Review Article: Improvement of Fluidization Quality Using Promoter and Modified Design of Distributor–A Review

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ABSTRACT

Persistent efforts were made by various researchers to improve the quality of gas-solid fluidized bed by means of hindering the bubble formation, avoiding the coalescence of bubbles through bubble breakage, thereby exhibiting a meticulous bed expansion and fluctuation. To conquer these difficulties and to achieve a better fluidization quality use of distributor, turbulence and promoter has been recommended and emerged as an alternative and effective technique. The introduction of promoters performs well in smoothening the fluidization operation in terms of infringement of bubbles, and regulating the bubble size through inhibiting their growth. A better fluidization quality is achieved by various fluidization characteristics such as bed expansion, fluctuation, mixing index and minimum fluidization velocity. Therefore, various researchers investigated the usage of different types of promoters and distributors and recommended that by using a suitably designed promoter and distributor possess an improved fluidization quality with marginal fabrication cost, however the operational cost remains unchanged. Most of the literature studies are dealt with the qualitative effect of the usage of promoters and distributors in a gas-solid fluidized bed though investigation of various hydrodynamics characteristics viz. bed expansion ratio, bed fluctuation ratio, bed pressure drop ratio.

Introduction

Fluidization is a traditional gas-solid or liquid-solid contacting technique that has been utilized in the process industries for the last ten decades. As compared to fix the bed, fluidized bed technique has the inimitable benefit of an even, liquid-alike flow of solid particles, thereby allowing continuous and involuntarily-inhibited operation with an improvement of managing

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and hurried solid mixing [1]. The mixing of solids in-turn directs in achieving the isothermal conditions all through the bed of particles, which further improves the rate heat and mass transfer between the used fluid and solid particles, thereby eliminating the risk of overheating of heat-sensitive materials [2].

Although the fluid-solid fluidized bed techniques possess many advantages, the integral shortcomings viz. channelling, abandoned bed expansion and fluctuation restrict the applications of gas-solid fluidized bed [3]. The shortcomings may be owing to the creation of bubbles and their consequent action in terms of collapsing and slugging [4]. The inherent shortcomings, thereby influencing the fluidization quality and moreover reduce the rate of heat transfer synergistically. Upon increasing the gas velocity beyond its minimum fluidization conditions, the fluctuation of particles at the top surface of bed causes instability in the fluidization operation. Moreover, while dealing with the fine or sticky particles, the nature of particles stimulates the cohesive force, thereby forming channels and thus the gas rather than dispersing through interstices in the bed of particles, it escapes through the formed channels [5]. Furthermore, while dealing with large and heavy particles, the gas velocity need to be increased to attain the fluidization conditions, however, it tend to increase the turbulence and formation of spouts within the bed [5].

To overcome the above said problems lying with the gas-solid fluidized bed in the line of augmentation of the quality of fluidization, various remedial measures viz. impartment of vibration and rotation of the column and unification of promoters such as orifices, internals and inserts within the bed of particles [6]. Moreover, design and usage of suitable gas distributor can be an alternative for the enhancement of the fluidization quality [7]. Shukrie *et al*. [8] critically reviewed on the investigation associated with the various designs of distributor and provided with critical findings as the foremost factor which influences the intensity of mixing. Further, they predicted that the various factor viz. flow area through the distributor, bed aspect ratio and pressure drop

have synergistically effect on the performance of the distributor at higher temperature. The low perforated ratio in the distributor results in less pressure drop and further the fluidized bed becomes steadier. However, as the bubbles of smaller size after escaping from the distributor may tend to coalesce and thus increases their size, which in-turn the quality of fluidization degrades.

The introduction of vibration in the fluidized bed improves the fluidization quality of cohesive particles. Various investigators have proposed that for large particles variation of pressure drop with respect to superficial velocity of gas with vibration is completely different from the conventional non- vibrational fluidized bed. Moreover, the minimum fluidization can't be obtained from the relationship for the vibrational mode [9,10]. The introduction of vertical vibration mode in fluidized bed of cohesive particles, the minimum fluidization velocity and the minimum porosity tend to decrease as compared to conventional mode [11,12]. Mori *et al*. [13] had introduced the twist vibrational mode in the fluidized bed of cohesive particles, and further classified the particles into three sub-categories. However, the complexity nature of the twist vibrational mode makes it tough to understand the vibration propagation, thereby resulting in less investigation. Thus, introduction of promoters reasonably an effective method for achieving the quality of fluidization.

