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# EXPERIMENTAL MEASUREMENT OF DIFFERENT FLUIDIZATION PARAMETERS

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**Abstract**

The minimum fluidization velocity and height are the important hydrodynamic parameters of fluidized bed which required primary data for mathematically modeling of fluidized bed reactor. Minimum fluidized velocity for particles can be obtained through existing theoretical and experimental equations, but the minimum bed height can be only achieved via experiments. The minimum fluidized velocity is a function of physical properties of particles, gas and porosity of bed($\varepsilon mf$) in the minimum fluidized velocity. The minimum fluidized height is a function of catalyst weight, bed diameter, gas distributor, and fluidized bed hydrodynamic. In order to run the experiments, an experimental system including glass column (fluidized bed) with gas distributor and manometer is made. The dropping pressure in fluidized bed, gas distributor with variation of particle height in the column is measured versa entering gas velocity. Diagram of dropped pressure versus velocity ($\Delta$P-U) is drawn in the log-log and linear scales, and where direction of the curve is changed the minimum of fluidized velocity will be obtained. Using the curve of catalyst height versus velocity (H-U), the minimum fluidized height (Hmf) is obtained where is the minimum of fluidized velocity (Umf).

***Key Words**:* Minimum Fluidization Velocity; Minimum Fluidized Height; Fluidized Bed; Hydrodynamic; Minimum Fluidized Porosity.

## 1. Introduction

Literature survey in fluidization area and minimum fluidization measurement shows that in searched literatures measurement method for minimum fluidization velocity and height were not reviewed also researcher in this area will nor review of all applicable minimum fluidization velocity in one point . This manuscript will review best equation for calculating of minimum fluidization and compare it with experimental data.

The contribution of this manuscript will be summarized as following:

- A reference manuscript for find all equation for minimum fluidization velocity calculating
- Predicting a experimental method for measurement minimum fluidization velocity
- Predicting different method for measurement of minimum voidage
- Predicting experimental method for measurement minimum fluidization height

### 1.1. Fluidization phenomena [4,5,9]

Whenever a fluid is entered a packed column with a low velocity, the occluded particles in fluid do not have any movement then there is a fixed bed condition. When the fluid velocity is going up, the particles take the distance from each other and many of them start rotating in a limited space. In this case, the bed will be expanded to provide an *Expanded Bed* and the porosity and pressure drop will be increased versa *Fixed Bed*. Beginning this status, it starts moving from point **A** to reach the point B on the curve

Figure 3, bed still is in Loasent status in this situation and there is still the contact among the solid particles. With increasing the velocity of entering fluid, all solid particles in gas and liquid phases are floating up. At this situation, friction force between solid particles and fluid is equal with particles' weight force, vertical vector of existing pressure force among adjacent particles are eliminated and pressure drop of every sections of bed is almost equal with weight of fluid and occluded particles in that fluid section. In this case, the bed is fluidized and named primary fluidized bed or minimum fluidization state. In liquid-solid systems, increasing of fluid flow at higher than minimum fluidization state terminates to a gently expanding state, but permanently bed is developing. Instability of flow pretends toward a limited expansion and will be reached a constant value at a fixed quantity, and generally, large bubbles and heterogeneity are not observed. The bed with these characteristics is called particle fluidized bed; homogeneous fluidized bed, permanent fluidized bed, or smoothly fluidized bed.

Behavior in the gas-liquid systems is often different from above behavior. With increasing the velocity of fluid after the minimum fluidization, high instability is observed like forming bubbles and channeling of gas flow. At even higher velocity, highly disturbance is observed and the movement of solid particles are more complicated. Generally, the expanding of bed concerning of its volume is not noticeable at minimum of fluidization. Such a this bed is named comprehensive fluidized bed, heterogeneous fluidized bed, bubbling fluidized bed, or more simple, gas fluidized bed. In the recent situation, pressure drop is decreased a little bit until reaches to the point **C** on the curve. The velocity of fluid in the bed and among particles are more than its upper air space and as a result, all particles are flowed towards outside of tube, but only very small particles are carried by fluid in this position (Entrainment). With increasing velocity of fluid at point **D**, all particles in the bed are carried by fluid, so that its porosity is shifted to 1. This phenomenon causes generating two phases and the primary fluidity is called bath fluidization because the particles out of the bed are not carried by fluid, but after that fluidization is named continuous fluidization.

## 2. Explanation

### 2.1. Minimum velocity

It is common knowledge that the minimum fluidization velocity is the basic information required for the design and development of fluidized bed processes; however in industrial practice fluidized bed reactors are mostly operated at superficial gas velocities well above the minimum fluidization velocities and, therefore, the minimum fluidization velocity is not a quantity with a precise significance for industrial applications. From the point of view of engineering practice even inaccuracies of up to 40% in the prediction of the minimum fluidization velocity values are more or less acceptable (Molerus and Wirth, [16]. Despite that and many earlier works, discussion on accurate prediction of the minimum fluidization velocity still seems to remain of much scientific interest. The best method to determine the minimum fluidization velocity would be by measurement. In practice, however, this is not always possible, especially at different operating conditions, such as at elevated pressure. In the absence of the facility to carry out experiments to determine the minimum fluidization velocity, approximate computation of this is necessary for any design or study of the fluidized bed process and numerous correlations have been proposed in the literature [12] .

