Effect of fluidization on pressure drop and power consumption in packed bed columns

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Abstract. This research investigated the effect of different initial heights of packed bed on pressure drop and minimum fluidisation velocity in packing bed columns. Glass beads with different sizes (6 and 8 mm \(d_p\)) with density 2, 550 kg/m\(^3\) were used, with distilled water as the liquid. In order to study the effect of changing the properties of water, a heater with a temperature controller was used to control the temperature of water for each experimental run at one of two levels (20 and 30 °C) to examine any changes occurring in terms of values of pressure drop at minimum fluidisation velocity.

All experimental runs were carried out in a cylindrical Perspex column with a 5 cm inside diameter, 5 mm wall thickness, and 100 cm length; six pressure sensors of transducer type (analogue) were fixed on wall of the column and at bed height in order to measure the pressure; the pressure differences between any two points were thus calculable. The pressure sensors were located 10 cm apart, beginning at the bed mesh holder. The pressure data were recorded using the LabVIEW program and an Arduino.

A comparison was made between a stationary (fixed) and fluidised bed by measuring the pressure drop in each case via pressure sensors at the same flow rate, temperature level, and initial height of bed. It was observed that the minimum fluidisation did not change with changes in the initial height, though pressure drop increased with the increase in initial height and both pressure drop and minimum fluidisation velocity increased with increase in temperature. The fluidisation state showed a lower pressure drop than the fixed state, reducing power consumption. For validation Octave Levenspiel’s equations for both pressure drop and minimum fluidisation velocity were applied, showing good agreement with the experimental work in terms of using the liquid as a working fluid.

1 Introduction

The multi-phase flow system began to be used in the industrial sector in the early 1920s, and the fluidised bed system was patented by Fritz Winkler in 1923 for gasification in Germany industry. Researchers in Germany outlined the fluidised bed system in detail in 1930, developing this technique and building a commercial fluidised bed system plant in 1940. Germany industry used the fluidised bed system in two main sectors: coal gasification and metal refining processes [1, 2].

In 1878, one technique rose to dominate the field of catalytic cracking processes, technique now known as fluidisation. Two researchers from an oil development company, Gilliland and Lewis, conducted extensive research on the measurement of the properties of using dense phase solid packing and working fluid gasses, thus developing the concept to the liquid-solid fluidised bed system. The first plant to use the principle of the two-phase flow (solid-liquid) fluidised bed system for industrial applications was operational by 1930. In 1953, Dale Worsted patented a method of coating the tablets by using the principles of the fluidised bed system, using warm air to suspend the tablets and spraying the coating solution on the bed to facilitate the coating process. In 1960, the fluidisation technique was used in large areas in the paramedical industry, and Dale Worsted
also used a fluidised bed in the granulation of powder. In 1980, there was an increase in research, commercialisation, and new applications of the fluid-solid fluidised bed system arose [3]. The fluidised bed system was used significantly in thermal catalytic cracking as well as in various industrial fields such as coal gasification, creating gasoline from other petroleum fractions, creating gasoline from natural and synthesis gases, synthesis reactions, calcination, separation, mixing and drying, heat exchange, ion exchange, and bio fluidisation [4, 5].

The fluidised bed offers several advantages over fixed beds and it is thus used in a range of wide fields and industries, particularly those with continuous operations with automatic controls that require easy to handle processes with good mass and heat transfer properties; in addition, for those processes that require large exothermal reactions or high thermal stability, the circulation of packing between more than one fluidised bed makes it possible to add or remove heat in large reactors, managing the rate of mass and heat transfer between the immersed object and fluidised bed more effectively than in a fixed bed [6]. There is more one system of the fluidised bed, and they can be classified depending on working fluid: major types are liquid-liquid fluidising beds, liquid-liquid-solid fluidising beds, and liquid-gas-solid fluidising beds [7, 8, 9].

Due to the increase in the number of industrial areas using fluidised bed systems, more researchers have begun seeking developments in terms of fluidised bed system design, scale-ups, control procedures, and operating measurements. Fluidised bed systems offer many advantages, such as lower energy consumption compared to other multi-phase flow systems, maximum product quality, easy operation conditions, and environment-friendly systems [3, 9].

