

A Study of the Hydrodynamics Behavior of Cylindrical Gas-Solid Fluidized Beds for pharmaceutical material “Paracetamol”

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ABSTRACT

The hydrodynamics behavior of gas - solid fluidized beds is complex and it should be analyzed and understood due to its importance in the design and operating of the units. The effect of column inside diameter and static bed height on the minimum fluidization velocity, minimum bubbling velocity, fluidization index, minimum slugging velocity and slug index have been studied experimentally and theoretically for three cylindrical columns of 0.0762, 0.15 and 0.18 m inside diameters and 0.05, 0.07 and 0.09 m static bed heights. The experimental results showed that the minimum fluidization and bubbling velocities had a direct relation with column diameter and static bed height. The minimum slugging velocity had an inverse relation with static bed height and a direct one with column diameter. There was no agreement between the experimental and calculated values of U_{mb} for $D_i=0.0762m$, this was a result to the assumption used in the correlation development. The fluidization index values were around 1 in all cases and that proved that the material is of Geldart type B. There was not a significant dependence of fluidization index and slug index on static bed height and column diameter.

Key words: gas-solid fluidization, cylindrical beds, Paracetamol.

دراسة للسلوك الهيدروديناميكي للأبراج المميعة الاسطوانية لمادة دوائية "الباراسيتامول"

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مدرس

كلية الهندسة - جامعة بغداد

الخلاصة

ان السلوك الهيدروديناميكي للأبراج المميعة نوع غاز- صلب معقد ويجب ان يحلل لأهميته في تصميم وتشغيل الوحدات. تمت دراسة تأثير القطر الداخلي و الارتفاع الستاتيكي للطبقة عمليا و نظريا لثلاثة ابراج اسطوانية بقطر 0.076, 0.15, 0.18 متر وارتفاع الطبقة 0.05, 0.07, 0.09 متر على سرعة التميع الدنيا وسرعة الفقاعات وسرعة slugging و معامل التميع

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ومعامل slug. أوضحت النتائج ان العلاقة مباشرة مع قطر العمود و ارتفاع الطبقة ما عدا سرعة slugging فالعلاقة عكسية مع ارتفاع الطبقة. لوحظ انه لا يوجد تطابق بين القيم العملية و النظرية لسرعة الفقاعات و slugging وهذا يعود للفرضيات بالمعادلات النظرية. قيم معامل التميع كانت بحدود 1 في جميع الحالات وهذا يعني ان مادة الباراسيتامول من نوع ب حسب تصنيف العالم كيلدارد. لا يوجد تأثير واضح لأرتفاع الطبقة المميعة و قطر العمود على كلا من معامل التميع و معامل slug. الكلمات الرئيسية: التميع غاز- صلب, الأبراج الاسطوانية, باراسيتامول.

1. INTRODUCTION

The application of gas- solid fluidized beds technique in chemical processes is widely used such as gas cleaning, blending of solids, heat sensitive materials drying, crude oil thermal cracking, gasification and combustion of waste material or coal, coating and granulation (Gajbliye, et al., 2013), mentioned that more applications of fluidization technique can be seen in circulation of solid particles, pneumatic transportation, chemical synthesis and chemical regeneration. (Sulaymon, et al., 2013), used fluidized bed for heavy metals removal by algal biomass. (Ebrahim, 2016) , worked on using fluidized beds in Fluoride ions removal from waste water using blue and green algae biomass. This technique needs stable and well defined operating conditions, maintaining these conditions is a major challenge in the design of fluidization beds because of limited understanding of the hydrodynamics characteristics of fluidized beds. (Jaiswal, et al., 2018) .

A fluidized bed means a bed of particles is transformed from static state to a dynamic one. (Yang, 2003), mentioned that there are six fluidization regimes for fluidized beds: 1- fixed bed 2-bubbling fluidization 3- slugging fluidization 4- turbulent fluidization 5-fast fluidization 6- pneumatic fluidization. The gas superficial velocity, when the frictional force between the gas and solid particles are counter balanced by the bed weight, is called minimum fluidization velocity. With an increase in gas velocity, fluidized system reaches the regime of bubbling fluidization where the formation and coalescing of bubbles which cause solid mixing, at the appearance of the first bubble the gas velocity is called minimum bubbling velocity. Further increase in gas superficial velocity causes the bubbles size and rise velocity to increase, they collapse and grow in size until they reached (0.3-0.6) of column diameter and the bed starts to slug, the gas velocity at this point is called minimum slugging velocity. The flow shifts from slug in to turbulent with further rise in gas velocity, the turbulent flow marked with the bubbles and slug absence in the bed and that followed by particles movement and transportation .(Jaiswal, et al., 2018) and (Agu, et al., 2017) mentioned that the transition from a type of flow to another depends on many factors i.e. static bed height, fluidizing gas flow rate, column diameter, particle density, particle size and others.

The hydrodynamics behavior is complex and determined with respect to the minimum fluidization velocity, the minimum bubbling velocity, fluidization index, particulate fluidization range, the minimum slugging velocity, bubbling bed index and slug index.



1.1 Minimum Fluidization Velocity (U_{mf}):

It is a very important parameter when the hydrodynamics of fluidized column is characterized. It can be defined as the gas superficial velocity at which the bed is fluidized. At this point the pressure drop is equal to the bed weight per unit cross sectional area. Minimum fluidization velocity depends on material properties, the geometry of the bed and gas properties and it is usually obtained experimentally. (**Ramos, et al., 2002**), mentioned to the most common equations used to calculate the theoretical minimum fluidization velocity:

$$U_{mf} = \frac{(\Phi dp)^2}{150} \left(\frac{\epsilon^3}{1-\epsilon} \right) \left(\frac{\rho_s - \rho_g}{\mu_g} \right) g \quad \text{for } Re_p < 20 \quad (1)$$

$$U_{mf}^2 = \frac{\Phi dp}{1.75} \left(\frac{\rho_s - \rho_g}{\mu_g} \right) g \epsilon^3 \quad \text{for } Re_p > 1000 \quad (2)$$

It can be noted from the equations that the column diameter and static bed height are not included among the parameters even though they have a large influence on minimum fluidization velocity, so the effect of each them should be studied experimentally.

