

# Hydrodynamic Studies on Conventional and Tapered Fluidized Bed

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**Abstract:** *In this project the determination of maximum pressure drop, minimum fluidization velocity, minimum bubbling velocity and fluidization index for various powders are to be studied using a conventional cylindrical bed and a tapered bed. Tapered beds have a velocity gradient along the axis unlike conventional beds. They have been used as an alternative to conventional cylindrical beds to deal with problems such as slugging. Initially air at ambient conditions (310 K) is to be used as the fluidizing medium. High temperature studies will also be carried out using a High temperature fluidized bed (Air temperature up to 1173 K).*

**Keywords:** Tapered, conventional, fluidized bed, ambient and high temperature

## 1. Introduction

Most of the gas–solid fluidization behaviour studies that have been reported are for straight cylindrical or columnar fluidized beds; although a considerable portion of the fluidized beds have inclined walls or have a tapered bottom section. A velocity gradient exists in the axial direction in tapered bed leading to unique dynamic characteristics of the bed. Due to this characteristic, tapered fluidized beds have found wide application in many industrial processes such as, waste water treatment, immobilized biofilm reaction, incineration of waste materials, coating of nuclear fuel particles, crystallization, coal gasification, roasting sulfide ores and food processing, etc. Tapered fluidized beds can be operated smoothly without any instability, i.e. with less pressure fluctuations. Tapered fluidized beds are very useful for fluidization of materials with a wide particle size distribution, as well as for exothermic reactions and also for extensive particle mixing. Despite their widespread application, much of the development and design of tapered fluidized bed reactors have been empirical in nature as the complex behaviour of gas–solid flow in these systems makes flow modelling a challenging task. In addition, numerical solutions of complex non-linear equations, with moving phase boundaries, are difficult to obtain. However, with increasing computational capabilities and computational fluid dynamics (CFD) tools in recent years, several researchers are involved in studying the hydrodynamics of gas–solid systems, which would be useful in the design, optimization and scale-up process.

Due to good solid mixing, high heat transfer, large contact surface area, etc., fluidized bed technology has been exhaustively used in many industrial processes such as coal combustion, gasification, polymerization and catalytic cracking. There are many variables determining the smooth running of fluidized beds like bubble frequency, bed expansion and minimum fluidization velocity  $U$  that plays a critical role in the design and operation of fluidized beds is one of the most important parameters. Therefore, numerous investigations have been undertaken to study the variation of  $U$  then,  $U$  under different operational conditions. To predict  $U$  precisely, many previous studies have proposed the

empirical correlations, as listed in Table 1.

Among them, most correlations have been developed for narrow-cut particles. However, when dealing with natural minerals or coal combustion etc., wide PSD particles are usually fluidized as bed materials. At this point, the prediction errors for the traditional correlations are too large to give useful results. In order to remedy this situation, some correlations have been developed for the particles with wide PSD, such as C8 in Table 1 proposed by Zhejiang University (in what follows, it's referred to as "ZJC"), C9 proposed by Northwest University (China) (referred to as "NWC") and C10 proposed by Huazhong University of Science and Technology (referred to as "HZC"). However, these correlations also have their short comings.

## 2. Literature Survey

Application of tapered fluidized bed technique is versatile in the field of high temperature processes viz. coal combustion and gasification, gas–solid catalytic reaction, etc. Prediction of minimum fluidized velocity at elevated temperatures is a pre-requisite in the design and operation of high temperature fluidized beds. Several investigations have been made during the last few decades and the Ergun's [13] equation has been used to correlate the experimental data in most of the cases [14–20]. Equations with particle sphericity range have been proposed by Lukas et al. [21] based on Brownell's approach [22] of bed compactness to eliminate the minimum bed porosity dependence of the minimum fluidization velocity. Yamazaki et al. [23] have developed a model for predicting minimum fluidization velocity at elevated temperatures based on quantitative relationship between the velocity and the adhesive force of particles. In most of the above elevated temperature studies, spherical particles were frequently used in order to eliminate eventual complications caused by non-spherical shapes of solids which is encountered in actual practice. Further all the studies were conducted in conventional (cylindrical) beds. In the recent past, fluidization in tapered bed is gaining importance in view of its potential application in gas–solid system involving continuously decreasing particle size in combustion and gasification of coal, reduction of iron ores [24], etc. Although some

