Techno-Science
Scientific Journal of Mehmet Akif Ersoy University
www.dergipark.gov.tr/sjmakeu

Original Research Article

hosted by Turkish JournalPark ACADEMIC

Testing of New design of Mesh-coupled Axial Blade Distributor for Swirling Fluidization Operation

Muhammad Yasin Naz[1*], Shazia Shukrullah[1], Shaharin Anwar Sulaiman[2], Afsin Gungor[3]

[1]*Department of Physics, University of Agriculture, 38040 Faisalabad, Pakistan.*
[2]*Department of Mechanical Engineering, Universiti Teknologi PETRONAS, Malaysia.*
[3]*Department of Mechanical Engineering, Akdeniz University, Turkey.*

ARTICLE INFO

Article History

Received	:	20/11/2019
Revised	:	30/12/2019
Accepted	:	30/12/2019
Available online	:	30/12/2019

Keywords
Swirling fluidized bed, mesh-type distributor, particle image velocimetry, hydrodynamics of particles.

ABSTRACT

All the fluidized beds have air distributor as their major component, which can have several variances. The distributor design worth proper investigations as it affects the quality of the fluidization. In this study, Particle Image Velocimetry (PIV) of a SFB, operated with a mesh-coupled air distributor, was carried out from top and side of the bed. The velocity vector fields of the particles and then the average velocities at different locations of the bed were generated by considering particle size and density as operational parameters. The experimental finds showed that there exist several partially differentiable layers during a swirling operation of the longer beds. The segregated bed includes a thin layer of particles swirling at the bottom of the bed and a thick layer in the middle of the bed followed by a slight bubbling at the top. It was also observed that the overall velocity of the particles increased with an increase in superficial air velocity. The uniform swirling of 3, 4 and 6 mm particles happened at superficial air velocities of 1.6, 1.8 and 2.2 ms-1, respectively.

1. INTRODUCTION

Fluidization is a technique in which solid particles in a bed are transformed from a static solid-like state to a dynamic fluid-like state and the bed behaves like a liquid [1]. When the fluid flows through the bed, it applies a force on the particles, which is referred to as the drag force. As the rate of vertically upward flow increases, the drag force exerted on the particles also increases and at an instant it becomes large enough to unlock the arrangement of the particles. The FB's designs have many types, as developed by the researchers in the past [2-4]. They include: circulating fluidized bed, centrifugal fluidized bed, tapered fluidized bed, vibro-fluidized bed, spouted fluidized bed [5]. These beds have been in use for past several decades. They have been applied in drying, combustion, gasification of biomass, surface treatment, heat and catalytic cracking, mass transfer, surface coating [6, 7]. But these conventional designs of fluidized beds beside of their many advantages, also have many shortcomings which affect their performance, like: they have moving parts, high pressure drop, complex hydrodynamics, inefficient particle mixing and limitation of using for variety of particle sizes etc. None of the design can be considered as optimized even for a single chemical or mechanical processes because of above discussed shortcomings. These shortcomings affect the quality of fluidization, yield of the process, energy consumption etc. The conventional fluidized beds also have limit on the size of the solid particles to be fluidized, partial particles mixing, high pressure drop and complex hydrodynamics [8]. The high pressure drop across the distributor is an issue requiring special emphasis as it determines the power utilization for pumping.

SFB is the latest version of the well-known bubbling fluidized bed. It can overcome the defects and shortcomings of above mentioned fluidized beds. The feature of SFB distinguishing it from the conventional fluidized beds is its swirling mode of operation. It has many other characteristics making it superior to conventional fluidized beds, which are: no moving parts, outstanding mixing of particles, uniform thermal distribution and excellent heat and mass transfer. The swirling motion of bed particles makes it possible to fluidize particles at lower pressure drops in contrast to conventional fluidized beds, having approximate value of 350 mm of water for conventional beds.

* Corresponding Author: yasin306@uaf.edu.pk

To cite this article: Naz, M.Y., Shukrullah, S., Sulaiman, S.A., Gungor, A., (2019). Testing of New design of Mesh-coupled Axial Blade Distributor for Swirling Fluidization Operation. *Techno-Science*, vol. 2, no. 4-Special Issue p. 111-118.