The introduction of promoters in the fluidized bed delays the formation of bubbles, disrupts the formed bubbles, and minimizes the formation of slug, and thus decrease the bed expansion and fluctuation, which in turn increases the quality of fluidization [14]. Moreover, the lateral dispersion and the quality of fluidization tend to improve owing to the swirling motion developed through the inclined injection of the gas promoters [15]. Furthermore, from the experimental result on particle mixing Wang and Huang [13] projected about if the orifice diameter of the distributor is small then the performance is better. Thus, various researchers [16-24] have recommended the use of promoters.

Bed dynamics with bed internals

The enhancement in the quality of fluidization can be achieved through the introduction of the various types of bed internals viz. distributors, promoters, nozzles, slotted, horizontal and vertical orifices. The introduction of orifices exhibits a low pressure drop with an additional advantage of enhanced rate of heat and mass transfer through better physical contact between the gas and liquid [25]. Moreover, an appropriate promoter design along with the gas distributor can reciprocate the quality of fluidization through reduction in bubble size, restricting their growth and further minimizing the slugging [26]. Therefore, the fluidization quality in terms of bed expansion and bed fluctuation can be optimized through the introduction of various configuration of promoters and distributors.

Bed dynamics with orifices/Promoters

The use of an appropriate promoter, being an effective method for improving the fluidization quality, augments the fluidization quality through minimizing the bubble, channel, and slug formation. Specifically, the promoters control the fluctuation and expansion through restricting the bubble formation, its growth, size of the bubble. **Figure 1** shows a schematic diagram of fluidized bed consisting of bed internals. The two categories of promoters such as rod and disc type used by various researchers for investigations are as strained in **Figure 2(a) and 2(b)**.
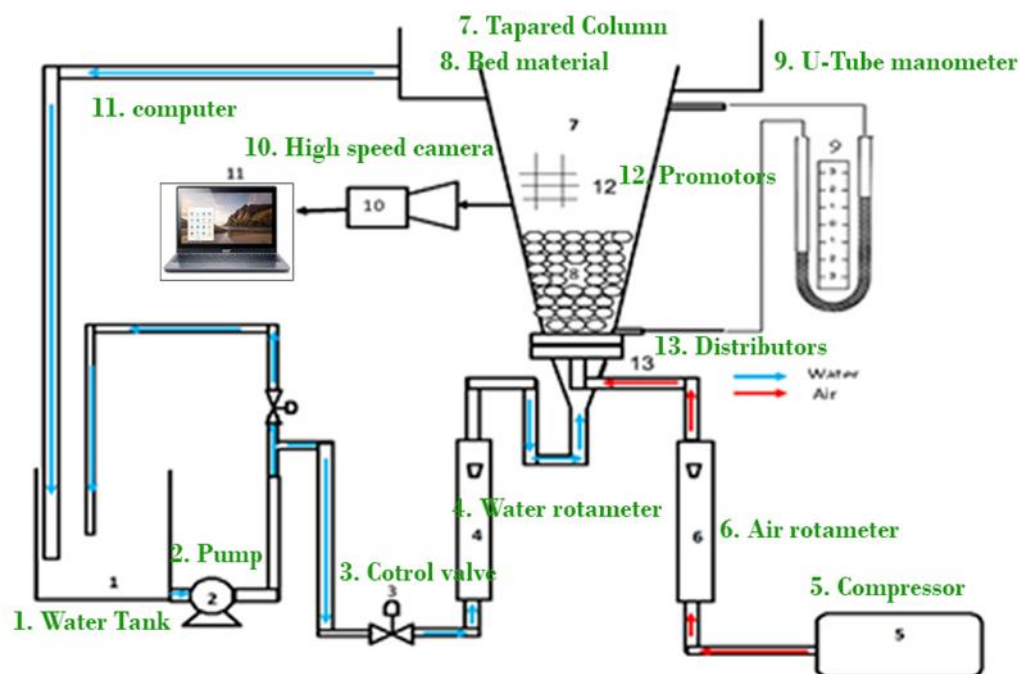
Figure 1. Schematic diagram of experimental set-up