Many researchers such as Vollmari, Jasevičius, and Kruggel-Emden have studied the fluidisation behaviour for spherical and non-spherical particles experimentally and theoretically, and by using simulation. The experimental and theoretical results for spherical particles show good agreement, while for non-spherical particles, the pressure drop depends on the height and orientation, and differences in particle height lead to odd predictions close to column walls or in corners. CFD can accurately reproduce the orientation behaviour of elongated particles; however, for elongated particles with ratios below a certain elongation, this behaviour cannot be observed, results confirmed by both simulation and experiments [3].

Ergun and Orning showed that the relationship between the pressure drop and superficial velocity in a packed bed is a linear function that depends on many parameters such as particle specific surface area, porosity, and viscosity of the fluid [4]. Khan et al. studied the effect of mixing more particle sizes on expansion behaviours, determining the types of segregation seen: glass beads were used with diameters of 3, 5, and 8 mm, with water as the working fluid. They observed that type of segregation depended on the diameter of particles mixed, specifically the diameter ratios; the height-diameter ratio of the mixing zone was small and decreased with increases in superficial liquid velocity. For lower diameter ratios, the mixing zone was larger and increased with increases in superficial liquid velocity [5]. Khan et al studied the expansion bed in binary liquid-solid fluidised beds both experimentally and numerically using three different sizes of borosilicate glass beads (3, 5, 8, mm) with density 2,230kg/m³ with equal and unequal mass ratios in water. They observed that the segregation increased with increases in differences in the size of particles vice versa; the mixing zone, however, decreased with increases in flow rate [7]. Lee found evidence that any increase in velocity is accompanied by an increase in pressure drop, with maximum pressure drop occurring when the velocity reaches minimum fluidisation velocity, beyond which the pressure drop decreases slightly, becoming almost constant with increase in flow rate [8]. Rao and Curtis studied the effect of various bed heights, column diameters, and walls to see their effects on the minimum fluidisation velocity. They concluded that increasing the height of the bed leads to increases in minimum fluidisation velocity, reducing the diameter of the column.
leads to increases in minimum fluidisation velocity, and that wall effects also have an effect on minimum fluidisation velocity [10].

2 Hydrodynamic Properties of Fluidising Beds

The hydrodynamics of fluidised beds are very important to any understanding of the behaviour of these multiphase flow systems [11]. The hydrodynamics of multiphase flow systems, especially in liquid-solid systems, are very complex, as there are multiple interactions inside the bed during the process, such as interactions between the working fluid in dense phase and particles as well as the interactions of the dense phase with the wall of the reactor. These hydrodynamics become more complex when the system has more than one size of particle, such as binary systems, particularly when the dense phase particles differ in density or shape as well [12].

Several researchers have thus investigated the hydrodynamics of multi-phase flow systems and fluid-solid fluidised bed systems for upward flow types (Kwauk and Wilhelm, 1948; Ergun 1952; Zaki and Richardson, 1954; Watson and Begovich 1978; Epstein et al., 1981; Fan et al., 1985; Kwauk, 1992; Zhang et al., 1995; Lee et al., 1999, 2000). However, little research on the hydrodynamics of two-phase flow fluidised bed systems of downward flow type has been disseminated [8].

2.1 Minimum Fluidising Velocity

The hydrodynamics of fluidised beds are very complex, though they must be understood to improve fluidised bed operations. One of the most important parameters of the hydrodynamics a fluidised bed is the minimum fluidisation velocity, the separation point between the fixed and fluidisation states.