The main issue still under the study of research community is the high pressure drop across the distributor. It has a great significance as pressure drop determines the power used in pumping by the blower, so disturbs the economy of the operation. The pressure drop across distributor during the swirling motion of beads in SFB increases with increase in superficial velocity but decreases in case of conventional beds. The reason is that the bed pressure is caused by the centrifugal weight of bed particles [9]. Hence, a new design of distributor for SFB operation is needed to be introduced to make a better operation of SFBs. Such beds have not been fully developed, even though they have been applied in various applications. So, a new design must be developed for improvement in operation efficiency. Mesh-type distributor is the newest design and not much reported before. There is scant literature on the operation of SFBs utilizing the mesh-type distributors. Hence, a study was needed to investigate the particle hydrodynamics to make a better understanding of the bed response to this design of a SFB. The main objective of the presented work was to study the effects of the particle size and density on the hydrodynamics of a bed by using a mesh-type distributor. The effect of the bed weight on the fluidized particle velocity was also investigated using a PIV technique

2. Experimental Setup

The experimental setup of this study consisted of a bed column, a gas distributor and the imaging apparatus. It is a laboratory scale setup employed to study the swirling fluidized bed operation. A plexiglass cylinder of 50 cm length and 300 cm diameter was used as a bed column, as shown in Fig 1. The air distributor was made of wire mesh having pore size of 2.8×2.8 mm2 and air entry angle of 45°. The distributor was placed at the top opening of the wind box. The function of the wind box, as shown in experimental setup, is to reduce the air fluctuations and to ensure smooth air flow from blower to air distributor and into the bed. For the said purpose, the height of plenum was set to 500 mm. The angle of air entry into the plenum also controls the air fluctuations therefore the plenum chamber was designed for air entry at tangential level without having any central body in the plenum.

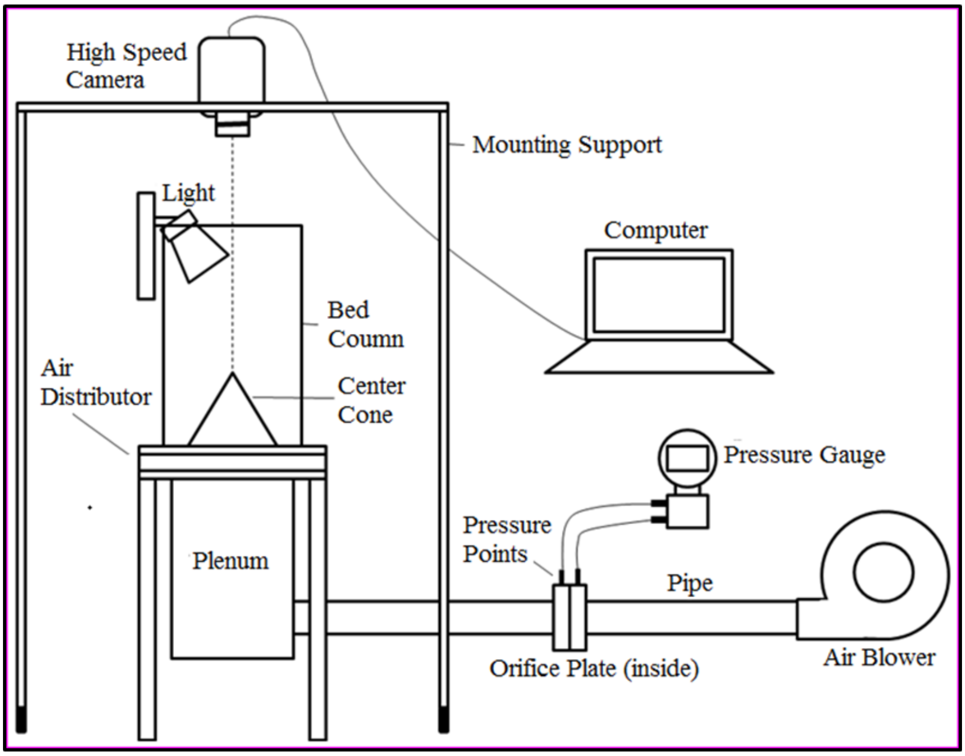

Fig 1. Schematic of the custom-made SFB.