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normal

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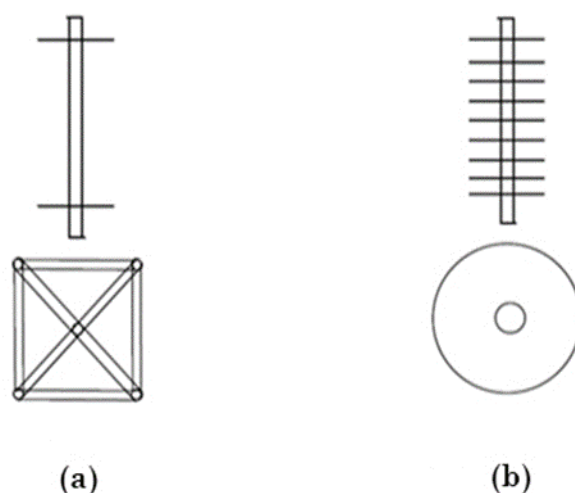
Figure 2. (a) Promoter rod type; (b) Promoter disc tipe

Bed behaviour qualitative

Based on the use of catalyst, generally two categories of fluidized bed reactors viz. catalytic and non-catalytic have been used in the chemical industries. The catalytic fluidized bed reactor, the catalytic particles (solid phase) doesn't take part in the chemical reaction, while the non-catalytic one the solid phase involve in the chemical reaction [27]. In both, to maintain the enviable operating parameters, heat transfer plays a significant role so that the process rate of reaction can be controlled. Moreover, the temperature of the process needs to be controlled at its desired value to alleviate the formation of the desired product, yield and efficiency. Thus, various categories of submerged surfaces viz. plates, tubes, and baffles, along with their configurations and orientation methods of the reactors viz. vertical and horizontal were emerged and were used by different investigators [28-31].

Simulation techniques were used for the investigation of the hydrodynamic characteristics of a gas-solid fluidized bed with and without internals [32]. It was concluded that the used of horizontal submerged internals act as a heat exchanging surface, thereby absorbing the heat of reaction, which in-turn helps in maintaining the required operating parameters of the bubbling fluidized bed reactors. Various researchers have investigated the bubble hydrodynamics viz. size of the bubble, bubble rise velocity, gas holdup and bubble frequency using different sized vertical internals with the consideration of the tube to tube spacing and tube arrangements viz. square and triangular [26,33]. They concluded that the use of the vertical internals decreases the bubble size, augments the bubble frequency and thereby affecting the bubble rise velocity.

Kumar and Roy [34] compared the promoted and un-promoted beds based on the minimum bubbling velocities and found that formation of a bubble is delayed owing to the augmented peripheral contacts of the bed with the fluid. Further, forecast with correlation subjected with experimental data about the bed expansion ratio using dimensional analysis approach and ANN-models. The ANN method had shown a precisely good agreement with the system behaviour than the dimensional analysis approach. The model equation for the bed fluctuation ratio by using the statistical approach method is tinted as Equation 1 and concluded as using rod type promoter bed fluctuation is greatly reduced which in turn reduce the fluidizer size hence the process becomes more economical.

$$(1)$$

Bed fluctuation ratio (r)= 1.668 + 0.309 X1+ 0.173 X2 -0.114 X3+ 0.112 X4 + 0.079 X1 X2

Where

$$X_1 = flow\ Parameter = \left(\frac{G_f - G_{mf}}{G_{mf}}\right) \quad X_2$$
$$= Density\ Parameter = \left(\frac{\rho_s}{\rho_f}\right)$$

$$X_3 = Bed\ Height\ Parameter = \left(\frac{h_s}{D_c}\right) \quad X_4$$
$$= Promoter\ Parameter = \left(\frac{D_e}{D_c}\right)$$

In industrial processes, apart from the quality control and superior solid handling efficiency, the bed internals also possess significant contributions in the fluidization operations thereby altering the hydrodynamic characteristics through transforming the flow arrangements of the gas-solid. Thus, Law et al [35] studied the effects of the bed internals on transforming the flow pattern in the gas-solid fluidized beds. They observed that the bubble growth is reduced, and further the coalescence of the bubble minimizes, which in-turn reduces the bubble growth. Moreover, the bubbles of lesser size can weaken the solid entrainment along the bubble, thereby smoothing the fluidization, which in-turn increasing the interphase heat and mass transfer. Furthermore, the hydrodynamic characteristics viz. pressure drop, slugging, solid elutriation, and fluctuation at the upper surface of the bed were found to decrease through the use of bed internals. The bed internals contribute to effectively breaking the formed bubbles, thereby, making the smaller bubbles, in-turn reduction in the bed pressure drop [36].