Fluidisation is controlled by varying the liquid flow rate. Suppose that a liquid is passed vertically through a bed of particles at low liquid velocity; the particles are stationary, and the bed is a packed bed. The velocity, known as the superficial liquid velocity, can be calculated as

\[ U_s = \frac{Q}{A_c} \]  

When the liquid flow is increased to the point where the pressure drop across the bed is equal to the weight of the particles per unit area, the bed is said to be fluidised, and that velocity is called the minimum fluidising velocity [1]. For a fixed state, both Levenspiel [6] and Geankoplis [13] used the Archimedes number and the Reynolds number under minimum fluidisation conditions to calculate the minimum fluidising velocity, using the formula

\[ k_1Re_{p,mf}^2 + K_2Re_{p,mf} = Ar \]  

where

\[ K_1 = \frac{1.75}{\epsilon_{mf}^3 \phi_p} \]  

and

\[ K_2 = \frac{150(1 - \epsilon_{mf})}{\epsilon_{mf}^2 \phi_p^2} \]

Wen and Yu [14] were the first to note that (K1, K2) remain almost constant for multiple types of particles over a wide range of conditions (Reₚ = 0.001 to 4000), thus allowing predictions of minimum fluidising velocity with a ±34% standard deviation. Since then, other investigators, as shown in table 1 have reported further on K₁ and K₂.
Table 1 Researchers’ values for K1 and K2.

| Investigation/focus                              | first $K_2/2K_1$ ($C_1$) | second $1/K_1$ ($C_2$) |
|--------------------------------------------------|----------------------------|------------------------|
| Wen and Yu (1966)                                | 33.7                       | 0.0408                 |
| 284 data points from the literature              |                            |                        |
| Richardson (1971)                                | 25.7                       | 0.0365                 |
| Saxena and Vogel (1977)                          | 25.3                       | 0.0571                 |
| Dolomite at high temperature and pressure        |                            |                        |
| Babu et al. (1978)                               | 25.3                       | 0.0651                 |
| Correlation of reported data to 1977             |                            |                        |
| Grace (1982)                                     | 27.2                       | 0.0408                 |
| Chitester et al.                                 | 28.7                       | 0.0494                 |
| Coal, char, balotini at up to 64 bar              |                            |                        |

2.2 Pressure Drop ($\Delta p$)

The pressure drop for fluid flow through a packed bed in a column with solid particles is one of the important hydrodynamic measures that must be determined. In the fixed state, this is commonly evaluated by means of the empirical Ergun (1952) equation [14]. The Ergun equation is a general equation used for all levels of Reynolds number [13].

$$\frac{\Delta p_{fr}}{l_m} = 150 \frac{(1 - \epsilon_m)^2}{\epsilon_m^3} \frac{\mu_f \ u_s^2}{(\phi_p \ d_p)^2} + 1.75 \frac{(1 - \epsilon_m) \ \rho_f \ u_s^2}{\epsilon_m^3 \ \phi_p \ d_p}$$

(5)

- Octave Levenspiel, 1991[6]

The frictional pressure drop for upward flow is always positive through fixed beds of length ($L$) containing a single size ($d_p$), as correlated by Ergun.

The measured pressure drop is

$$\Delta p_b = \Delta p_{fr} \pm \frac{\rho_f \ l_m}{g_c}$$

(6)

where the + sign stands for the upward flow of fluid. The last term may be appreciable for flowing liquids, but it can safely be ignored for flowing gases unless deep beds at high pressure are involved. Thus, in most cases with gases, the equation may be written:

$$\Delta p = \Delta p_b = \Delta p_{fr}$$

(7)

2.3 Porosity ($\epsilon$):

One of the most important hydrodynamic factors of a fluid flowing in a packed bed column is porosity, which depends on the shape and size of the particles involved. This can be calculated experimentally or theoretically.
2.3.1 Experimental method:

Two measuring cylinders were used, with one filled to any level with water and the other filled with packing. The levels of the two cylinders are recorded. The water from the first is then added to the packing cylinder until it just covers all packing, and the two cylinder levels are recorded again, with equation (8) used to calculate the porosity of the packing:

\[ \epsilon = \frac{\text{the volume of water cylinder before add} - \text{volume of water cylinder after add}}{\text{the volume of packing cylinder after add}} \]  

(8)

2.3.2 Theoretical method:

Using equation (9) [13],

\[ \epsilon = \frac{\text{volume of the voidge}}{\text{volume of the inter bed}} \]  

(9)

3 Experimental work

3.1 Experimental setup

A fluidised bed is a complex system. The experimental cylindrical test section used in experiments was made from acrylic of 1m length, 5cm ID, and 6cm OD, and six transducer sensors were connected to the wall, with an Arduino and the LabVIEW program along with other equipment completing the experimental rig, as shown in figure 1. The rig consisted of two parts, a solid part and an electronic part, as defined in table 2.