(**Gunn, et al., 1997**), investigated gas-solid cylindrical fluidized beds. They used glass beads of (100-500) μm diameters, column diameters of 89, 290 mm and bed heights of (20,30,40,50) cm, the results showed that there was not a significant variation in the U_{mf} when the static bed height was increased. (**Hilal, et al., 2001**), studied the effect of bed diameter on U_{mf} for two cylindrical beds (0.29, 0.089) m and they concluded that the U_{mf} decreased with an increase in the bed diameter. (**Ramos, et al., 2002**), studied 2D rectangular gas-solid fluidized bed filled with different diameters of glass ballotoni ranges (160-700) μm and for different static bed heights (2,4,8,16,20,40,60) cm, they found that the U_{mf} decreased with increased bed width and increased when the bed height increased. (**Akhil, et al., 2010**), used spherical particles in cylindrical gas-solid fluidized bed, they noticed that the U_{mf} increased as the column diameter is reduced, or the bed height is increased. (**Escudero, 2010**), studied the effect of material density and static bed height on the U_{mf} for ground walnut shells, ground corncob and glass beads in a 3D cylindrical fluidized bed at different static bed heights and found that the U_{mf} approximately remained constant. (**Sarker, et al., 2012**), investigated the most effective parameters on gas-solid fluidized beds, these are bed diameter, static bed height and particle shape. They used laboratory scale systems having 3.5 and 12.4 cm inside column diameter and a static bed height ranges 2 to 7 cm, they observed that the U_{mf} decreased by decreasing the column diameter and there was not a significant dependence of U_{mf} on bed height. (**Yupeng, et al., 2017**), studied the effect of column inner diameter of (8,12, and 16) mm and different static bed heights for Geldart B particles behavior in micro fluidized beds, they concluded that both increasing static bed height and decreasing the diameter of the column increases the minimum fluidization velocity.

Therefore, based on the mentioned literatures the bed diameter and static bed height are among the parameters those affect on U_{mf} in gas-solid fluidized beds.

1.2 Minimum Bubbling Velocity (U_{mb}) and Fluidizing Index (FI):

By Increasing the gas velocity larger than the U_{mf} , the gas-solid system exhibit either bubbling or non-bubbling condition. The non-bubbling is known as homogeneous or particulate fluidization while the bubbling is referred as heterogeneous or aggregative



fluidization. The upper limit of the gas velocity at Particulate condition corresponds to the first bubble appearance and the velocity at this point is called minimum bubbling velocity. **(Abrahamsen, and Geldart, 1980)**. developed an equation for minimum bubbling velocity, it can be observed that the column diameter and static bed height are not among the equation parameters:

$$U_{mb} = \frac{2.07 d_p \rho_g^{0.06}}{\mu_g^{0.347}} \quad (3)$$

(Xiaotao, 1994), cleared that the U_{mb} depends mainly on particles properties and it is equal to minimum fluidization velocity for Geldart B type. **(Singh, and Roy, 2005)**, defined the minimum bubbling velocity as the gas superficial velocity where the first bubble appears. Also they mentioned that particulate fluidization lies between minimum bubbling velocity and minimum fluidization velocity. They defined an important parameter, the fluidization index (FI), it is a measure of the uniform expansion of the fluidized bed after bubbling. The fluidization index values range from 1 to 2 as a great value but in special cases it is more, the high values of FI means the bed capacity for holding gas is larger. It is an important parameter i.e., in fluidized bed reactor, for catalysts in industry means there is a plasticity and the bed can bent around the corners, contract and expand while the low value means the fluidization state is brittle where a small air velocity change cause a break from uniform fluidization state to a packed or bubbling bed state. FI can be expressed as:

$$FI = \frac{U_{mb}}{U_{mf}} \quad (4)$$

They predicted a dimensionless relation for U_{mb} in cylindrical columns for gas –solid systems at ambient conditions fluidized by air:

$$U_{mb} = 0.5231 \left(\frac{dp}{Di} \right)^{1.13} \left(\frac{Di}{h_s} \right)^{-0.0384} \left(\frac{\rho_s}{\rho_f} \right)^{0.74} \quad (5)$$

(Leu, and Tsai, 2009), studied the effect of static bed height for Plexiglas fluidized bed 1.8 m high and 10 cm bed inside diameter using type A material, they concluded that the U_{mb} was independent on static bed height. **(Escudero, et al., 2011)**, mentioned that for cylindrical bubbling fluidized beds, the U_{mb} was independent on static bed height. **(Panda, 2013)**, studied the U_{mf} , U_{mb} and FI for 3D tapered gas –solid fluidized bed with angles 7.36° and 9.28°, he used Limestone, Hematite and Dolomite with range of particle size 20-80 µm. He observed that they increase with increased static bed height. **(Gajbliye, et al., 2013)**, studied the hydrodynamics and stability of a fluidized bed with respect to minimum bubbling velocity, minimum fluidization velocity and fluidization quality, they used Geldart group B particles, they named the fluidization quality for U_{mb}/U_{mf} and it was equal to one for this type. **(Agu, et al., 2017)**, had an experimental study on samples of 150-450 µm limestone particles and two samples of 100-550 µm glass beads, they concluded that the



bubbling velocity is determined when the first bubble was observed in the top of the fluidized bed also they mentioned that for a bed of Geldart B type with particles size range 100-1000 μm , the bubbles appear as soon as the fluidization is started.

1.3 Minimum Slugging Velocity (U_{ms}) , Bubbling Bed Index (BBI) and Slug Index

(Singh , and Roy, 2008), defined the minimum slugging velocity as the gas velocity at which the size of bubble equals to the bed diameter, at this velocity the bed is lifted by the gas and this phenomenon is called slug formation. Slugging increases entrainment problems and lowers the bed performance. They called the difference between U_{ms} and U_{mb} as the bubbling regime. They developed a Correlation for U_{ms} prediction in cylindrical gas –solid fluidization systems. They used air at ambient temperature and non-spherical particles of diameter range 324-925 micron and density of 1500-4800 Kg/m^3 :

$$U_{ms} = 0.136 \left(\frac{dp}{Di} \right)^{0.6324} \left(\frac{Di}{h_s} \right)^{0.044} \left(\frac{\rho_s}{\rho_f} \right)^{0.6559} \quad (6)$$