Bed behaviour quantitative

The quality of fluidization can be extrapolated through the bubble formation and their growth in the flow direction. The alteration of gas flow endorses small size bubbles rise and travel in the fluidized bed, effectively drive the solid particles resulting in reduced bed expansion and fluctuation, intensive particle mixing and enhanced overall heat transfer coefficient [37]. The imperative investigations on hydrodynamics characteristics such as bed pressure drop, bed expansion, bed fluctuation, bubbling, channeling, and slugging which drives the fluidization quality in conventional and promoted beds are discussed in the following section.

Prediction of bed expansion ratio

It is the ratio of the average of the highest and lowest height of expanded fluidized bed to the initial static bed height [38]. For a fixed fluidized bed height, it is an important parameter for a particular service. The synergistic effect of both the introduction the promoter and gas distributor on the bed expansion is depicted as here under.

It was observed that the expansion ratio is dependent on the various the various system parameters viz. excess gas velocity (Gf-Gmf), average particle size (dp) and the initial static bed height (hs) [39]. Though the bed expansion ratio is sufficiently more in a 2-D bed than in a 3-D bed, thus various methods were adopted by the different researchers for the measurement of the bed expansion ratio. The different methods for measuring the bed expansion ratio focuses on radiation absorption methods like X-ray absorption, g-ray absorption, positron emission tomography, electrical capacitance methods digital image analysis [40].

Various works related to prediction of un-promoted beds were attempted [41-46]. Simultaneously, substantial work [35,47-52] were carried out using promoted beds on different hydrodynamic features viz. enhancement of homogeneity in fluidized bed, motion of the particles, the phenomenon of the bubbles, mixing of fluid-solid element, minimum fluidization velocity and pressure drop etc. For improvement of fluidization quality regarding bed expansion less quantitative information is available.

The synergistic effect of promoter and distributor on the bed expansion of the fluidized bed was investigated by Kumar and Roy [34]. In their investigation they had studied the effect of various system parameters on the bed expansion using varied rod promoters, disk promoters, and distributors. Further they observed that compared to un-promoted bed and bed expansion reduces quite effectively by using promoted bed. For similar system parameters, the bed expansion is directly

proportional to the gas velocity [53]. Though the bed expansion is notably prejudiced by the varied distributor and promoter designs and additional system parameters, thus usage of all categories of promoters is helpful in plummeting the bed expansion over the un-promoted ones. Moreover, from the investigations it was concluded that the categories of disk and blade promoters are more competent in the reduction of bed expansion than the conventional un-promoted ones [34]. Furthermore, the authors have depicted that the blade type promoter were observed perform better than the disk promoters in the improvement of the performance of fluidized bed in the line of reduction of bed expansion. The reduction in bed expansion was owing to the usage of lessen open area of the distributor. Moreover, the promoters and the distributor are manifest for the almost inclusive fluidization regimes.

However, the region which did not fully stabilize i.e. under minimum fluidization conditions was exempted. The different correlations developed (**Table 1**) for the bed expansions by using different combinations of Bed configurations were predicted by Kumar and Roy [34]. The calculated values of bed expansion ratio from various correlations presented in **Table 1** for both un-promoted and promoted bed were found to be in concurrence with the experimental values. Moreover, the radial promoter possesses superior fluidization quality owing to the disappearance of the channelling and slugging. Furthermore, the large number of rods in the rod promoter instigate more blockage volume, thereby reducing the bed expansion. Whereas, in case disc promoter the thickness and diameter of disc prevents the slugging, and thus reduces the bed expansion.