![Experimental rig](image)

Figure 1. Experimental rig.

Table 2. Parts of the experimental rig.

| No. | Name         | No.  | Name         |
|-----|--------------|-----|--------------|
|     |              |     |              |
4 Results and discussion

4.1 Pressure drop

4.1.1 Effect of the flow rate on pressure drop

The flow rate of the working fluid has a significant positive effect on pressure drop: increasing the flow rate of the working flow leads to an increase in the interaction between the working fluid and the particles and the working fluid and the wall, increasing friction and leading to an increase in the pressure drop. This case occurs with the bed both in the fixed state and in the initial state of a fluidised bed. At the point when the weight of the packed bed is equal to the buoyance force in the fluidised bed, however, the relationship become constant and the pressure drop reaches a maximum value. After that point, increases in the flow rate of the working fluid lead to pressure drop decreases due to the increasing flow rate forcing bed expansion, which leads to increases in the spacing between the particles causing a decrease in pressure drop. These effects are shown in figures 2 and 3 for fluidised and fixed beds.

4.1.2 Effect of particle diameter on pressure drop

Figures 4 and 5 show the effects of particle diameter for two sizes (6 and 8 mm) of sphere. It is clear from these that increasing in the diameter of packing leads to a decrease in the pressure drop for both fluidisation and fixed states. In the fluidisation state, the maximum pressure drop for 8 mm particles is 9305.357 Pascals and for 6 mm, it is 9420.612 Pascals. The increase in
diameter leads to increases in void space in the packing and this in turn leads to increases in spacing in the bed, reducing interaction between the fluid and particles, leading pressure drop to decrease. In the fixed state, the pressure drop for 8 mm particles is less than for 6 mm particles for the same flow rate.

**Figure 4.** Effect of diameter of particles on pressure drop at fluidisation state.

**Figure 5.** Effect of diameter of particles on pressure drop at fixed state.

### 4.1.3 Effect of temperature on pressure drop.

Figures 6, 7, 8, and 9 show the pressure drop in both fluidisation and fixed states for 6 and 8 mm spheres increasing with increases in the fluid temperature. This is because increasing temperature leads to a decrease in several key properties of the working fluid such as density and viscosity, which leads to a decrease in the force of the fluid and an increase in flow rate. Thus, for 8mm spheres, the maximum pressure drop at 20 °C is 9,305.357 Pascals and at 30°C, it is 9,682.749 Pascals; for the 6 mm sphere at 20°C it is 9,420.612 Pascals and at 30°C, it is 9,753.393 Pascals. Similar behaviour is seen in the fixed state: increases in temperature cause pressure drop increases.
4.1.4 **Effect of initial height on pressure drop**

Figures 10, 11, 12, and 13 show that this parameter has more effect on pressure drop than changes in temperature and diameter.

For both fluidisation and fixed state and for both sizes of particle tested (6 and 8 mm), an increase in initial height leads to an increase in pressure drop, which may be attributed to the fact
that increasing initial height leads to increasing loading and weight, thus increasing the interaction between solid and liquid.

4.1.5 **Validation**

As shown in the figures 14, 15, 16, and 17, there is good agreement between the pressure drop levels from the experimental result and the theoretical results from the equation by Ergun and Octave.
Levenspiel for the fixed and fluidisation states, with all error percentages below the ±25% mark, suggesting that these are acceptable values. Figures are given in tables 3, 4, 5, and 6 [6].

![Figure 14](image1.png) 8 mm glass sphere at 20°C.

![Figure 15](image2.png) 8 mm glass sphere at 30°C.