investigations have been carried out in conical bed [25,26] and tapered bed [27,28] for prediction of minimum fluidization velocity under ambient conditions, similar data at elevated temperature conditions in tapered bed are not available. In this present work, investigations have been carried out at elevated temperatures to obtain the correlation of minimum fluidization velocity (independent of bed voidage) in tapered bed for gas–solid system and the correlation has been compared with values obtained by equation of Sau et al. [28] proposed for tapered bed in ambient conditions in gas–solid systems. In addition, the best fit to the presented results have also been compared with the equations developed for conventional (cylindrical) beds available in the literature under ambient conditions and at elevated temperatures.

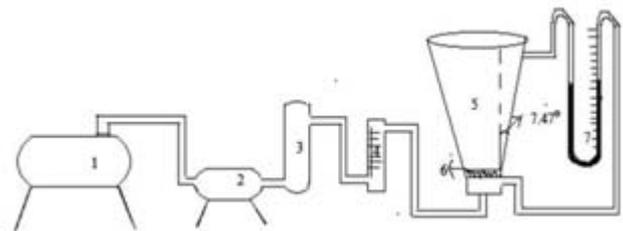
Although many fluidized beds are operated at high temperature, before 1980's, most experiments determining U have been conducted at ambient temperature. Along with the wide applications of fluidized beds to the industry, the prediction errors of U<sub>mf</sub> at high temperature began to attract more attention. Botterill et al. (1982) [29] have observed that the U of Geldart B-type particles decreases with the increasing bed temperature, opposite to the behaviour of Geldart D-type particles. Pattipati and Wen [30] have pointed out that the increasing temperature causes a slight decrease of U<sub>mf</sub> for particles, whose diameter is smaller than 2 mm, unlike particles larger than 2 mm. Goo et al. [31] have noted that U<sub>mf</sub> decreases as the bed temperature increases, regardless of particle diameters or particle types. Wu and Baeyens [32] and Subramani et al. [33] have also observed similar behaviour. Formisani et al. [34] have showed that the increase in bed temperature has no effect on the U<sub>mf</sub> of fine particles but leads to a decrease in U<sub>mf</sub> for coarse particles. Chiou et al. pointed out that bed temperature and PSD can influence the U<sub>mf</sub> simultaneously, making U<sub>mf</sub> vary non-monotonically with temperature [35].

### 3. Method and Experimental Details

#### 3.1. Tapered fluidized bed at ambient temperature

The tapered bed used in this study has column bottom and top diameters of 41.5 mm and 540 mm respectively. The tapered angle was 7.47°. A cloth having a pore size less than 20 μm was used as the distributor. A U Tube Manometer having Carbon tetrachloride as the Manometric fluid (density = 1630 Kg/m<sup>3</sup>) was used to measure the bed pressure drop. Air was used as the fluidizing medium (At Temperature=310 K, Density = 1.17 kg/m<sup>3</sup>). A silica gel tower was used to remove the moisture present in the air. A rotameter having a range of 0-50 LPM was used to measure the air flow rates. Hematite powder of particle size 20μm was used. The ratio of mass/poured volume of the powder gives us the bulk density. Before the start of the tests the bed was fluidized once so that the powder bed is level so that the initial height can be noted properly.