They defined the Bubbling bed index (BBI) as the ratio of minimum slugging velocity to minimum bubbling velocity. BBI predicts the bubbling range for gas- solid fluidization system:

$$\text{Bubbling Bed Index} = \text{BBI} = \frac{U_{ms}}{U_{mb}} \quad (7)$$

(Ho, et al., 1983) ,used different beds of sand and glass particles (358-1112) μm , they observed that the U_{ms} is independent on the static bed height and column diameter. (Yang, 2003) ,considered that the slug appears in fluidized beds when the static bed height (H_s) /the diameter of the bed (D_i) is larger than 2 to ensure that the bubbles have the time to coalesce to larger bubbles and the slugging velocity is reached when the bubbles become 2/3 of the internal diameter of the bed. (Dimattia, et al., 2009) ,studied a bed of Geldart type D for coarse sand, 10mm glass beads, whole peens and long rice at different static bed heights, they mentioned that the U_{ms} was independent on static bed height because those particles offer a low resistance to gas flow and the shape of the particles has the important role on U_{ms} .(Kong, et al., 2017) ,investigated the slugging velocity in 0.05 m inside bed diameter for Geldart A particles type; they concluded that for fine particles when the diameter is less than 100 μm , the U_{ms} was independent on static bed height. (Agu, et al., 2017) ,used glass and limestone particles in a cylindrical column with 1.4m height and 10.4 cm internal diameter using air at ambient conditions as the fluidizing medium and static bed height of 52 ,58 and 64 cm ,they concluded that the transition from bubbling to slugging is related to bed height and the slug formation depends on aspect ratio ,which is defined as the ratio between bed height to column diameter, also they suggested that the U_{ms} decreases with an increase in bed height up to $h_{mf}=60 D_i^{0.175}$ after this it increases . (Jaiswal, et al., 2018),



studied 3D cylindrical bed of 0.084m diameter and 1.5m height, sand of mean particle diameter of 234.74 μ m was used .They used aspect ratio (H_s/D_i) of 0.7, 1.0, 1.5, 2.0 and 2.5 .They concluded that as the static bed height increased the slugging velocity decreased because the turbulence increased .Also the U_{mf} dropped to a stable value, U_{mb} remained constant and U_{ms} decreased with the increase of static bed height.

(Agu, et al., 2017),defined a slug index which indicates the ease of bed slugging, the low values means the bed can easily slug. Slug index slightly depends on bed height:

$$\text{Slug index} = \frac{U_{ms} - U_{mf}}{U_{mf}} \quad (8)$$

Bubbles and slugs flow influence the solid and gas interaction in fluidized beds, so identifying the velocities of bubbling and slugging are important in the study of the hydrodynamics behavior of fluidized columns.

The main objective of this work was to study the effect of column inside diameter and static bed height on the minimum fluidization velocity, minimum bubbling velocity, fluidization index, minimum slugging velocity, bubbling bed index and slug index and the experimental results were compared with the theoretical values by using the correlation equations.

2. MATERIALS and METHOD

2.1 Material and Chemicals

The material used as a bed was a pharmaceutical material, Paracetamol, its properties is shown in **Table1**. The density is 600 g /m³ and a mean particle size of 1.06*10⁻³ m. (Geldart, 1986) ,classified powders according to the relation between particle diameter and density difference between the solid and the fluidizing gas, **Fig.1**. The material is classified of type B according to Geldart classification and the bed behavior is identified according to this practical classification.

Table1. Particle and bed properties.

- Material Property	
Name	Paracetamol (Geldart type B)
Shape	spherical
Color	White
Size (dp)	1.06*10 ⁻³ m
Density (ρ_s)	600 kg/m ³
Density of fluidizing medium ,air (ρ_f)	1.18 kg/m ³ at 25°C
Column inside diameters (D_i)	0.0762, 0.15, 0.18 m
Static bed heights (h_s)	0.05, 0.07, 0.09 m

Temperature within $\pm 2^\circ\text{C}$



(Cocco, et al., 2014) mentioned that Geldart group B particles fluidize easily and are widely used in unit operations because they cause few difficulties. Most pyrolysis and fluidized bed combustors units use particles of this type. They form bubbles at the onset fluidization so the minimum fluidization velocity almost equals to the minimum bubbling velocity. Also they cleared that this type allows the formation of large bubbles size nearly equal to 2/3 of the internal column diameter and that cause slug formation.

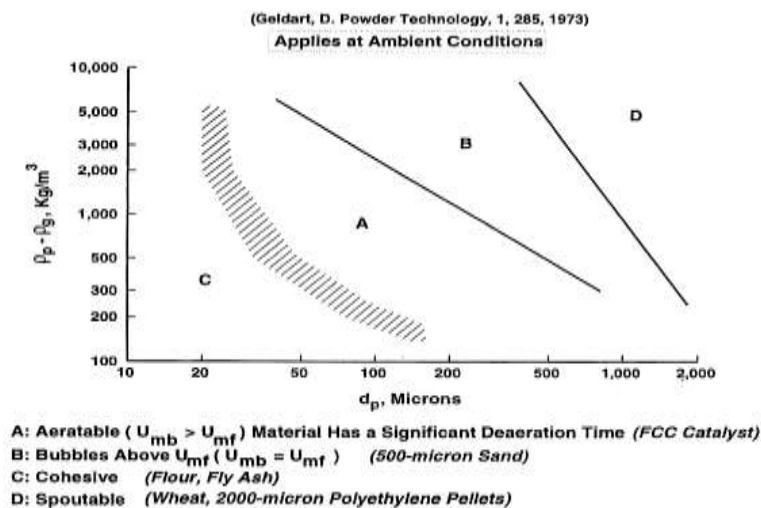


Figure1. Geldart classification.

2.2 Method:

The experimental set up is shown in Fig.2 and it consists of the following parts:

1- Air Compressor:

2- Silica gel column:

A column was filled with silica gel granules to adsorb any humidity exits in the air. The color of the silica was blue and became pink when it was saturated with moisture.

3- Calming section: A column was used to get a stable rotameters readings.

4- Rotameters: Two rotameters were used .They calibrated at 20°C and 0.981 bar abs. m^3/hr . air.

5- Manometer.

6-Fluidization column: It consists of a cylindrical column of 0.0762, 0.15 and 0.18 m inside diameters and 1.5 m height and made of Perspex material. A calming section with Rachig rings was used at the air entrance to the column for uniform distribution of the air throughout the bed. Air distributor was fitted at the column bottom.

A measured amount of the material was charged from the top of the fluidized bed ,the bed fluidized once and the particles allowed to settle down slowly, the desired static bed height was recorded (The bed was fluidized once before each experiment to get the particles in a loosely state).

Compressed air at ambient temperature was allowed to pass through the silica gel column, then through the calming column .The air passed through a porous plate distributor to the fluidization column and its velocity was increased until the minimum fluidization velocity was observed

experimentally and the height at minimum fluidization velocity was recorded. The air velocity was increased until the first bubble observed visually and the first bubble erupted from the bed free surface and the velocity recorded as the experimental minimum bubbling velocity then the air velocity was further increased till the slug formation started, the velocity recorded as experimental minimum slugging velocity. This procedure was repeated for each column diameter, three times for each bed height. Each experiment was repeated for several times and the average value was reported.



Figure2. Experimental set-up .