![Figure 16](image3.png) 6 mm glass sphere at 20°C.

![Figure 17](image4.png) 6 mm glass sphere at 30°C.

**Table 3.** Error percentages for 8 mm sphere at 20°C.
### Table 4. Error percentages for 8 mm sphere at 30°C.

| velocity | Experimental | Ergun | %Error | Octave | %Error |
|----------|--------------|-------|--------|--------|--------|
| 0.042462845 | 4590.642857 | 3875.210417 | 15.58458068 | 4301.202697 | 6.30502782 |
| 0.056617127 | 7520.107143 | 6664.072062 | 11.38328304 | 7090.064342 | 5.71857279 |
| 0.08492569 | 9181.285714 | 8436.672986 | 8.110113889 | 8890.599186 | 3.166076484 |
| 0.099079972 | 9112.357143 | 8167.417466 | 10.36987096 | 7900.861663 | 9.228400883 |
| 0.113234253 | 8891.472403 | 7754.921634 | 12.78428098 | 8248.753434 | 7.228487475 |
| 0.127388535 | 8704.112013 | 7382.088863 | 15.18848962 | 7900.861663 | 9.228400883 |
| 0.141542817 | 8333.150974 | 7108.678164 | 14.69399526 | 7647.403764 | 8.229146594 |

### Table 5. Error percentages for 6 mm sphere at 20°C.

| velocity | Experimental | Ergun | %Error | Octave | %Error |
|----------|--------------|-------|--------|--------|--------|
| 0.035385704 | 5555.75 | 4251.749345 | 20.61337171 | 4670.758145 | 12.78983999 |
| 0.038924275 | 7651.071429 | 5921.882114 | 22.60061654 | 6340.890914 | 17.12414434 |
| 0.056617127 | 9489.920408 | 8449.607668 | 10.96229152 | 8902.450968 | 6.190456983 |
| 0.099079972 | 9324.638776 | 8179.939338 | 12.27607273 | 8647.711538 | 7.259554542 |
| 0.113234253 | 9166.243878 | 7766.811088 | 15.26724368 | 8259.464788 | 9.892591789 |
| 0.127388535 | 9040.404453 | 7393.406709 | 18.2181865 | 7910.941909 | 12.49349572 |
| 0.141542817 | 8856.340816 | 7119.576831 | 19.61040142 | 7657.017231 | 13.54197642 |

### Table 6. Error percentages for 6 mm sphere at 30°C.

| velocity | Experimental | Ergun | %Error | Octave | %Error |
|----------|--------------|-------|--------|--------|--------|
| 0.035385704 | 5486.714286 | 4693.88781 | 14.44993187 | 5111.89701 | 6.83161296 |
| 0.038924275 | 7750.69 | 7128.100011 | 8.032704049 | 7546.109211 | 2.639517119 |
| 0.056617127 | 9560.392857 | 8647.88366 | 9.544683058 | 9090.77436 | 4.91225552 |
| 0.070771408 | 9298.67316 | 8079.199931 | 10.0921701 | 8553.078931 | 5.921865697 |
| 0.08492569 | 8722.597403 | 7831.877484 | 11.13421701 | 8813.26275 | 7.345691491 |
| 0.099079972 | 8864.903571 | 7843.885401 | 11.51753272 | 8320.721084 | 6.016317779 |

### 4.2 Minimum fluidisation velocity

#### 4.2.1 Effect of diameter on minimum fluidisation velocity

As shown in figure 4, the effect on minimum fluidisation velocity of increasing particle diameter is an increase in minimum fluidisation velocity. This is because increases in particle diameter create increases in the weight of particles, leading to increased minimum fluidisation velocity.

The minimum fluidisation velocity for 8 mm spheres was 0.084926 m/s, and for 6 mm spheres, it was 0.070771 m/s.
4.2.2 Effect of working fluid temperature on minimum fluidisation velocity

As shown in figures 6 and 8, increasing the temperature leads to an increase in the minimum fluidisation velocity due to decreases in key properties of the working fluid such as viscosity and density that are responsible for the fluid force; thus, decreasing these properties decreases the force, increasing in flow rate and generating a fluidisation state in the fluidised bed.