As shown in Fig 2, a hybrid model of the air distribution plate was constructed whereby a mesh type distributor was coupled with an annular blade distributor. A small sized wire mesh was fixed at the top of an annular blade distributor, which not only make the system suitable for study of larger particles' dynamics but fine particles as well. A hollow metal cone of 15 cm base diameter was placed in the center of the distributor to avoid particles' clogging. The base annular spiral distributor was consisted of metal blades of trapezoidal shape with openings between successive blades. The blades were overlapped to form an annulus between inner and outer radii of the hybrid air distributor. The blades were inclined at 45° for tangential air entry into the bed.

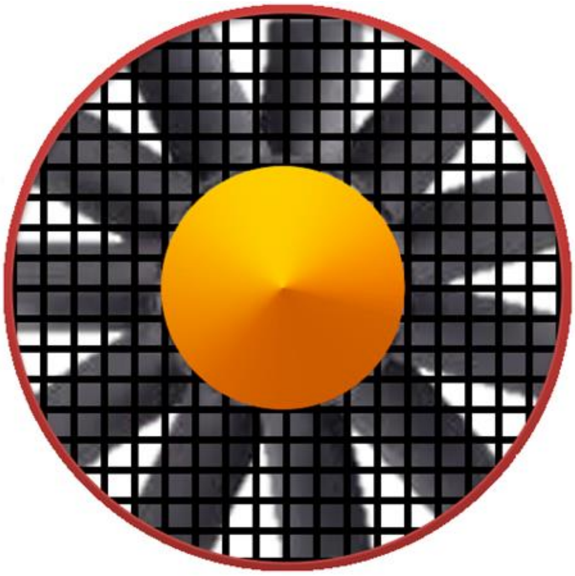

Fig 2. Top view of the mesh-type air distributor fitted with annular blades.

Table 1. Details of the bed material.

Particle size	D = 3mm Mass = 0.025 g	D = 4mm Mass = 0.1 g	D = 6mm Mass = 0.25 g	
Particle Density	D = 3mm Density = 1.77 g/cm3	D = 4mm Density = 2.99 g/cm3	D = 6mm Density = 2.21 g/cm3	D = 6mm Density = 2.65 g/cm3
Superficial Velocity	Velocity ranges in steady state swirling regime of operation and early bubling regime of operation			
Bed Weight	500, 750, 1000, 1250			

PIVlab measurements were conducted on a set of images of the swirling bed taken with a high-speed camera. As the central cone and its shade was there, it was difficult to photograph the whole bed with same brightness. To keep the measurements simple, only half of the bed was photographed for PIV analysis, as shown in Fig 3. A phantom M 110 high-speed camera was used to photograph the swirling bed at a maximum frame rate of 1000 fps. The illumination arrangements were adjusted according to the frame rate and the velocity of the particles. The resolution of camera was set to 856×846. The coordinates of the displacement and velocity of bed particles were determined by taken an image of the static bed and scaling the diameter of the bed column and the central cone in the image.

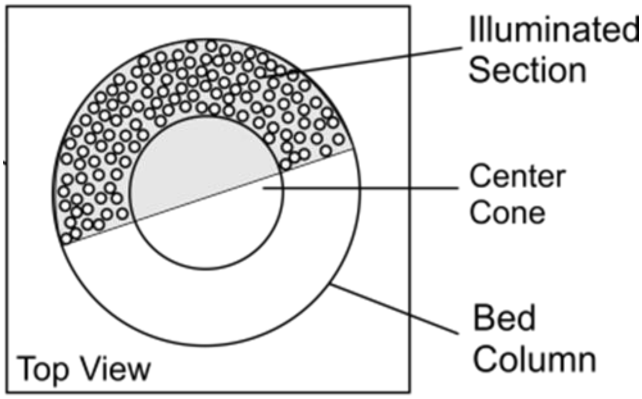

Fig 3. The shaded half are of the swirling bed for PIV analysis.