3. RESULTS and DISCUSSION

The experiments were carried out for measuring the minimum fluidization velocity, minimum bubbling velocity, fluidization index, particulate fluidization range, minimum slugging velocity, bubbling bed index and slug index. A pharmaceutical material “Paracetamol” was used. The variables were the column inside diameter 0.0762, 0.15 and 0.18m and static bed height 0.05, 0.07 and 0.09m. The results are shown in **Figs.3-14 and Tables 2 and 3.**

It can be noted from **Table2. and Figs. 3 and 4** that the U_{mf} increased by the increasing of the static bed height for the same column diameter and that agreed with **(Akil, et al., 2010) , (Ram, 2013) and (Panda,2013)**, this can be explained due to the drag force affecting on the particles during the process of fluidization. The friction of the wall opposes the weight of the bed so increasing the bed height means increasing the bed weight and that increases the friction enhancement which causes to increase the U_{mf} . Also the U_{mf} increased with



increasing column diameter for the same static bed height and that agreed with (Sarker, et al.,2012), this can be explained that by increasing column diameter and for the same bed height means larger amount of the material and that needs more force and larger velocity to raise the bed to reaches the minimum fluidization velocity. (Yupeng, et al., 2017) ,explained that the bed height increase cause to increase the wall effects: the boundary wall effect and the particles wall friction which are contributed, the second one effect mainly on the particles causing higher values of U_{mf} . Also the particle wall friction and column diameter are coupled and can not be separated, they are responsible for the U_{mf} variation.

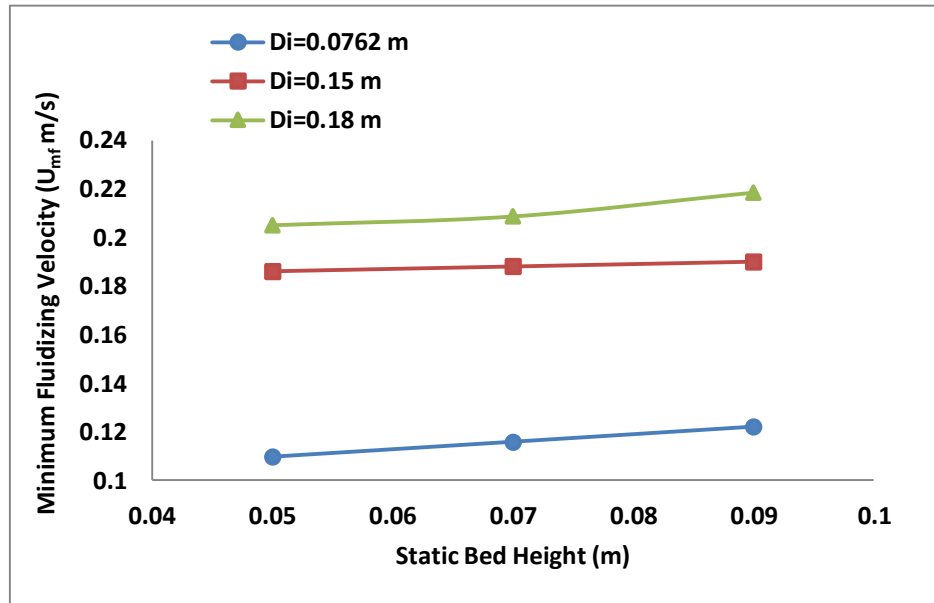


Figure 3. The relation between U_{mf} and static bed height for different column diameters.

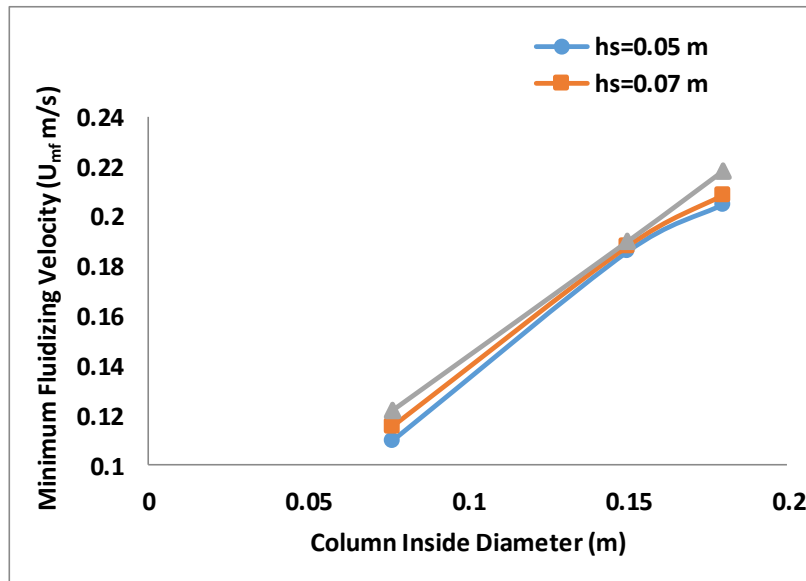


Figure 4. The relation between U_{mf} and column diameter for different static bed height.

**Table2.** Results and calculations of minimum fluidization, minimum bubbling and FI.

D_i (m)	h_s (m)	Q (m ³ /h)	G_{mf} (Kg/m ² .s)	U_{mf} (Exp) (m/s)	U_{mb} (Cal) Eq.5	U_{mb} (Exp)	FI (-) Eq.4	Particulate Fluidization Range (m/s)
0.0762	0.05	1.8	0.129	0.1097	0.41	0.11	1.0	0.004
	0.07	1.9	0.1366	0.1158	0.42	0.122	1.05	0.006
	0.09	2.0	0.144	0.12195	0.423	0.134	1.1	0.01
0.15	0.05	11.8	0.2195	0.186	0.187	0.187	1.01	0.001
	0.07	12.0	0.222	0.188	0.188	0.189	1.01	0.001
	0.09	12.1	0.224	0.19	0.191	0.192	1.01	0.002
0.18	0.05	18.8	0.242	0.205	0.15	0.206	1.01	0.001
	0.07	19.1	0.246	0.2086	0.152	0.209	1.0	0.004
	0.09	19.5	0.258	0.213	0.154	0.215	1.01	0.002

The gas superficial velocity was further increased until the first bubble was noticed at the surface of the fluidized bed and the velocity recorded as the experimental U_{mb} as shown in **Table 2.**, and **Figs. 5, 6 and 7**. The calculated values were obtained by using **Eq. (5)**. **Fig.5** shows a comparison between the experimental and calculated values of U_{mb} , it can be observed that for $D_i = 0.0762\text{m}$ there was no agreement between them, a very good agreement for $D_i = 0.15\text{m}$, and a fairly one for $D_i = 0.18\text{m}$. This diversion can be explained due to one of the assumptions in the correlation development was the particles are non- spherical. **(Singh, and Roy, 2005, 2008)** ,explained that the reason of the difference is due to the particle shape and its effect on drag and U_{mf} then on U_{mb} (There was not another correlation for sphere particles including the bed height and column diameter in the literatures to recalculate the U_{mb}). **(Singh, and Roy, 2005)** ,whom predicted the correlation equation comment that” It is always better to rely on experimental values than on the theoretical from the correlations”. From **Fig.6** and **Fig.7** it can be concluded that the U_{mb} increased by increasing the static bed height and column diameter and that agreed with **(Ram, 2013)** and **(Panda, 2013)**. The increase in U_{mf} and subsequently in U_{mb} can be explained due to the increase of frictional interaction between the particles and the wall, for the same static bed height and larger diameter that means bigger weight and that increases the drag forces and wall friction which cause to an increase in U_{mb} .