The generally accepted approach is to estimate the effect of pressure on minimum fluidization velocity by employing Ergun [17] equation for pressure loss through a packed bed, which in slightly modified form as Equation (1) appears in many books on fluidization; or its numerous simplified variations, such as the most popular Wen and Yu [12,13 ] correlation Yang [15] noted that in the design of fluidized bed systems designers encounter two completely different situations: the area of interest has very little information or it has many correlations available and their predictions give sometimes very different results. There are numerous studies and proposed correlations on the prediction of the minimum fluidization velocity at ambient conditions as well as a number of comprehensive reviews comparing these correlations (Adanez and Abanades [1]; Couderc [3], Grewal and Saxena [8]; Lippens and Mulder [10]). The great majority of the correlations proposed in the literature are based on the Ergun equation, modified after an experimental evaluation based on limited numbers of data and materials, and quite often they are purely empirical. Those empirical correlations have limited value and according to Lippens and Mulder [10 ], preference should be given to complete characterization of the fluidized bed including determination of particle shape factor φ and voidage at minimum fluidization ε*mf* as described by Geldart [7]. Lippens and Mulder [10] explain the widely accepted success of the Wen and Yu [13,14] correlation in engineering practice as a first approximation because it simply offers the correct order of magnitude.

### 2.2 Pressure drop and minimum velocity at fluidized bed [11, 9]

Many researchers (Blake [18], Carman [19], Kozeny[20] and Ergun [17]) have presented equations for pressure drop at fluidized bed. Ergun's equation is applied in a Wide range of Reynold's number for non-sphere particles. Ergun [17] presented that pressure drop at bed is equal total energy lost because of viscosity and kinetic effects.

For low Reynold's number (small particles and high temperature), losing of viscosity energy is more than losing of kinetic energy and total pressure drop is explainable using the losing of viscosity energy.

For high Reynold's number (large particles), the kinetic energy is important and total pressure drop is explainable using the losing kinetic energy. Ergun's equation is presented as the following:

$$\Delta P/L = 150\,(1-\varepsilon)^2/\Phi_s^2\,(\varepsilon^3)\,.\,\mu U/d_p^{\,2} + 1.75\,(1-\varepsilon)\,U^2\rho/\,\varepsilon^3\,\Phi_s\,d_p \qquad (1)$$

As it is expressed previously that the solid or gas is flowing up where fluidized action happens, the following equation is applied:Particle weight = the Drag force that is applied by moving gas or, it can be written as following:

(solid specific gravidity) X (the fraction of solid at bed) X (bed's volume) = (cross sectional of column) X (pressure drop at bed)

As a result of that, the following equation can be written:

$$\Delta P_b \,X\, A_t = (A_t \,X\, H_{mf})\,(1-\varepsilon_{mf})\,(\rho_s - \rho_g)\,g/g_c$$

The equation of pressure drop at bed:

$$\Delta P_{b/}\,H_{mf} = (1-\varepsilon_{mf})\,(\rho_s - \rho_g)\,g/g_c \qquad (2)$$

Using the equation of pressure drop at bed (Ergun's Equation) where fluidized velocity is minimum, pressure drop can be obtained by the following equation:

$$1.75\,K_1\,Re_{mf}^{\,2} + 150\,K_2\,Re_{mf} = Ar \qquad (3)$$

$$K_1 = (1/\Phi_s\,\varepsilon_{mf}^{\,2})\ ,\ K_2 = (1 - \varepsilon_{mf}\,/\Phi_s^2\,\varepsilon_{mf}^{\,3})$$

$$Re_{mf} = U_{mf}.\,d_p.\,\rho_g\,/\,\mu\ ,\ Ar = d_p^3.\,\rho_g\,(\rho_s - \rho_g)\,g\,/\,\mu 2 \qquad (4)$$

Wen and Yu [13,14] are the first researchers that the quantities of $K_1$ and $K_2$ at the Reynold's number range of 0.001 to 4000; while, suggested $U_{mf}$ has the standard deviation +/- 34%. The Table 1 shows the values of $K_1$ and $K_2$ that are obtained by the other researchers.