Table 1. Bed expansion ratio (R) by using different type of bed

Different type of bed	Bed expansion ratio (R')
Un-promoted bed	$0.3(G_R)^{0.85}\left(\dfrac{\rho_S}{\rho_f}\right)^{0.59}\left(\dfrac{h_s}{D_c}\right)^{-0.32}\left(\dfrac{A_{do}}{A_c}\right)^{0.20}\left(\dfrac{d_p}{d_o}\right)^{0.40}$
Rod promoter bed	$0.18(G_R)^{0.74}\left(\dfrac{\rho_S}{\rho_f}\right)^{0.56}\left(\dfrac{A_{do}}{A_c}\right)^{0.19}\left(\dfrac{D_e}{D_c}\right)^{0.23}\left(\dfrac{d_p}{d_o}\right)^{0.29}\left(\dfrac{h_s}{D_c}\right)^{-0.40}$
Disk promoter bed	$0.08(G_R)^{0.75}\left(\dfrac{\rho_S}{\rho_f}\right)^{0.56}\left(\dfrac{A_{do}}{A_c}\right)^{0.19}\left(\dfrac{D_k}{D_c}\right)^{-0.48}\left(\dfrac{d_p}{d_o}\right)^{0.26}\left(\dfrac{h_s}{D_c}\right)^{-0.47}\left(\dfrac{t}{D_c}\right)^{-0.24}$
Blade promoter bed	$0.24(G_R)^{0.73}\left(\dfrac{\rho_S}{\rho_f}\right)^{0.51}\left(\dfrac{h_s}{D_c}\right)^{-0.71}\left(\dfrac{A_{do}}{A_c}\right)^{0.17}\left(\dfrac{d_p}{d_o}\right)^{0.22}$

They observed that the improvement the quality of fluidization in terms of bubble formation and slugging can be attained through the collective effort of the suitable design of the promoter and reduction in the open area of the distributor. Further, the improvement of quality of fluidization was driven though the prevention of significant reduction transport dis-engaging height.

Sahoo [54] investigated the quality of fluidization improves using stirrer. Introducing revolving effect of promoter inside a two-phase fluidized bed, particle mixing improves to a considerable extent as compared with an immobile promoter. They depicted that the

better particle- particle mixing was observed using revolving rod promoter rather than the revolving disc promoter. However, as the segregation of the particles in concern, the disc promoter was proven to be better than the rod promoter in both revolving and immobile conditions. Therefore, the optimization of the quality of fluidization process can be attained through the suitably designed revolving promoters. They further developed correlations (**Table 2**) for bed expansion ratios with respect to various system parameters using rod promoter, disc promoter under static and rotational conditions. The different correlations developed were observed to be significant over

the entire range of the system parameters investigated and also can be used for the design of gas-solid fluidized bed for their industrial application.

Table 2. Bed expansion ratio (R) by using different type of promoter

Different type of bed	Bed expansion ratio (R')
Un-promoted bed	$1.8\,E+05\left[\left(\dfrac{H_S}{D_C}\right)^{1.17}\left(\dfrac{d_p}{D_c}\right)^{-0.46}\left(\dfrac{\rho_S}{\rho_f}\right)^{-1.86}\left(\dfrac{U}{U_{mf}}\right)^{3.32}\right]$
Rod Promoter bed without stirring effect	$8.6\,E+02\left[\left(\dfrac{H_S}{D_C}\right)^{0.64}\left(\dfrac{d_p}{D_c}\right)^{-0.74}\left(\dfrac{\rho_S}{\rho_f}\right)^{-1.26}\left(\dfrac{U}{U_{mf}}\right)^{2.13}\right]$
Disk promoter bed without stirring effect	$1.29\,E+02\left[\left(\dfrac{H_S}{D_C}\right)^{0.14}\left(\dfrac{d_p}{D_c}\right)^{-0.71}\left(\dfrac{\rho_S}{\rho_f}\right)^{-0.93}\left(\dfrac{U}{U_{mf}}\right)^{2.06}\right]$
Rod Promoter bed with stirring effect	$3.19\,E+04\left[\left(\dfrac{H_S}{D_C}\right)^{0.10}\left(\dfrac{d_p}{D_c}\right)^{-0.66}\left(\dfrac{\rho_S}{\rho_f}\right)^{-2.26}\left(\dfrac{U}{U_{mf}}\right)^{2.80}\left(\dfrac{N_S}{N_f}\right)^{-7.02}\right]$
Disk promoter bed with stirring effect	$1599+04\left[\left(\dfrac{H_S}{D_C}\right)^{0.31}\left(\dfrac{d_p}{D_c}\right)^{-0.76}\left(\dfrac{\rho_S}{\rho_f}\right)^{-1.68}\left(\dfrac{U}{U_{mf}}\right)^{3.50}\left(\dfrac{N_S}{N_f}\right)^{-4.43}\right]$

Prediction of bed fluctuation ratio

The gas flow through the bed of particles causes prevalence of bubbles and when the gas mass velocity is more than the velocity at incipient fluidization, stimulates the fluctuation in the upper surface of bed, thereby leads to instability in the operation and adversely affecting the fluidization quality [55]. The superiority of fluidization can be quantified by two methods viz. uniformity index method and the fluctuation ratio method [56]. However, the later one is more exact approximation of the quality of fluidization. The fluctuation ratio is the ratio of the highest to the lowest height attained during fluidization condition [38].