The voidage was measured by the formula-  
 Voidage = 1 - (bulk density/ Particle density)

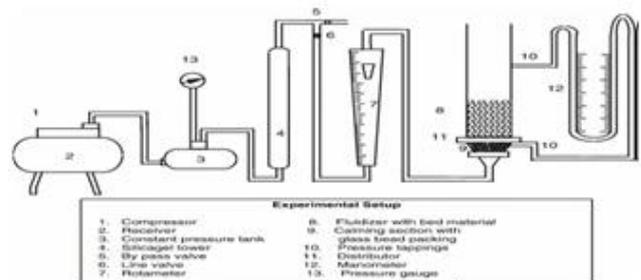


**Figure 1:** Experimental Setup: (1) Compressor, (2) Receiver, (3) Silica gel tower, (4) Air Rotameter, (5) Tapered bed with tapered angle 7.47o, (6) Powder bed, (7) U-tube manometer.

#### 3.2. Conventional fluidized bed at ambient temperature-

The experiment was carried out using hematite as solid particle, compressed air as gas. Initially the column is filled with hematite and for different gas flow rate variation of pressure drop and bed height is taken. The scope of the experiment has been presented in the Table below.

Pressure drop and the expanded bed height were noted. Experiments were conducted at normal temperature of (30±5) 0C. The temperature of the air was presumed to be at normal condition i.e.at 25 0C. The procedure was repeated for different gas flow rate, particles of 10micrometer sizes and varying initial static bed heights.



**Figure 2:** Experimental set-up of conventional fluidized bed

## 4. Observations and Calculations

### 4.1. Conventional fluidized bed at ambient temperature

**Table 1:** At Bed height = 6cm for hematite

Flow rate (LPM)	Velocity (M/S) *10 <sup>-3</sup>	Manometer (Maximum) H <sub>1</sub>	Manometer (Minimum) H <sub>2</sub>	$\Delta H=(H_1-H_2)$ (cm)	$\Delta P= \rho g \Delta H$ (Kpa)	Bed height Lf (Cm)	(Lf-Lm)/Lm
1.5	6.48	25	24.8	0.9	1.19	12	1
4.5	19.86	25.2	24.1	1.1	1.58	14	1.3
6.5	28.04	25.3	24.0	1.3	1.73	17	1.83
7	30.46	25.4	23.9	1.4	1.86	18	2
8	34.76	25.3	24.0	1.4	1.86	20	2.33
9	38.88	25.4	24.0	1.4	1.86	21	2.5

**Table 2:** At Bed height = 7cm for hematite

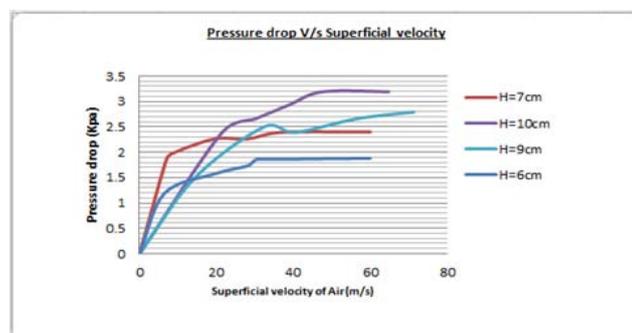
Flow rate (LPM)	Velocity (M/S) *10 <sup>-3</sup>	Manometer (Maximum) H <sub>1</sub>	Manometer (Minimum) H <sub>2</sub>	$\Delta H=(H_1-H_2)$ (cm)	$\Delta P= \rho g \Delta H$ (Kpa)	Bed height Lf (Cm)	(Lf-Lm)/Lm
1.0	4.28	25.1	24.0	1.2	1.86	8.5	0.46
2.0	8.64	25.3	23.9	1.4	1.98	12.0	0.54
4.5	19.36	25.5	23.8	1.7	2.26	15.5	1.18
6.5	28.12	25.6	23.6	1.7	2.26	19.0	2.47
8.6	36.79	25.7	23.5	1.8	2.39	20.0	2.63

**Table 3:** At static Bed height = 9cm for hematite

Flow Rate (LPM)	Velocity (M/S) * 10 <sup>-3</sup>	Mano meter (Maximum) H <sub>1</sub>	Mano meter (Mini mum) H <sub>2</sub>	$\Delta H=(H_1-H_2)$ (cm)	$\Delta P=\rho g \Delta H$ (Kpa)	Bed height Lf (Cm)	(Lf-Lm) /Lm
4.5	14.44	25.7	24.5	1.2	1.5	24	1.66
7.5	32.4	25.8	23.9	1.9	2.5	30	2.33
9.5	40.36	25.0	23.2	1.8	2.39	36	3
13.5	56.58	25.1	23.1	2.0	2.66	38	3.33
16.5	71.28	25.1	23.0	2.1	2.79	37	3.01