It was observed that directly after the U_{mf} observation, the bubbling velocity was noticed where many bubbles appeared above the distributor plate, they became larger and moved upwards through the bed. Also the values of U_{mf} and U_{mb} are approximately the same and that agreed with **(Agu, et al., 2017)** and **(Cocco, et al., 2014)** ,that the behavior of this kind of material was according to Geldart classification and can be considered of kind B. Also that agreed well with **(Jaiswal, et al., 2018)**, For shallow beds ($h_o/D_i < 2$) the bubbles



appear as soon as the bed fluidized where h_o/D_i for the three bed heights and column diameters equal to (0.66, 0.92, 1.2), (0.3, 0.47, 0.6) and (0.28, 0.39, 0.5) respectively.

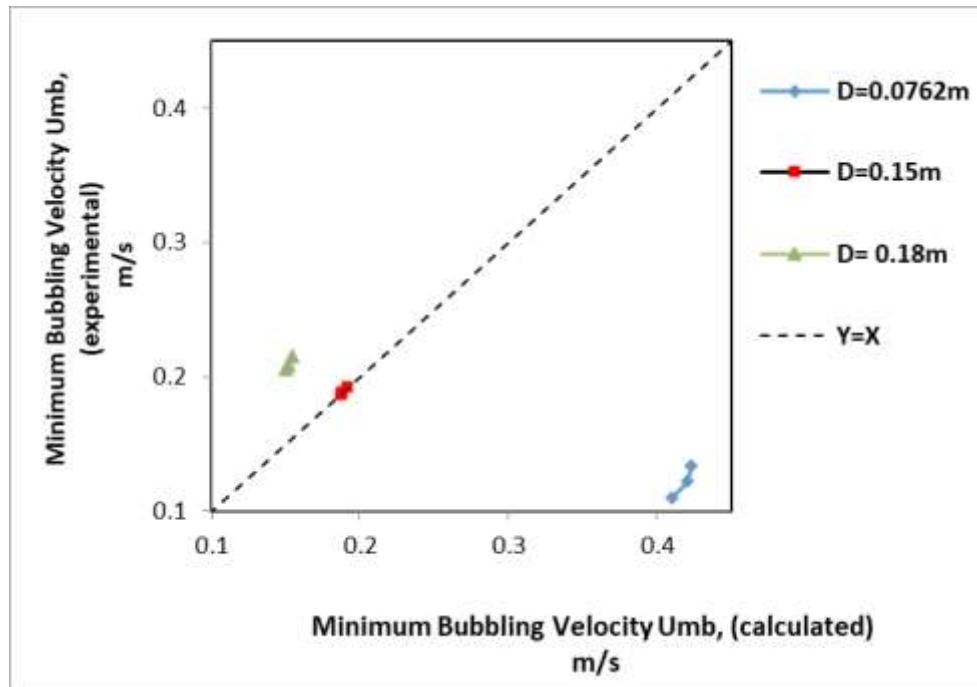


Figure.5 Comparison between experimental and calculated values of minimum bubbling velocity.

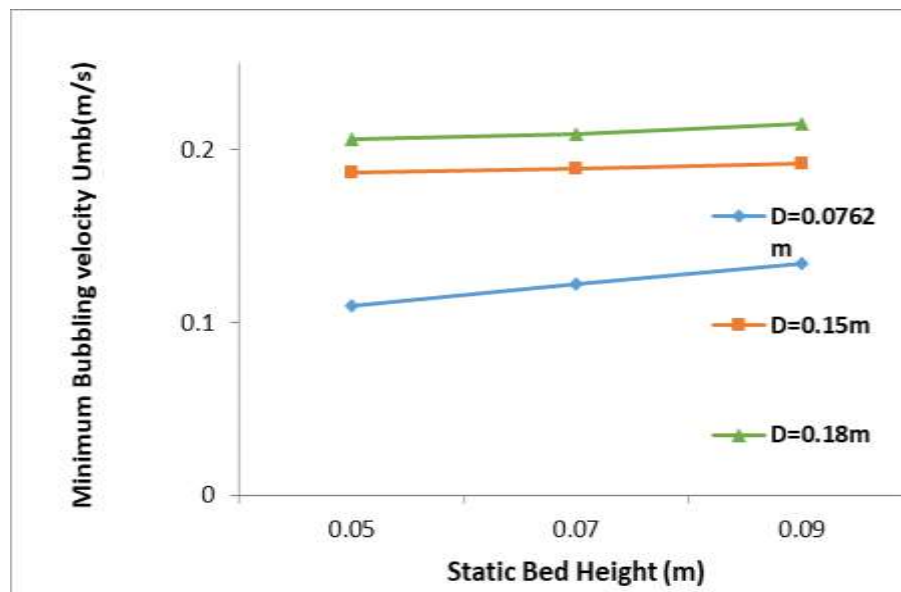


Figure.6 The relation between U_{mb} and static bed height for different column diameters .

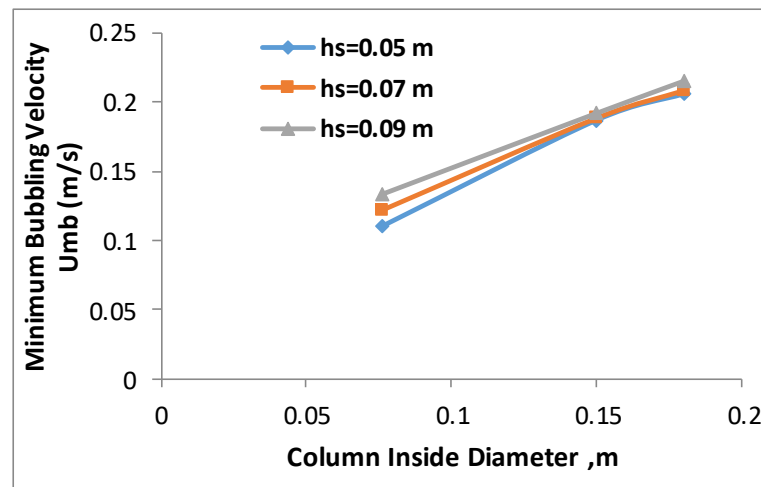


Figure.7 The relation between U_{mb} and column diameter for different static bed heights.