Chitester *et. al.* [26 have presented the following equation for large particles:

$Re_{mf} = [(28.7)^2 + 0.0494\ Ar]^{1/2} - 28.7$ (5)

The following equation is also presented for small particles (Wen and Yu [13,14]:

$1.75/\varepsilon_{mf}^{\ 3}\ \Phi_s(d_p\ .\ U_{mf}\ \rho_{g/}\ \mu)^2 + 150(1-\varepsilon_{mf})/\varepsilon_{mf}^{\ 3}\ \Phi_s^{\ 2}(d_p.U_{mf}\ \rho_{g/}\ \mu) = (d_p^{\ 3}\rho_g\ (\rho_s-\rho_g)\ g\ _{/}\mu)$ (6)

$Rep_{mf} = [(33.7)^2 = 0.0408\ Ar]^{1/2} - 33.7$ (7)

Wen and Yu [13,14] have also presented the following equation for small particles:

$U_{mf} = d_p^{\ 2}\ (\rho_s - \rho_g)g/150\ \mu\ X\ \varepsilon_{mf}^{\ 3}\Phi_s^{\ 2}/1 - \varepsilon_{mf}$ $\quad Rep_{mf} < 20$ (8)

For large particles the equation is:

$U_{mf} = dp^2\ (\rho_s - \rho_g)g/1.75\ \rho_g\ \varepsilon_{mf}^{\ 3}\Phi_s$ $\quad Rep_{mf} > 1000$ (9)

Using equation (3) it can be written as following:

At the equation (3) for low Reinhold's numbers, the kinetic term is ignored and then the following equation is obtained:

$150\ C_2\ Rep_{mf} = Ar$ or $\mu\ U_{mf} = Constant$ (10)

According to above equation with increasing the gas temperature, viscosity decreases and $U_{mf}$ enhances for high Reinhold's number. Then viscosity term is ignored and the following equations are obtained:

$1.75\ C_1\ Rep_{mf}^{\ 2} = Ar$ or $\rho\ U_{mf}^{\ 2} = Constant$ (11)

Table 1 All existing equations, which are used to calculate ($U_{mf}$) [9]

| Type of equation | Theoretical an experimental equation, Umf(cm/sec) | |
|---|---|---|
| Ergun, 1952[17] | $1.75/\ \varepsilon_{mf}^{\ 3}\ \Phi_s\ Rep_{mf} + 150\ [(1 - \varepsilon_{mf})/\ \varepsilon_{mf}^{\ 3}\ \Phi s^2]Rep_{mf} = Ar$ | |
| | $Rep_{mf} = (d_p\ .\ U_{mf}\ \rho_{g/}\ \mu)$ , $Ar = dp^3\ (\rho_s - \rho_g)g/\ \mu^2$ | |
| | For small partides: $Rep_{mf} = [(33.7)^2 + 0.0408\ Ar]^{1/2} - 33.7$ | |
| | For $U_{mf} = dp^2\ (\rho_s - \rho_g)g/1650\ \mu$ | |
| | For Rep >1000, $U_{mf} = dp^2\ (\rho_s - \rho_g)g/24.5\ \rho_g$ | |
| General equation | $K_1Rep_{mf}^{\ 2} + K_2Rep_{mf} = Ar$ | |
| | $K_1$ | $K_2$ |
| Wen and Yu [13] | 24.5 | 1651.96 |
| Richardson [21] | 27.39 | 1408.21 |
| Saxena and Vogel [22] | 17.5131 | 886.164 |
| Babu *et.al.* [23] | 15.36 | 777.265 |
| Grace [24] | 24.509 | 1333.33 |
| Chitester *et al.* [25] | 20.242 | 1161.94 |
| Chitester *et.al.* [26] | For large particles: $Re_{mf}=[(28.7)^2+0.0494\ Ar]^{1/2}-28.7$ | |
| Goroshko *et.al.* [27] | $Re_{mf}=Ar/(150(1-\varepsilon_{mf}/\varepsilon_{mf}^3) + 1.754Ar/\varepsilon_{mf}^3))$ | |
| McKay and McLain [11] | $400\ [(1 - \varepsilon_{mf})/\ d_p]U_{mf}\ \mu\ + 2.4\ \rho_g\ U_{mf}^{\ 2} = (\rho_s - \rho_g)g\ \varepsilon_{mf}^{\ 3}\ d_p$ | |
| Leva [28] | $U_{mf} = 0.0093\ d_p^{1.82}\ (\rho_s - \rho_g)^{0.94}/\ \mu^{0.88}\ \rho_g^{\ 0.06}$ (SI) | |
| Davidson, Harrison .[29] | $U_{mf} = 0.0008\ g\ (\rho_s - \rho_g)\ d_p^{\ 2}/\ \mu$ (SI) | |
| Rowe and Yacono [30] | $U_{mf}=0.0085[\varepsilon_{mf}^3/(1-\varepsilon_{mf})^{0.973}]^{0.973}d_p^{1.92}d_p(\mu m); U_{mf}(cm/s)$ | |
| Yacono [31] | $U_{mf}=0.000512\ d_p^{1.77}U_{mf}(cm/s);\ d_p(\mu m)$ | |

At equation (11), gas density decreases when the temperature increases and $U_{mf}$ also decreases. For small particles and large particles, terms of viscosity and kinetic are important. Table 1 presents all existing equations in literature which are used to calculate fluidity.