PIV profiling of the grabbed images was performed to study the hydrodynamics of the particles at the top and side of the bed. The PIV profiles not only provided information about the angular velocity but also swirling motion of the bed material at the top and side of the bed. For PIV analysis, the images were firstly process through a Binary Image Cross-Correlation method in MATLAB software. These binary converted images were showing white tracer particles on a dark background of the black particles. This color contrast given by the minority tracer particles and majority black particles was helpful in tracking the tracer particles and generating the reliable data through PIV analysis.

3. RESULT AND DISCUSSION

Fig 4 shows a typical velocity vector field of particles swirling at the top of the bed. For this velocity vector field, the bed weight was measured about 750 g, the pressure drop was 103 mmH2O, particles size was 6 mm and superficial air velocity was 2.28 m/s. This field predicts slightly nonuniform particle velocity. The regions where the particles had the highest velocity are marked by the orange color followed by the yellow, green, sky blue and dark blue. The fast-moving particles were found mostly in the middle area of the bed. The blue color showed the lowest velocity of the particles located near the bed walls or cone walls. The lowest velocity near the walls was due to frictional effects associated with the centrifugal force on the particles [10, 11].

The different patterns of particles velocity in the different locations of the bed were unaffected by the change in superficial air velocity. In the same manner, the velocity profiles of the particles were generated using of 500 g bed weight while increasing superficial air velocity. Each velocity profile presented a different area in the interrogation window, as is lustrated in Fig 5. When fluidizing agent is denser or pressure is high, or bed particles are very light, the bed undergoes relatively stable expansion, resulting in fluidization of the particles. However, as observed in the presented work, it is still difficult to deal with the fluidization of the fine particles. For 3 mm particles, the particle to particle cohesive forces were much higher than the gravitational force.

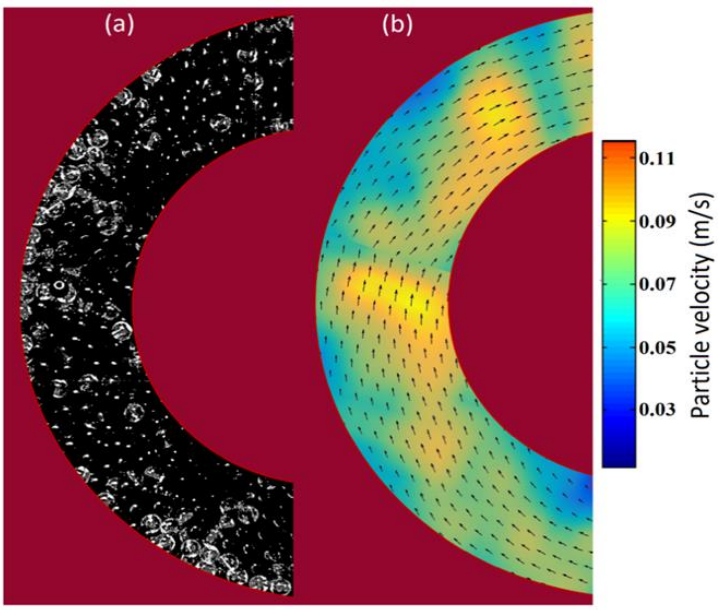

Fig 4. Typical velocity vector field of 6 mm particles.