There was a short period of particulate fluidization ($U_{mb} - U_{mf}$) with a range of (0.01-0.004 m/s) shown in **Table 2.**, this period disappeared in bubbling fluidized beds where nearly the $U_{mb} = U_{mf}$. The fluidization index was calculated using **Eq. (4)** and the results are shown in **Table 2. and Figs. 8 and 9**, it is obvious that there was not a large dependence on static bed height and column diameter also the values were around 1, these values mean the bed has a less capacity to hold gases and a large possibility to form bubbles for a small increase in air velocity above the minimum fluidization velocity. It can be concluded that for materials of Geldart type B, the fluidization index was nearly unity and that agreed well with (**Gajbliye, et al., 2013**) and (**Ram, 2013**).

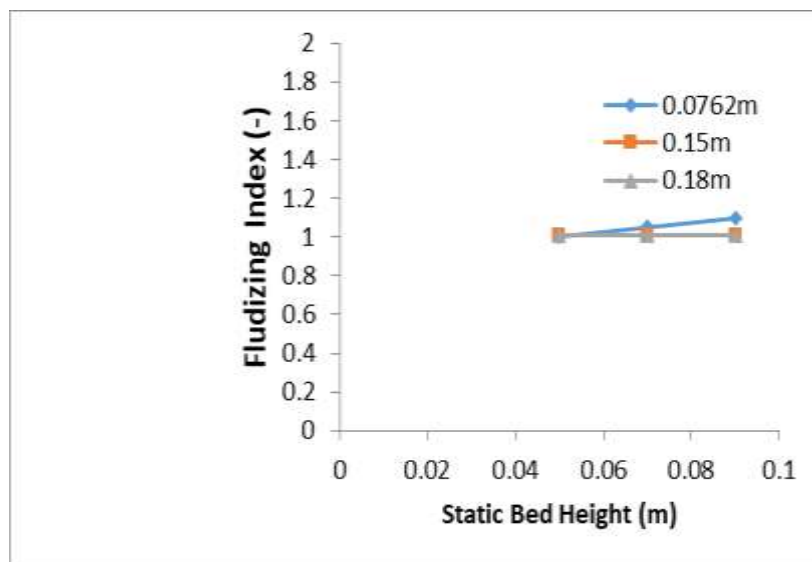


Figure 8. The relation between fluidizing index and static bed height for different column diameters.

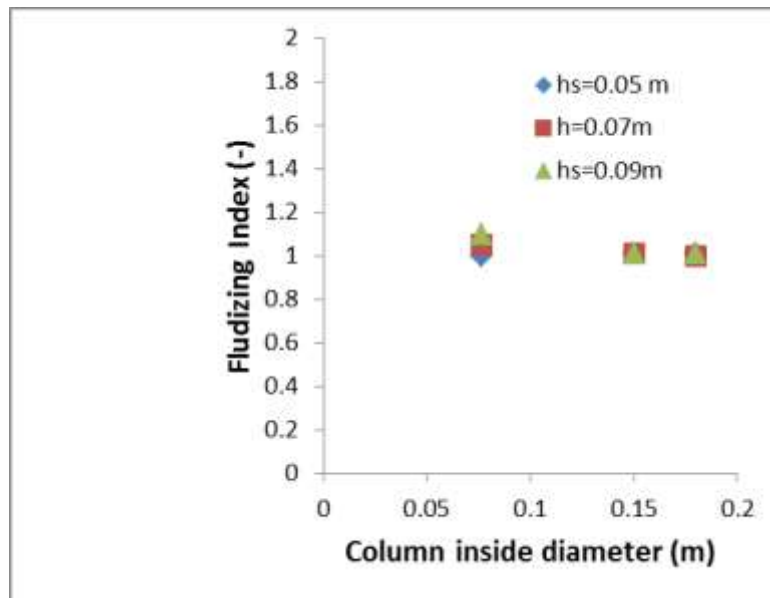


Figure 9. The relation between fluidizing index and column diameter for different static bed heights.

For further increase in gas velocity, slug started to form, it was noted visually with the structure of rounded- nosed, that agreed with (Agu, et al., 2017) , for easy fluidized materials, the material used was considered of this type, the slug had that structure and it can be noted visually for narrow (shallow) beds. The experimental and calculated values of U_{ms} are shown in **Table 3.** and **Figs.10, 11 and 12.**

Table 3 . Results and calculations of minimum slugging velocity, BBI and slug index.

D_i (m)	h_s (m)	G_{mf} (Kg/m ² .s)	U_{ms} (Cal.) (m/s) Eq.6	U_{ms} (Exp.) (m/s)	U_{ms} (Cal.) (m/s) Eq.9	BBI (-) Eq.7	Slug Index (-) Eq.8
0.0762	0.05	0.129	0.55	0.267	0.28	2.4	2.4
	0.07	0.1366	0.545	0.260	0.278	2.1	2.2
	0.09	0.144	0.539	0.257	0.272	1.9	2.1
0.15	0.05	0.2195	0.371	0.39	0.43	2.1	2.1
	0.07	0.222	0.365	0.376	0.42	2.0	2.0
	0.09	0.224	0.361	0.37	0.40	1.9	1.9
0.18	0.05	0.242	0.333	0.439	0.44	2.1	2.1
	0.07	0.246	0.328	0.435	0.432	2.1	2.1
	0.09	0.258	0.325	0.43	0.43	2.0	2.0

Fig.10 shows a comparison between the experimental and calculated values for U_{ms} , it can be concluded that for $Di= 0.0762m$ there was no agreement between them, a very good agreement for $Di=0.15m$, and a fairly one for $D=0.18m$. The calculated values were obtained from **Eq. (6)**. The difference can be explained due to an assumption used in the development of the correlation and that was the particles are non- spherical. **(Singh, and Roy,2005,2008)** ,explained that the reason of the difference is due to the particle shape and its effect on drag forces. **(Dimattia, et al., 2009)** ,concluded that the shape of the particles has an important role on U_{ms} . A good agreement between the experimental results and the calculated values of U_{ms} was obtained for the three column diameters and static bed heights when another equation have been used, **Eq. (9)**, to calculate the U_{ms} , this equation developed by **(Geldart, and Baeyens, 1974)**. The results are shown in **Table3.** and **Fig. 10 .** **(Geldart, and Baeyens, 1974)** had their equation for different fluidized beds : bed diameter 0.05-0.3 m , different densities 550-2800 Kg/m³ and the range of particle size 55-3380 μm . Their equation was widely used for U_{ms} prediction for different bed diameters, heights and particle properties. (The units are in centimeters)

$$U_{ms} = U_{mf} + 0.0016(60Di^{0.175} - h_{mf})^2 + 0.07(gDi)^{0.5} \quad (9)$$