## 3. Experimental work

### 3.1.Manufacturing a set-up of glass bed for measurement of $U_{mf}$ and $H_{mf}$ [6]

A glass fluidized bed is made to measure $U_{mf}$ and $H_{mf}$ set up is including:

a) A glass column is used with internal diameter 4.8 cm and the glass column has two parts. Its top part, is a 100 cm distributor including one outer pathway and two other pathways located at both sides of distributor to measure the pressure drop. Down part of distributor has one pathway for entering gas and another pathway to measure pressure drop and the height of upper part is varied from 13 to 20 Cm. Because three different sizes of distributors of porous type are used, each of columns has a specific size.

b) The distributor of porous type which is used at the lower of column is separate-able. Three different sizes of porous plates are used with 1, 2 and 3 meshes. The mesh 1 is smaller than meshes of 2 and 3.

c) Other equipment which are used are a water manometer, a gas flowmeter, the tygon hoses and a metal frame. The water manometer and the gas flowmeter are used to measure pressure drop and gas flow respectively. The gas flowmeter is calibrated using calibrating equipment with nitrogen gas. The tygon hoses are used to measure pressure. All equipment is installed on the metal frame. Glass fluidized bed system is shown at Figure 1.

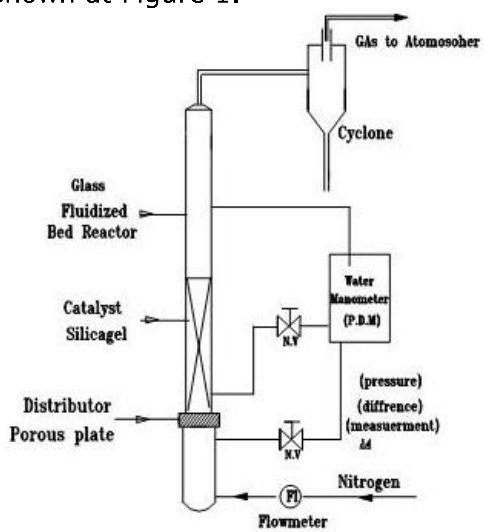


Fig.1 Glass fluidized reactor set-up for measuring Umf and Hmf of silicagel catalyst

### 3.2. The distributing size of silicagel particles (averaged diameter of particles), physical properties of particles and nitrogen gas [6]

Table 2 Distributing size of particles of silicagel catalyst coupled with physical properties of catalyst and physical properties of gas.

| Particle size per micrometer | Percent of particles | Particles density, (g/cm3) $\rho_s$ | 1.96 |
|---|---|---|---|
| 500 | 0.7 | Bulk density, (g/cm3) $\rho_B$ | 0.43 |
| 300 | 43.96 | Temperature of nitrogen Gas(T), $^oC$ | 27 |
| 250 | 18.55 | Nitrogen gas pressure(P-atm) | 1 |
| 180 | 24.32 | Gas viscosity(kg/m.sec) $\mu$ | 0.00001784 |
| 150 | 6.56 | Gas Density(Kg/m3) $\rho_g$ | 1.1421 |
| 125 | 3.45 | | |
| 100 | 2.47 | | |

The averaged diameter of particles is calculated from the following equation:

$d_{pAverage} = 1/\sum_i (X/d_p)_i = 1/4.697 \times 10^{-6} = 212.88\ \mu m = 0.21288\ mm$

### 3.3. Calculation of $U_{mf}$ and $H_{mf}$ via experiments

400 grams of silica gel powder catalyst (silicagel is the catalyst of conversion urea to melamine) is weighed and transfer into a column equipped with porous distributor with porous mesh 1. Nitrogen gas is flown with flow rate in which flow meter showed the numbers higher than 150 mm (Flow = 50 $cm^3$/sec.) then the pressure drop in the bed and total are measured (pressure drop is obtained using variation of pressure drop of bed and total). The bed height at fixed bed position is chosen zero as a baselines and the variation of catalyst height with gas flow rate is measured (using of length scale on the column). The value of pressure drop is measured from high velocity to low velocity and these amounts have a little variation from measuring pressure drop from low velocity to high velocity because of the cohesive force among particles. For distributor with mesh size 1, the variations of total pressure drop, bed pressure drop, and distributor pressure drop versus gas flow velocities are drawn. The experiment for distributors with porosities of mesh sizes 2 and 3 are also drawn. The situation of fluidized bed is studied in all above cases (fixed bed, bubbled fluidized bed, slugging phenomena and carrying particles at high velocity).