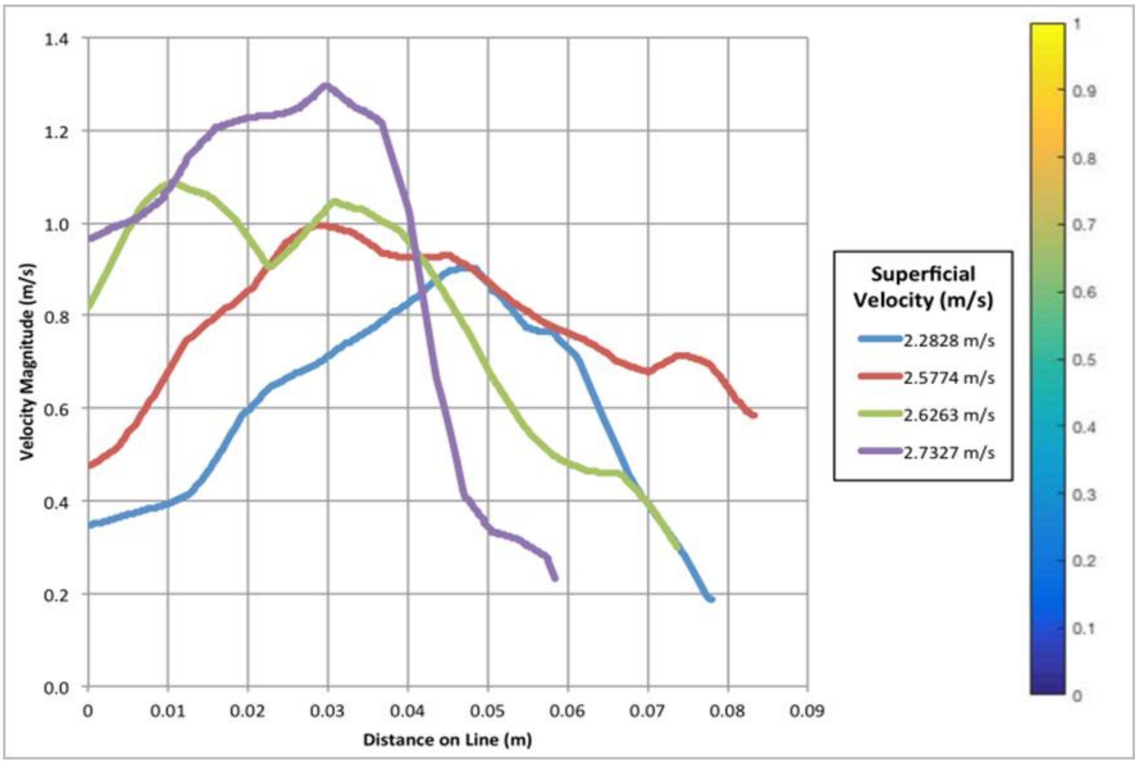

Fig 5. Velocity profiles of a swirling bed at different superficial air velocities.

At lower superficial velocities, particles in the bed remained intact by letting the fluid to pass through the bed via blowing channels. In response, the bed started to bubble to some extent. In case of heavier particles, the bed remained settled at lower superficial velocities and did not underwent unwanted bubbling. With an increase in superficial velocities, all particles sizes and weights entered into uniform swirling regime expect 3 mm particles. Some low level of bubbling still remained in the bed of 3 mm particles which changed to turbulence at higher superficial velocities.

The velocity variations of 3 mm, 4 mm and 6 mm spherical particles with superficial air velocity are shown in Fig 6. Two differentiable layers of particles were observed in longer beds. A thin layer near the air distributor, which swirls at faster rate and a thick layer covering the rest of the bed volume above the first layer. The particles in the layer were swirling slightly slower as compared to the particles in the first layer. However, in case of 3 mm, there were also some fast decaying bubbles in the bed. The overall trend of bed particles' velocity was that increasing superficial velocity of air increased their velocity. It was observed that 3 mm particles started to swirl first, followed by 4 mm and 6 mm particles. When the bed weight was kept 500 g, the 3 mm spherical particles started to swirl at a superficial air velocity of 1.6 m/s while 4 mm and 6 mm particles at 1.8 m/s and 2.2 m/s, respectively. Also, the stable swirling of the smaller particles ended earlier as compared to the heavier particles.

