - 28.7$$

These are compared with the observed values of minimum swirl velocities obtained in this work. All results are given elsewhere [6], but some typical data are shown in Table 2. The minimum swirl velocities are found to be 1.2–3.3 times the minimum fluidization velocities for the range of variables used in this work.

It should be noted here that the solids used by Chistester et al. [3] in their work for developing the equation given above are finer (<1 mm size) when compared to the coarser solid particles used in this study; however, the particle densities are comparable. Hence, the minimum fluidization velocities in CFB computed here for coarse solids may not be very accurate. Nevertheless a comparison of minimum velocities is done just to show that the fluid consumption would always be more for swirl fluidizing coarse solid particles.

4. Conclusions

The following conclusions can be drawn based on the results obtained in this study:

1. A satisfactory operation of a swirl fluidization unit as designed and operated in this work is possible.
2. The minimum swirl velocity is dependent on static bed depth, fluid inlet diameter, column diameter, number of fluid inlets and particle properties; this is greater than the minimum fluidization velocity needed in conventional beds.
3. There exists a maximum swirlable bed depth beyond which stable swirling of entire bed is not possible.

Appendix A. Nomenclature

Ar	Archimedes number, $gd_p^3\rho_f(\rho_s-\rho_f)/\mu^2$ (dimensionless)
D_c	column diameter (m)
De	modified Dean's number, $N_{\mathrm{rep}}\,(P/D_c)^{0.5}$ (dimensionless)
P	pitch, $\pi D_c/N$ (m)
D_i	fluid inlet diameter (m)
H	static bed height (m)
d_p	particle size (m)
g	acceleration due to gravity (m/s^2)
N	number of fluid inlets
N_{rep1}	particle Reynolds number at V_1, $d_pV_1\ \rho_f/\mu_f$ (dimensionless)
N_{rep2}	particle Reynolds number at V_2, $d_pV_2\ \rho_f/\mu_f$ (dimensionless)
N_{rep3}	particle Reynolds number at V_3, $d_pV_3\ \rho_f/\mu_f$ (dimensionless)
N_{rep4}	particle Reynolds number at V_4, $d_pV_4\ \rho_f/\mu_f$ (dimensionless)
ΔP_1	bed pressure drop at V_1 (N/m^2)
ΔP_2	bed pressure drop at V_2 (N/m^2)
ΔP_3	bed pressure drop at V_3 (N/m^2)
ΔP_4	bed pressure drop at V_4 (N/m^2)
$\mathrm{Re}_{p.mf}$	particle Reynolds number at minimum fluidization velocity, $d_pV_f\ \rho_f/\mu_f$
V_1	minimum superficial velocity for fountain formation (m/s)
V_2	minimum superficial velocity for fountain region (m/s)
V_3	minimum superficial swirl velocity (m/s)
V_4	minimum superficial velocity for expanded swirling (m/s)
V_f	minimum fluidization velocity (m/s)

Greek letters

ρ_b	Bulk density (kg/m^3)
ρ_f	Fluid density (kg/m^3)
ρ_s	Solid density (kg/m^3)
μ_f	Fluid viscosity (kg/ms)

Subscripts

b	bed of solids
s	solid phase
f	fluid phase

References

[1] B. Sreenivasan, V.R. Raghavan, Hydrodynamics of a swirling fluidized bed, Chem. Eng. Process. 41 (2002) 99–106.
[2] C.-s. Chyang, Y.-C. Lin, A study in the swirling fluidizing pattern, J. Chem. Eng. Jpn. 35 (6) (2002) 503–512.
[3] D.C. Chitester, R.M. Kornosky, L.S. Fan, Characteristics of fluidization at high pressure, Chem. Eng. Sci. 39 (1984) 253–261.
[4] A.F. Dolidovich, Hydrodynamics and interface heat transfer in swirled spouted bed, Can. J. Chem. Eng. 70 (1992) 930–937.
[5] S. Eiamsa-ard, Y. Kaewkohkiat, W. Lelaphatikul, C. Thianpong, P. Promvonge, Experimental investigation of combustion characteristics in a multi-staging vortex combustor firing rice husk, Int. Commun. Heat Mass Transfer 35 (2008) 139–148.
[6] S. Harish Kumar, Hydrodynamic studies on swirled gas–solid contacting devices. Ph.D. thesis, Manipal University, Manipal, India, 2009.
[7] R. Kaewklum, V.I. Kuprianov, Experimental studies on a novel swirling fluidized-bed combustor using an annular spiral distributor, Fuel 89 (2010).
[8] R. Kaewklum, V.I. Kuprianov, P.L. Douglas, Hydrodynamics of air–sand flow in a conical swirling fluidized bed: a comparative study between tangential and axial air entries, Energy Convers. Manage. 50 (2009) 2999–3006.
[9] J.K. Lee, C.G. Hu, Y.S. Shin, H.S. Chun, Combustion characteristics of a two-stage swirl-flow fluidized bed combustor, Can. J. Chem. Eng. 68 (1990) 824–830.
[10] K.B. Mathur, N. Epestein, Spouted Beds, Academic Press, New York, 1974.
[11] D.V.R. Murthy, P.N. Singh, Minimum spouting velocity in multiple spouted beds, in: Nicholas P. Cheremiscinoff (Ed.), Mixed-flow Hydrodynamics: Advances in Engineering Fluid Mechanics Series, Gulf Publishing Co., Houston, TX, 1996, pp. 741–758 (Chapter 32).
[12] W. Zhang, A numerical study of solid suspension layers in a swirl fluidized bed reactor, Int. J. Chem. React. Eng. 7 (2009) A15.
[13] J.D. Wilde, A.de. Broqueville, Experimental investigation of a rotating fluidized bed in a static geometry, Powder Technol. 183 (2008) 426–435.