The experiments indicate that with decreasing the porosity size of distributor, the condition of distributing of gas flow will become better and the amount of distributing pressure drop will be increased as Figures presented. The porosity size decreases with enhancing number of distributors. The height of 400grams catalyst in column is measured which is 50 cm and based on, the variation of total height versus gas flow velocity is drawn. The experiments indicate which the variation of catalyst height ($\Delta H$) increases with decreasing porosity size of distributor. Fluidized bed has better condition with high catalyst height variations rather than low catalyst height variations. Diagram of pressure drop ($\Delta$ P) is drawn versus gas velocity (U) in a logarithm scale. Consider that $U_{mf}$ is velocity that the curve of P-U has been broken and the value of pressure drop from its maximum reaches to a fixed value. The value of Umf is a function of the type of distributor and obtained values are not identical in all states. Considering that distributor is at the best state with the mesh size 3 and provides properly gas distribution at the bed, the value in which third state is acceptable. $U_{mf}$ (3) is located between $U_{mf}$ (1) and $U_{mf}$ (2).

$U_{mf}$ (1) = 3.9cm/sec, $U_{mf}$ (2) = 2.4 cm/sec and $U_{mf}$ (3) = 2.9 cm/sec

These values are obtained using the diagram of height versus velocity (H-U). The height of catalyst ($H_{mf}$) which obtained has the following values for 400 gram of catalyst and are valid only for a reactor with internal diameter 4.8 cm.

$U = U_{mf}(3) = 2.9$ cm/s $\rightarrow H = H_{mf}(3) = 54$

Related Figures of the experiments are drawn as figures 2,3,4, 5,6,7,8,9 and 10. Pressure drop distributor and pressure drop of bed have the following relationship:

$$\Delta P_{(Distributor)} = \Delta P_{(bed)} \times (0.2 \text{ to } 0.4) \qquad (12)$$

Therefore the averaged distributor pressure drop is 30% of bed pressure drop where distributor with porosity of type 3 is used. All experiments result were shown at following figures:

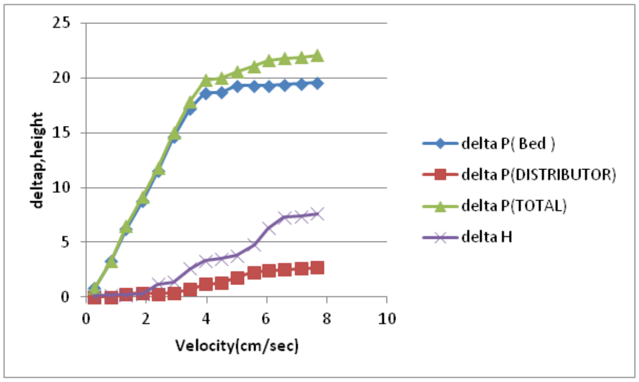


Fig. 2 Pressure drop and different height of catalyst vs gas velocity were measured.(for distributor mesh size1)-$d_r$=4.8cm,w=400gr, $h^{-1}$ 20cm

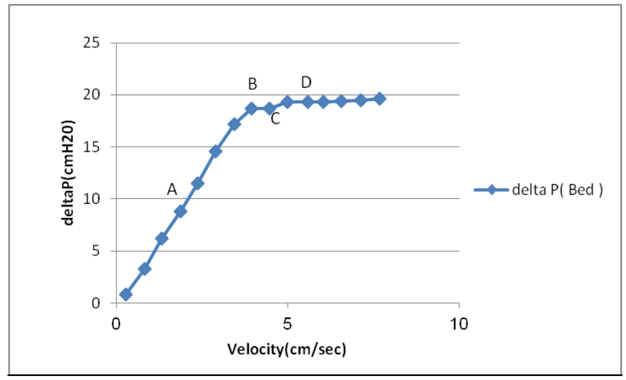


Fig. 3 Bed pressure drop vs gas velocity for distributor mesh size1

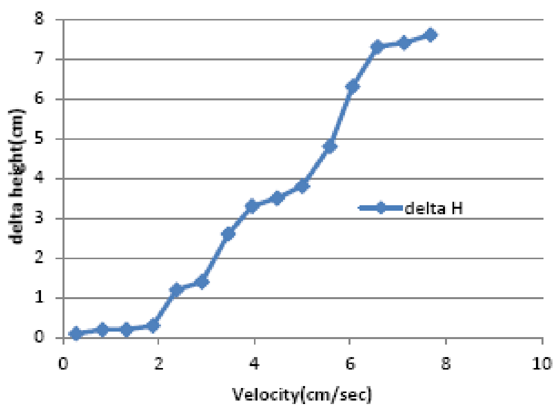


Fig. 4 Different height of catalyst vs gas velocity for distributor mesh size1

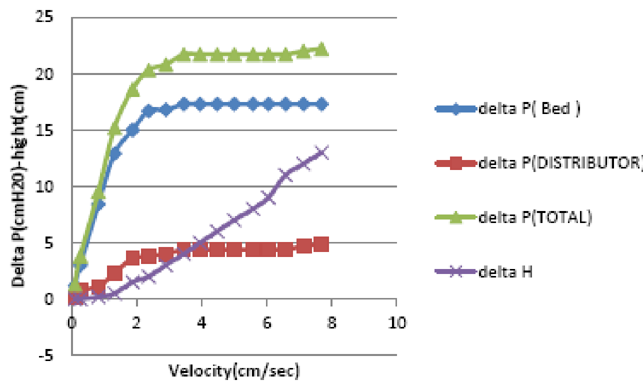


Fig. 5 Pressure drop and different height of catalyst vs gas velocity were measured.(for distributor mesh size2)-$d_r$=4.8cm,w=400gr,$h^{-1}$ 20cm