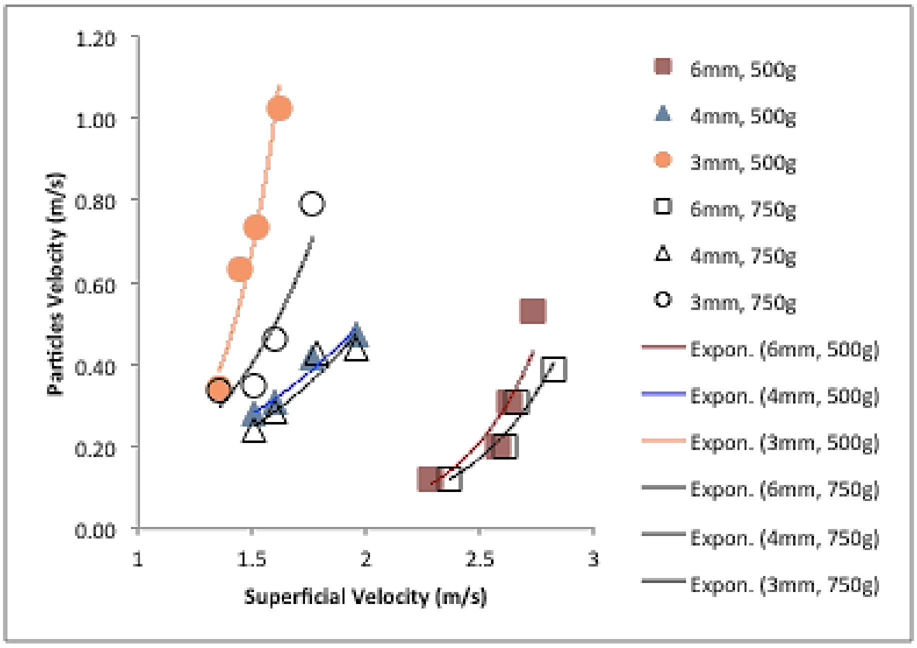

Fig 6. Variation of average velocity for different sizes of particle with superficial air velocity.

It was obvious to see that heavier beds needed higher superficial velocities to fluidize than the lighter beds where large drag force was required to lift the particles. Since mathematics of swirling fluid beds is not fully developed so far, it is generally difficult to calculate the minimum fluidization velocity of the bed. however, it is anticipated that from packed bed to incipient fluidization phases, a SFB acts similar to the conventional fluidized beds. Therefore the equations for minimum fluidization velocity of a conventional bed would be valid SFBs [1]. The minimum fluidization velocity for swirling of 3 mm, 4 mm and 6 mm particles was estimated about 8.12 m/s. 1.04 m/s. 1.32 m/s, respectively.

The average particle velocity for different bed weights but of single sized particles is shown in Fig 7. As the bed weight was increased, the average particle velocity was observed to decrease. An increase in bed weight caused a decrease in air velocity as the air percolated through the interstices of particles. It was due to the transfer of momentum to the particles. Subsequently, the velocity of the particle in the uppermost layer of the bed remained slower as compared to the bottom of the bed. The increase in the bed weight caused the particle velocity less sensitive to the changes in superficial air velocity. The graph shows that the range of the particle velocities for 500 g bed was larger than the other bed weights and the particle velocity showed a decreasing trend with an increase in bed weight.