**Table 4:** At Static Bed height = 10cm for hematite

Flow rate (LPM)	Velocity (M/S) * 10 <sup>-3</sup>	Mano meter (Maximum) H <sub>1</sub>	Mano meter (Minimum) H <sub>2</sub>	$\Delta H=(H_1-H_2)$ (cm)	$\Delta P=\rho g \Delta H$ (Kpa)	Bed height Lf (Cm)	(Lf-Lm) /Lm
5	21.6	25.3	24.2	1.1	1.46	27	1.7
7	30.24	25.8	23.8	2.0	2.66	35	2.5
9	38.82	25.9	23.7	2.2	2.93	45	3.5
11	47.58	26.0	23.6	2.4	3.19	47	3.7
15	64.8	26.0	23.6	2.4	3.19	48	3.8

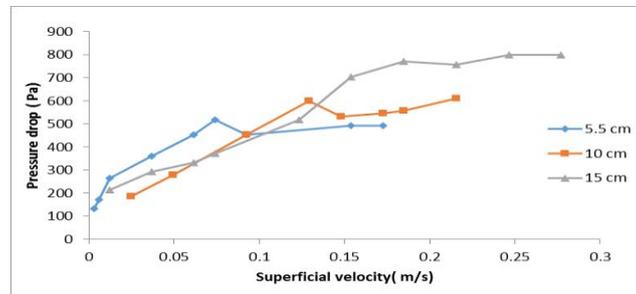


**Figure 3:** The graph between the superficial velocity and the pressure drop for conventional bed

## 5.2. Tapered fluidized bed at ambient temperature

**Table 5:** Experimental data in Tapered fluidized bed at ambient temperature

Material	Particle Diameter ( $\mu\text{m}$ )	Bed Height (cm)	Experimental Pressure Drop (Pa)	Experimental $U_{mf}$ (m/sec)	Experimental $U_{mb}$ (m/sec)
Hematite	20	5.5	518.8	0.074	0.074
		10	598.6	0.129	0.129
		15	705.0	0.154	0.154



**Figure 4:** Plot of Pressure drop vs superficial velocity for bed height of 5.5, 10 & 15 cm using hematite

## 5. Results and Discussion

### 5.1 Conventional bed at ambient

#### Pressure Drop

The bed pressure drop decreases with the increase in gas velocity and with the increase in bed mass the pressure drop increases.

#### Minimum liquid fluidization velocity

The minimum liquid fluidization velocity decreases with the increase in gas velocity, and increases with the increase in particle size. Minimum liquid fluidization velocity is independent of initial static bed heights.

### 5.2 Tapered bed at ambient temperature

#### Bed Height and Minimum fluidization velocity ( $U_{mf}$ )

As can be seen in Table 7 the minimum fluidization velocity of the sample of hematite increased with increase in bed height.

#### Minimum bubbling velocity ( $U_{mb}$ )

The minimum fluidization velocity and bubbling velocity for the hematite powder sample was equal at the various bed heights that they were studied.

#### Fluidization index

A fluidization index value of 1 was observed indicates that the beds formed by the various powder samples have very less capacity to hold gases i.e. are less aeratable (Abrahamsen and Geldart, 1980)

## 6. Conclusions

The experiments were carried out at using a conventional cylindrical bed and a bed having a constant tapered angle by varying the bed heights for a variety of gas-solid systems. The effect of initial bed height of powder samples on minimum fluidization velocity and bubbling velocity were studied. Previous researches on fluidization in tapered beds have stated that minimum fluidization velocity is independent of the tapered bed height. Our results using hematite in tapered beds didn't support this assertion

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