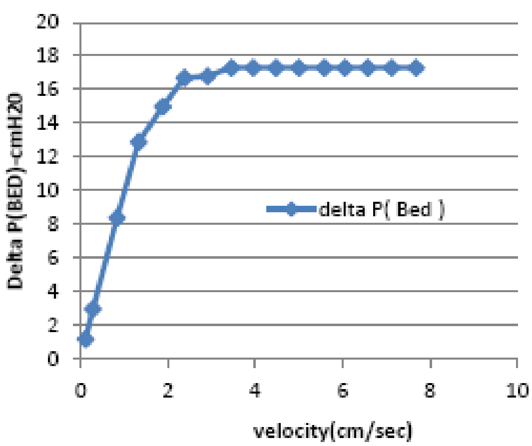


Fig. 6 Bed pressure drop vs gas velocity for distributor mesh size2

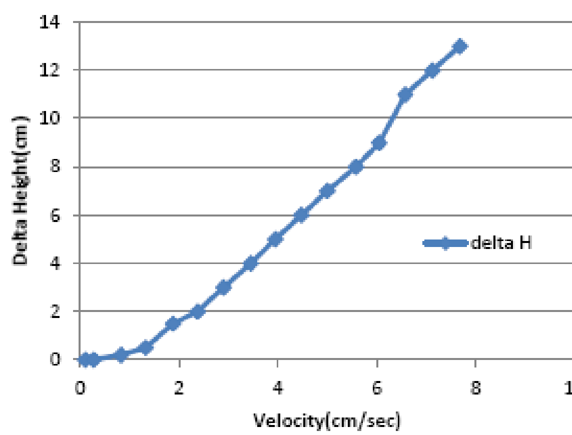


Fig. 7 Different height of catalyst vs gas velocity for distributor mesh size2

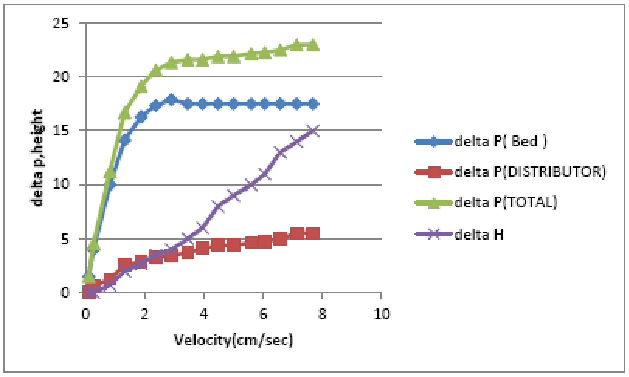


Fig. 8 Pressure drop and different height of catalyst vs gas velocity were measured.(for distributor mesh size3)-$d_r$=4.8cm,w=400gr,$h^{-1}$ 20cm