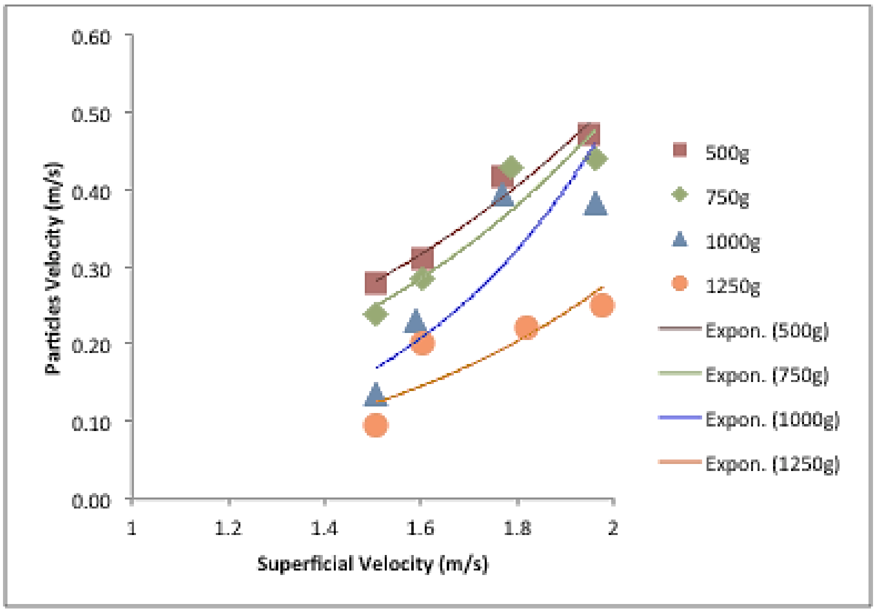

Fig 7. Variation of average particle velocity of different bed weights with superficial air velocity.

The densities of particles are observed in general to affect the bed hydrodynamics, if the size is same [2, 12]. In Fig 8, the graph is showing the effect of different densities of 6 mm particles to describe the density dependent bed hydrodynamics. The lighter the particles, the low superficial air velocity is required to fluidize the bed as compared to the bed of heavier particles. But in the comparison of particles of different sizes, as in the case of fluidization of 4 mm and 6 mm particles, the 4 mm particles required lower superficial air velocity than 6 mm particles although the density of 4 mm particle was high. Additionally, the increment in velocity of 4 mm particles was observed to be exponentially over superficial air velocity. The velocity of individual particles increased with increase in superficial air velocity but decreased with increase in particle density. The reason of such behavior was that the weight and close packing of the heavier particles played their role to lessen the velocities.

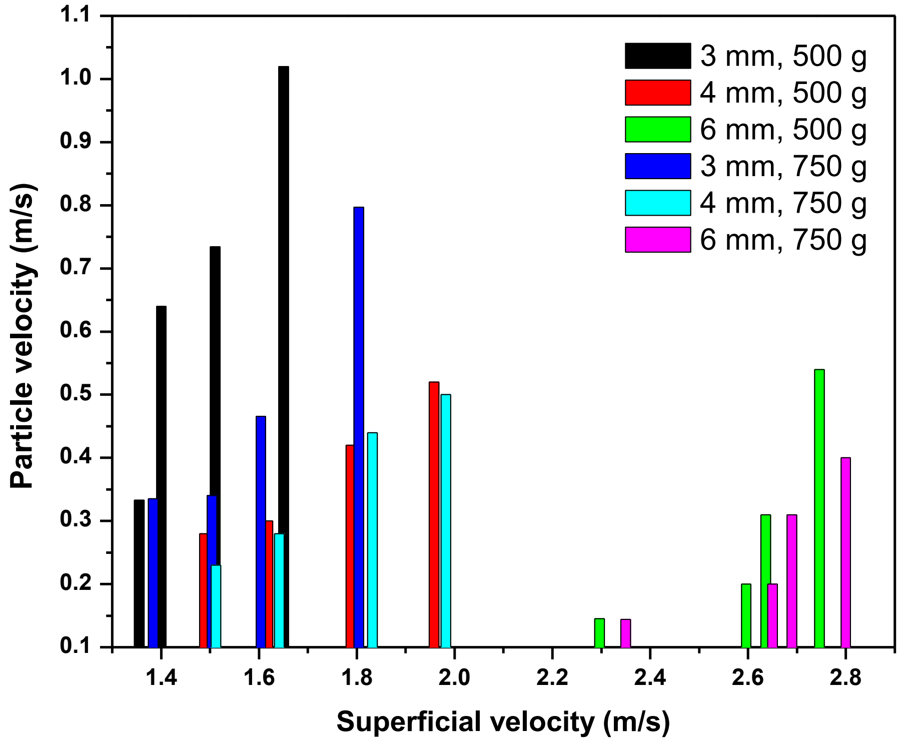

Fig 8. Variation of average velocity of different density particles with superficial air velocities.