The minimum fluidisation velocity for 8 mm spheres at 20°C was 0.084926 m/s and at 30°C, it was 0.093007 m/s, while for the 6mm sphere at 20 °C it was 0.070771 m/s and at 30°C, it was 0.07875 m/s. Validation between the experimental and theoretical results was done using equation 2, as shown in table 7.

Table 7. Error percentages for minimum fluidisation velocity.

| Particle diameter | Temperature °C | Experimental m/s | Theoretical m/s | %Error |
|------------------|----------------|------------------|-----------------|--------|
| 8mm glass sphere | 20             | 0.084926         | 0.086023791     | 1.29   |
| 8mm glass sphere | 30             | 0.093007         | 0.086821692     | 6.65   |
| 6mm glass sphere | 20             | 0.070771         | 0.0718217       | 1.48   |
| 6mm glass sphere | 30             | 0.07875          | 0.0727876       | 7.57   |

4.3 Power consumption

Operating and capital costs are very important in industry in terms of selecting equipment and processes. Comparing fluidised beds and fixed beds in these terms suggests, as that fluidised bed pressure drop increases with flow rate and reaches an expanded state when the pressure drops, after which it becomes constant while the fixed bed shows increasing pressure drop as the flow rate increases, and that this state to end flow levels, the cost of the fluidised bed is less than that of the fixed. Fluidised beds are thus widely used in industrial applications. Equation 10 can be used to determine the cost of power consumption.

Power consumption generally refers to the electrical energy supplied to operate an electrical appliance over time. Energy losses or friction losses refers to the difference in pressure required to overcome the pressure drop during flow through pipes; such losses only occur as a result of dynamic movement and can be thought of as dynamic differential pressure. Friction losses thus only occur when flow actually takes place.

\[
Power\ consumption = \frac{\Delta p \cdot Q}{\epsilon} \tag{10}
\]

Power consumption calculations are shown in table 8 and table 9.

Table 8. Costs in fluidisation state.

| Type of packing | \(\Delta P_{max}\) | Q       | \(\epsilon\) | cost   |
|-----------------|-----------------|---------|--------------|--------|
| 8mm sphere      | 9305.357        | 1.666*10^-4 | 0.44         | 3.52 w |
| 6mm sphere      | 9420.612        | 1.388*10^-4 | 0.433        | 3.01 w |

Table 9. Costs in fixed state.

| Type of packing | \(\Delta P_{max}\) | Q       | \(\epsilon\) | cost   |
|-----------------|-----------------|---------|--------------|--------|
### Nomenclature

| Symbols | notations | units |
|---------|-----------|-------|
| $\Delta p$ | = pressure drop | [Pascals] |
| $\Delta p_{fr}$ | = friction pressure drop | [Pascals] |
| $l_m$ | = fixed bed initial height | [m] |
| $l_{mf}$ | = expansion bed height | [m] |
| $\epsilon_m$ | = fixed bed porosity | --- |
| $\epsilon_{mf}$ | = fluidised bed porosity | --- |
| $\phi_p$ | = sphericity | --- |
| $d_p$ | = particle diameter | [m] |
| $\mu_f$ | = working fluid viscosity | [$kg/m/s$] |
| $\rho_f$ | = working fluid density | [$kg/m^3$] |
| $\rho_p$ | = particle density | [$kg/m^3$] |
| $u_s$ | = working fluid superficial velocity | [m/s] |
| $u_{mf}$ | = minimum fluidisation velocity | [m/s] |
| $u_s$ | = settling velocity | [m/s] |
| $Q^0$ | = Working fluid Volumetric flow rate | [$m^3/s$] |
| $A_c$ | = Test section Cross sectional area | [m$^2$] |
| $Ar$ | = Archimedes number | --- |
| $Re$ | = Reynolds number | --- |
| $Re_p$ | = Particle Reynolds number | --- |
| $Re_{p, mf}$ | = Particle Reynolds number at minimum fluidisation condition | --- |

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