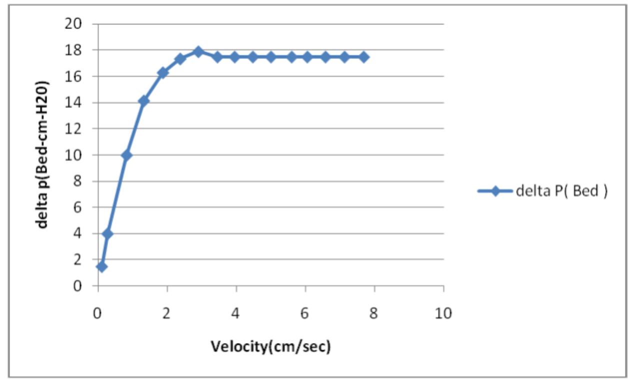


Fig. 9 Bed pressure drop vs gas velocity for distributor mesh size3

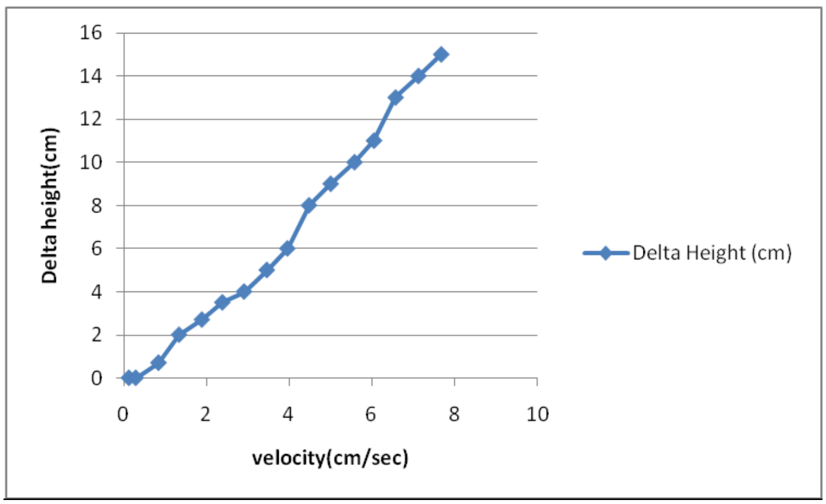


Fig. 10 Different height of catalyst vs gas velocity for distributor mesh size

## 4. Calculation of Umf Via the Equations and The Comparison of Obtained Values with Experimental Values [4,5,9]

Table 3 Calculation $U_{mf}$ according different equation is given.

| Type of Equation | $U_{mf}$(cm/sec) |
|---|---|
| Wen and Yu [13] | 2.915 |
| Richardson [21] | 3.4089 |
| Saxena and Vogel [22] | 5.388 |
| Babu *et.al.* [23] | 6.13 |
| Grace [24] | 3.6 |
| Chitester *et.al.*[25] | 4.1284 |
| Gorosho *et.al.* [27] | 14.3095, $\varepsilon_{mf}$=0.4151 |
| Ergun [17] | 3.9443, $\varepsilon_{mf}$=0.41516 |
| Leva [28] | 3.6509 |
| Davidson and Harrison [29] | 3.8982 |
| Rowe and Yacono [30] | 3.0604, $\varepsilon_{mf}$=0.4149 |
| Yacono [31] | 6. 7574 |
| McKay and McLain [11] | 1.493 |

## 5. The Values of $\varepsilon_{mf}$ at Fluidized Bed [2,9]

In order to calculate the value of $\varepsilon_{mf}$ , there are two following methods: The first method is the use of equation of Broadhurst and Becker and second method is based on the amounts of $K_1$ and $K_2$ in the equation which presented by Wen and Yu [13]. $K_2/K_1$ is obtained with using the following equation:

$$K_2 / K_1 = 85.714 (1- \varepsilon_{mf} )/ \Phi_s \qquad (13)$$

$\Phi_s$ is spherical factor which is equal 1 for spherical particles. $\Phi_s$ is defined as following: $\Phi_s$ = Sphere Surface/ Particle Surface which has the Same Volume as Sphere ; $\Phi_s$ for Sharp Sand, Round Sand and for Fischer Tropsch Catalyst is 0.67, 0.86 and 0.58 resp.

The values of $\varepsilon_{mf}$ for spherical particles with presented $K_2/K_1$ for $K_1$ and $K_2$ with $\varepsilon_{mf}$ is calculateable. Table 4 presents the values which are calculated for different states of $\varepsilon_{mf}$. The calculated value of $\varepsilon_{mf}$ based on $K_2/K_1$ is a little smaller than the calculated values based on two separated values obtained using the equations of $K_1$ and $K_2$. The calculated values of $\varepsilon_{mf}$ at Table 4 and the obtained values using equations of $K_1$ and $K_2$ are nearly closed, and $\varepsilon_{mf}$ between two values obtained from $K_1$ and $K_2$ is varied.

$$\varepsilon_{mf} (K_2) < \varepsilon_{mf} < \varepsilon_{mf} (K_1) \qquad (14)$$

Table 4 Values which are calculated for different states of $\varepsilon_{mf}$

| Type of Equation | Wen and Yu [13] | Richardson [21] | Saxena & Vogel [22] | Babu *et.al.*[23] | Grace, [24] | Chitester *et.al.* [25] |
|---|---|---|---|---|---|---|
| $K_2/K_1$ | 67.4 | 57.4 | 50.6 | 50.6 | 54.4 | 57.4 |
| $1- \varepsilon_{mf}$ | 0.7863 | 0.5996 | 0.59 | 0.59 | 0.6346 | 0.6696 |
| $\varepsilon_{mf}$ | 0.2137 | 0.4004 | 0.41 | 0.41 | 0.3654 | 0.331 |
| $K_2$ | 1651.96 | 1408.21 | 886.16 | 777.26 | 1333.3 | 1161.94 |
| $K_1$ | 24.5 | 27.39 | 17.51 | 17.51 | 24.5 | 20.242 |
| $1/\varepsilon_{mf}^3$ | 14 | 15.655 | 10.007 | 10.007 | 14.005 | 11.566 |
| $\varepsilon_{mf\ (k1)}$ | 0.4149 | 0.3997 | 0.4646 | 0.4646 | 0.4149 | 0.4421 |
| $1\text{-}\varepsilon_{mf}/1-\varepsilon_{mf}$ | 11.013 | 9.388 | 5.9077 | 5.9077 | 8.888 | 7.7462 |
| $\varepsilon_{mf\ (k2)}$ | 0.3827 | 0.3998 | 0.4525 | 0.4525 | 0.4058 | 0.421193 |