4. CONCLUSIONS

PIV profiling of a SFB, operated with a wire-mesh distributor, was performed both from top and side wall. It was concluded that the velocity of the particles increases with the superficial air velocity and decreases with bed height. For 3 mm particles, at lower superficial velocities, particles in the bed remained intact by letting the fluid to pass through the bed via blowing channels. In response, the bed started to bubble to some extent. In case of heavier particles, the bed remained settled at lower superficial velocities and did not underwent unwanted bubbling. With an increase in superficial velocities, all particles sizes and weights entered into uniform swirling regime expect 3 mm particles. Some low level of bubbling still remained in the bed of 3 mm particles which changed to turbulence at higher superficial velocities. Full swirling of 3, 4 and 6 mm particles happened at superficial air velocities of 1.6, 1.8 and 2.2 ms-1, respectively. For any case, particles, motion in the middle of the bed showed the highest velocities than the regions near the bed boundaries. Some differentiable moving layers in the bed for each particle size were observed with an increase in bed height. The particle velocity found decreasing over bed height. For all the bed heights, the particle velocity at the top of the bed was found lower than the side of the bed. This different becomes more significant for larger bed heights showing that the bed might be composed of multi-layers.

REFERENCES

[1]. Sulaiman, S. A. Miin, C. S. Naz, M. Y. and Raghavan, V. R., "Particle Image Velocimetry of a Swirling Fluidized Bed at Different Blade Angles," Chemical Engineering & Technology, vol. 39, pp. 1151-1160, 2016.

[2]. Collado, F. J., "New one-dimensional hydrodynamics of circulating fluidized bed risers," Granular Matter, vol. 18, p. 78, 2016.

[3]. Geldart, D. and Baeyens, J., "The Design of Distributors for Gas-fluidized Beds," Powder Technology, vol. 42, pp. 67-78, 1985.

[4]. Jittanit, W. Srzednicki, G. and Driscoll, R. H., "Comparison Between Fluidized Bed and Spouted Bed Drying for Seeds," Drying Technology, vol. 31, pp. 52-56, 2013.
[5]. Yang, W. C., "Bubbling Fluidized Beds", in Handbook of Fluidization and Fluid-Particle Systems," Ed. Yang, W.-C., Marcel Dekker, N.Y., pp. 53-111, 2003.
[6]. Othomer, D. F., "Background,History and Future of Fluid bed systems, Fluidization,," Reinhold publishing corporation, New York, pp. 102-115, 1956.
[7]. Shukrullah, S., Mohamed, N. M., Shaharun, M. S., Saheed, M. S. M. and Irshad, M. I., "Effect of CVD Process Temperature on Activation Energy and Structural Growth of MWCNTs," Metallurgical and Materials Transactions A, vol. 47, pp. 1413-1424, 2016.
[8]. Miin, C. S., Sulaiman, S. A., Raghavan, V. R., Heikal, M. R. and Naz, M. Y., "Hydrodynamics of multi-sized particles in stable regime of a swirling bed," Korean Journal of Chemical Engineering, vol. 32, pp. 2361-2367, 2015.
[9]. Venkiteswaran, V. K., Jun, G. J., Sing, C. Y., Sulaiman, S. A. and Raghavan, V. R., "Variation of Bed Pressure Drop with Particle Shapes in a Swirling Fluidized Bed," Journal of Applied Sciences, vol. 12 pp. 2598-2603, 2012.
[10]. Sanvitale, N. and Bowman, E. T., "Using PIV to measure granular temperature in saturated unsteady polydisperse granular flows," Granular Matter, vol. 18, p. 57, 2016.
[11]. Hagemeier, T., Roloff, C., Bück, A. and Tsotsas, E., "Estimation of particle dynamics in 2-D fluidized beds using particle tracking velocimetry," Particuology, vol. 22, p. 39-51, 2015.
[12]. Karen, E. D. and Matthias, S. T., "Focus on granular segregation," New Journal of Physics, vol. 15, p. 035017, 2013.

Techno-Science *Paper ID: 649903*