The correlation developed for bed fluctuation ratio by Leva [57] is depicted as Equation 2.

$$(2)\qquad r=e^{m\frac{G_f-G_{mf}}{G_{mf}}}$$

Where, m, being the slope and function of particle size, also influenced by slugging. They observed that the non-slugging zone fluctuation ratio is followed slickly by the slugging zone. Though the slugging in the bed is dependent upon the aspect ratio, thus it is well understood that the fluctuation ratio is also dependent on the aspect ratio [57]. Various investigators had developed correlations for bed fluctuation ratio with respect to static and dynamic system variables for different beds viz. cylindrical [58,59], cylindrical beds with orifice promoted [14] and conical conduits [60,61].

The developed correlation by Krishnamurthy et al [62] for the bed fluctuation ratio using horizontal orifices in a gas-solid fluidized bed is as depicted as Equation 3.

$$(3)$$

$$r=\frac{59}{100}\left(\frac{G_f}{G_{mf}}\right)^{\frac{101}{100}}\left(\frac{D_P}{D_c}\right)^{-\frac{12}{100}}\left(\frac{D_c}{H_s}\right)^{-\frac{20}{100}}\left(\frac{\rho_s}{\rho_f}\right)^{-\frac{2}{100}}$$

The effect of stirrer type orifices on fluidization quality and proposed a correlation for the prediction of fluctuation ratio by Agrawal *et al.* [63].

$$(4)$$

$$R=\frac{249}{100}\left(\frac{G_f}{G_{mf}}\right)^{\frac{175}{100}}\left(\frac{D_P}{D_c}\right)^{-\frac{7}{100}}\left(\frac{D_c}{H_s}\right)^{-\frac{29}{100}}\left(\frac{\rho_s}{\rho_f}\right)^{-\frac{25}{100}}$$

Further, correlations were developed by Kumar [53] for bed fluctuation ratio ensuing the null-fluctuation at the incipient fluidization as td in **Table 3**.

Table 3: Fluctuation ratio (r) for different type of bed

Different type of bed	Bed fluctuation ratio (r)
Un-promoted bed	$\dfrac{24}{100}(G_R)^{\frac{85}{100}}\left(\dfrac{\rho_s}{\rho_f}\right)^{\frac{63}{100}}\left(\dfrac{h_s}{D_c}\right)^{-\frac{36}{100}}\left(\dfrac{A_{do}}{A_c}\right)^{\frac{21}{100}}\left(\dfrac{d_p}{d_o}\right)^{\frac{41}{100}}$
Rod Promoter bed	$\dfrac{16}{100}(G_R)^{0.69}\left(\dfrac{\rho_s}{\rho_f}\right)^{0.50}\left(\dfrac{h_s}{D_c}\right)^{-0.38}\left(\dfrac{D_e}{D_c}\right)^{0.27}\left(\dfrac{A_{do}}{A_c}\right)^{0.13}\left(\dfrac{d_p}{d_o}\right)^{0.25}$
Disk promoter bed	$\dfrac{9}{100}(G_R)^{0.67}\left(\dfrac{\rho_s}{\rho_f}\right)^{0.46}\left(\dfrac{A_{do}}{A_c}\right)^{0.14}\left(\dfrac{t}{D_c}\right)^{-0.23}\left(\dfrac{D_k}{D_c}\right)^{-0.67}\left(\dfrac{d_p}{d_o}\right)^{0.24}\left(\dfrac{h_s}{D_c}\right)^{-0.46}\left(\dfrac{h_s}{D_c}\right)^{-0.46}$
Blade promoter bed	$\dfrac{29}{100}(G_R)^{0.31}\left(\dfrac{\rho_s}{\rho_f}\right)^{0.33}\left(\dfrac{d_p}{d_o}\right)^{0.18}\left(\dfrac{h_s}{D_c}\right)^{-0.67}\left(\dfrac{A_{do}}{A_c}\right)^{0.13}$

Kar [64] developed the correlations for bed fluctuation ratio(r) for the gas-solid system and is presented as **Table 4**.