If the value of $\varepsilon_{mf}$ from equation of Broadhrst and Becker [2] for silicagel catalyst regarding of catalyst physical properties is calculated, $\varepsilon_{mf}$ will be calculated using the following equation:

$$\varepsilon_{mf} = 0.586 (\mu^2 / \rho_g \; g (\rho_g - \rho_s ) d_p{}^3)^{0.029}(\rho_g / \rho_s )^{0.021} \qquad (15)$$

The following value is obtained for nitrogen gas and silicagel catalyst: $\varepsilon_{mf}$ = 0.41516. The values of $\varepsilon_{mf}$ for spherical particles with presented $K_2/K_1$ for $K_1$ and $K_2$ with $\varepsilon_{mf}$ is calculatable.

## 6. Results and Discussion

The distributor pressure drop at distributor with mesh size 3 is maximum 5.5 cm-$H_2O$ which is almost 30% of bed pressure drop (17.5 cm- $H_2O$). Regarding that distributor pressure drop is appropriate and this distributor is appropriate for fluidized bed and the obtained value of $U_{mf}$ ($U_{mf}$ = 2.9 cm/sec) is acceptable.

1. The value of $U_{mf}$ is calculated using the equations which presented in articles and books. The calculations indicate that for silicagel catalyst with averaged diameter 212 micrometer and solid density 1.96 g/cm$^3$, the equation presented with maximum 14.3 cm/sec (Gorosho *et al.*, [27]), the equation presented with minimum 1.493 cm/sec (McKay and McLain [11]) and the equations presented with $U_{mf}$ 2.915 cm/sec (Wen and Yu [14]), 3.0 cm/sec (Rowe and Yacono [30]) and 3.6 cm/sec are more appropriate respectively. Then the only equation for silicagel catalyst which

predicts real $U_{mf}$ is equation of Wen and Yu [14] and the equation presented by Rowe and Yacono for $U_{mf}$ has a small variation from real $U_{mf}$.

2. Considering fluidization, fluidized bed with very small porousity of distributor is appropriate which distributor bed has proper pressure drop and the variations of catalyst height at the bed also are more in comparison with the others cases of catalyst height. In third status, distributor pressure drop is higher than first and second statuses.
3. The minimum value of fluidized height is obtained using $U_{mf}$ = 2.9 cm/sec while the variation of catalyst height versus velocity for bed with diameter 4.8 cm, distributor type of porous and 400 gr of catalyst are measured. This value is only applicable for the above condition and they have to obtain for the other states (diameter, type of distributor and catalyst weight through experiments.)
4. The porosity values in condition of minimum of fluidization ($\varepsilon_{mf}$ ) is calculated using Broadhurst and Becker [2] equation and presented values of $K_1$ and $K_2$ in general equation of $Re_{mf}$ for different states are based on the values of $K_1$ and $K_2$. Maximum and minimum of $\varepsilon_{mf}$ are 0.48 and 0.41 based on the values of $K_1$ and $K_2$ respectively. Basically, based on the ratio of $K_2$ to $K_1$ ($K_2$/ $K_1$), minimum and maximum of $\varepsilon_{mf}$ are 0.21 and 0.41. In order to calculate $U_{mf,}$ the value of $\varepsilon_{mf}$ which is resulted from Broadhurst and Becker equation is used.

## Symbols

| | |
|---|---|
| Ar , $(d_p^3\ \rho_g\ (\rho_s - \rho_g)/\mu'^2$ | *Arschmides* number |
| $K_1$ $(1/\Phi_s\ \varepsilon_{mf}{}^3)$ | A function of porosity and the figure factor in general equation |
| $K_2$, $(1 - \varepsilon_{mf}/\Phi_s\ \varepsilon_{mf}{}^3)$ | A function of porosity and figure factor in general equation |
| $d_p$, m | Particle diameter |
| g ,$m/s^2$ | Accelerative gravity |
| $\Delta P$, $N/m^2$ | Pressure drop |
| $H_{mf}$, cm | The height of bed at the condition of minimum of fluidized height |
| Re, $U.\ d_p.\ \rho_g\ /\ \mu$ | Reynolds number based on particle diameter |
| $Re_{mf,}$ $U_{mf}.d_p.\rho_g\ /\ \mu$ | Reynolds number in the state of minimum of fluidized velocity |
| U, m/s | Nominal velocity of fluid |
| $U_{mf,}$ m/s | Minimum of velocity of fluid |

## Greek Alphabets

| | |
|---|---|
| $\varepsilon$ | Porosity which is dimensionless |
| $\varepsilon_{mf}$ | Porosity in minimum of fluidized velocity which is dimensionless |
| $\Phi_s$ | Figure Factor (Spherical) which is dimensionless |
| M | $N/Sm^2$ Viscosity of Fluid |
| $\rho$ | $Kg/m^3$Density of Fluid |
| $\rho_s$ | $Kg/\ m^3$ Density of Particles |

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