From **Table 3.** and **Fig. 11**, it can be observed that U_{ms} had an inverse relation with static bed height and that agreed with **(Jaiswal, et al., 2018)** ,that when the static bed height increased, the slugging velocity decreased because the turbulence increased. According to **(Agu, et al. ,2017)** ,the values of $h_{mf}=60 Di^{0.175}$ were(86,96.4 and 99.5 cm) if the h_{mf} value exceeded those values for each individual column diameter ,the U_{ms} increased with bed height (The $60 Di^{0.175}$ value means the height of the fluidized bed at conditions of minimum fluidization for a flow of stable slug) ,experimentally h_{mf} for each diameter for the three bed heights were (6,7.5 and 10), (5.6, 7.5, 10) and(5.4, 7.3, 9.2) cm they did not exceeded the upper limit and that means there was a stable slug and the U_{ms} decreased with bed height . Also it can be noted that the lines were almost parallel to each other, so it can be concluded that the relation between the U_{ms} and static bed height is linear and that agreed with **(Agu, et al., 2017)**.

From **Table 3.** and **Fig.12** it can be observed that the U_{ms} had a direct relation with column diameter, this can be explained that bubbles first formed at the distributor then they rise and became larger so if the gas velocity increased the bubbles grow in size and become large enough to occupy nearly the whole diameter of the column then the slug starts to form. By increasing the column diameter and for the same height which means larger weight of the material, the wall friction increases and that increases the resistance to air flow and cause to an increase in U_{ms} .

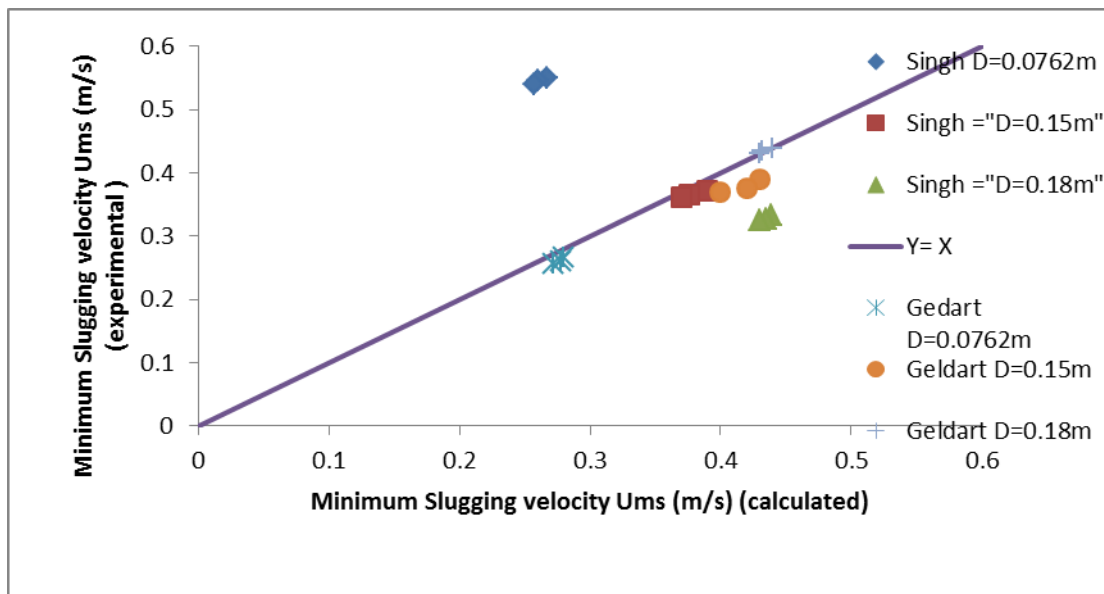


Figure.10 Comparison between the experimental and calculated values of U_{ms} .

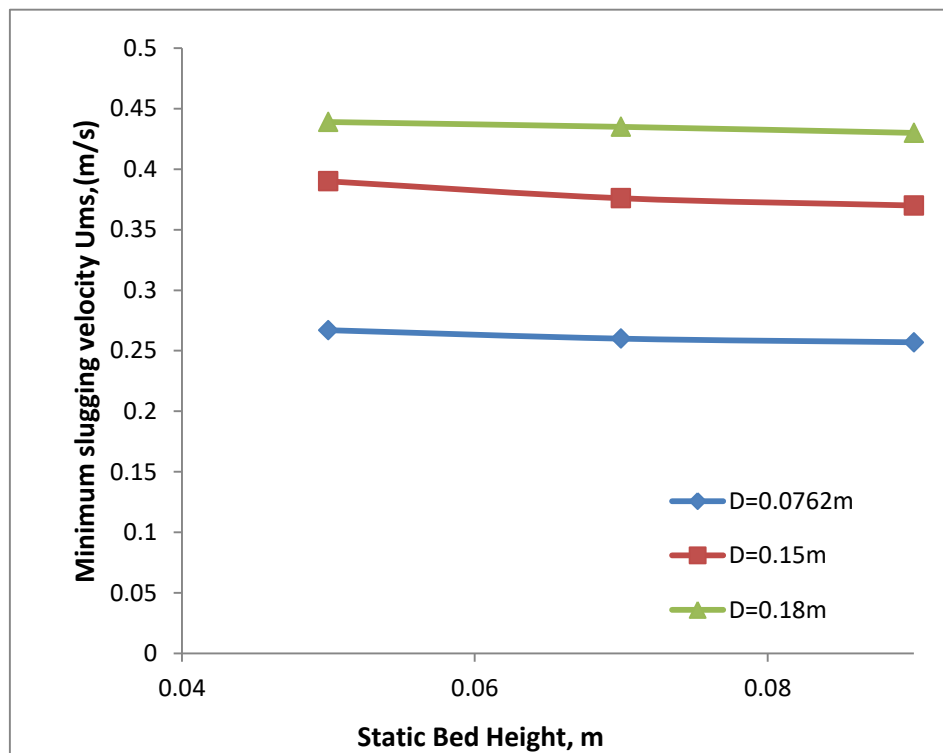


Figure .11 The relation between U_{ms} and static bed height for different column diameters.

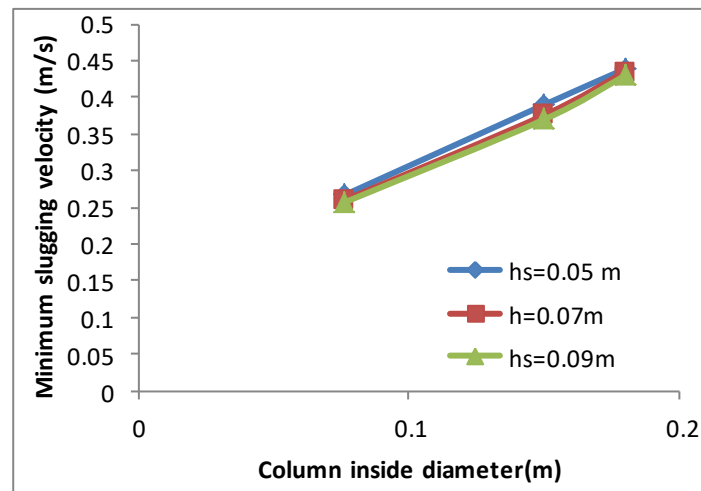


Figure.12 The relation between U_{ms} and column diameter for different static bed heights.

From **Table 3**, it can be observed that the bubbling bed index (BBI) calculated from **Eq. (7)** had the range of 1.9-2.4 and it decreased with increasing static bed height and there was a slightly dependence on it because by increasing the bed height, the minimum bubbling velocity slightly increased, so according to **Eq. (7)** the BBI decreased. The slug index calculated by using **Eq. (8)** and From **Table 3**. and **Figs. 13 and 14** the slug index also decreased with increased bed height and they had nearly the same values 1.9-2.4 and that agreed well with (**Agu, et al. , 2017**). Both of them decreased with column diameter but there were not a significant effect.

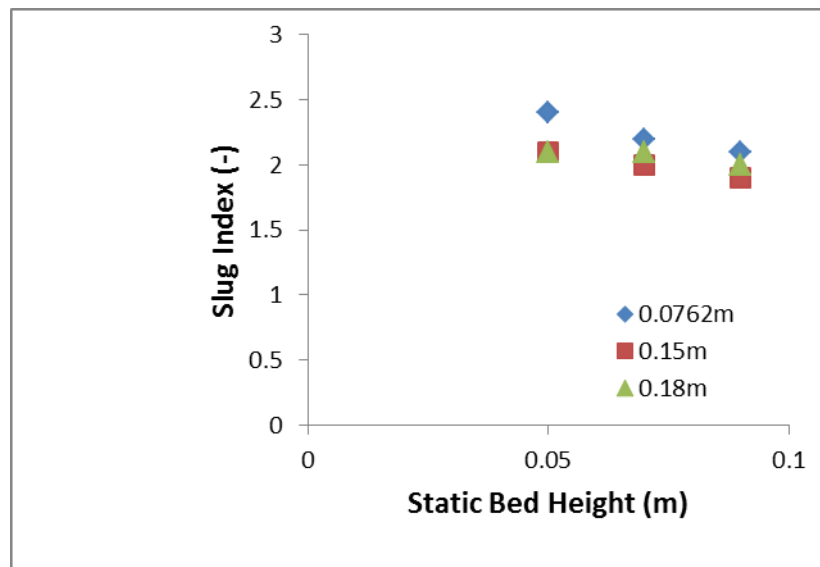


Figure. 13 The relation between slug index and static bed height for different column diameters.

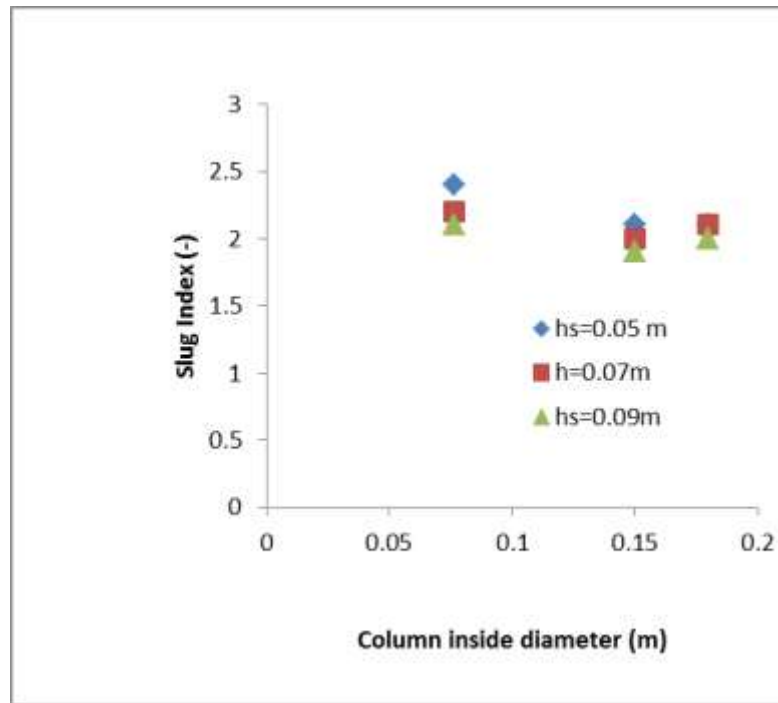


Figure.14 The relation between slug index and column diameter for different static bed heights.

4. CONCLUSIONS

- The minimum fluidization velocity nearly equals to the minimum bubbling velocity for this kind of material, the material considered of type B according to Geldart classification of particles. U_{mf} has a direct relation with column diameter and static bed height.
- The minimum bubbling velocity increased with an increase in static bed height and column diameter.
- The fluidization index values were around 1 in all cases and that ensures the Paracetamol material is of Geldart type B.
- The minimum slugging velocity had an inverse relation with static bed height and a direct one with column diameter.
- The bubbling bed index and the slug index slightly depend on static bed height and there was not a significant effect for both of them with column diameter change.

5. NOMENCLATURE

BBI= bubbling bed index, dimensionless

D_i = column inside diameter, m

d_p = mean particle diameter, m



FI = fluidization index, dimensionless

G_f = air mass velocity, Kg/m².s

G_{mf} = air mass velocity at minimum fluidization velocity, Kg/m².s

h_{mf} = bed height at minimum fluidization velocity, m

h_s = static bed height, m

U_{mf} = minimum fluidization velocity, m/s

U_f = air Superficial velocity, m/s

U_{mb} = minimum bubbling velocity, m/s

U_{ms} = minimum slugging velocity, m/s

Re= Reynolds Number, dimensionless

ρ_s = density of solid particles, Kg/ m³

ρ_f = density of solid particles, Kg/ m³

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