Table 4: Fluctuation ratio (r) for different type of bed developed by Kar

Different type of bed	Bed Fluctuation Ratio (r)
Un-promoted bed	$\dfrac{3}{1000}\left(\dfrac{H_s}{D_c}\right)^{0.11}\left(\dfrac{\rho_s}{\rho_f}\right)^{0.18}\left(\dfrac{D_p}{D_c}\right)^{-0.05}\left[\dfrac{(G_f-G_{mf})}{G_{mf}}\right]^{0.35}$
co-axial rod type promoter bed	$\dfrac{4}{1000}\left(\dfrac{H_s}{D_c}\right)^{0.15}\left(\dfrac{\rho_s}{\rho_f}\right)^{0.29}\left(\dfrac{D_p}{D_c}\right)^{-0.29}\left[\dfrac{(G_f-G_{mf})}{G_{mf}}\right]^{0.30}$
co-axial disk type promoter bed	$\dfrac{87}{100}\left(\dfrac{H_s}{D_c}\right)^{0.04}\left(\dfrac{\rho_s}{\rho_f}\right)^{0.02}\left(\dfrac{D_p}{D_c}\right)^{-0.04}\left[\dfrac{(G_f-G_{mf})}{G_{mf}}\right]^{0.04}$

According to Sahoo [54] fluctuation in a gas fluidized bed is unavoidable. As reported, with increase in the average particle size and density of the bed material the bed fluctuation ratio tends to decrease in both un-promoted and promoted beds. This may be owing to increase in the weight of the bed material causes a decrease in the fluidization capacity at a constant superficial gas velocity. The Fluctuation ratio (r) for different Type of bed developed by Sahoo is presented as **Table 5**.

Table 5: Fluctuation ratio (r) for different type of bed developed by Sahoo

Different type of bed	Bed fluctuation ratio (r)
Un-promoted bed	$2.4\,E+03\left[\left(\dfrac{H_S}{D_C}\right)^{1.32}\left(\dfrac{d_p}{D_c}\right)^{-1.30}\left(\dfrac{\rho_S}{\rho_f}\right)^{-1.74}\left(\dfrac{U}{U_{mf}}\right)^{3.04}\right]$
Rod promoter bed without stirring effect	$33.155+02\left[\left(\dfrac{H_S}{D_C}\right)^{0.63}\left(\dfrac{d_p}{D_c}\right)^{-0.60}\left(\dfrac{\rho_S}{\rho_f}\right)^{-0.75}\left(\dfrac{U}{U_{mf}}\right)^{1.27}\right]$
Disk promoter bed without stirring effect	$3.49\,E+02\left[\left(\dfrac{H_S}{D_c}\right)^{0.29}\left(\dfrac{d_p}{D_c}\right)^{-0.84}\left(\dfrac{\rho_S}{\rho_f}\right)^{-1.11}\left(\dfrac{U}{U_{mf}}\right)^{0.89}\right]$
Rod promoter bed with stirring effect	$3E+07\left[\left(\dfrac{H_S}{D_C}\right)^{0.38}\left(\dfrac{d_p}{D_c}\right)^{-0.67}\left(\dfrac{\rho_S}{\rho_f}\right)^{-2.87}\left(\dfrac{U}{U_{mf}}\right)^{0.51}\left(\dfrac{N_S}{N_f}\right)^{-4.98}\right]$

Disk promoter bed with stirring effect	$4.8E + 03 \left[\left(\dfrac{H_S}{D_C}\right)^{0.32} \left(\dfrac{d_p}{D_c}\right)^{-0.29} \left(\dfrac{\rho_S}{\rho_f}\right)^{-1.92} \left(\dfrac{U}{U_{mf}}\right)^{3.76} \left(\dfrac{N_S}{N_f}\right)^{-5.9}\right]$

The investigators had focused on the development of correlations for bed fluctuation ratio and the qualitative analysis related to fluidization quality for different conduits viz. cylindrical, non-cylindrical, conical and some of the promoted beds. However, the detailed investigations on the effect of various system parameters using different promoted beds stay uncharted. Moreover, the following broad conclusions may be drawn from the above investigations.

1.The introduction of holed (promoted) in cylindrical and conical conduits contributes towards a significant decrease in the bed fluctuation ratio.

2.Further, the bed fluctuation ratio tends to decrease for binary mixture of homogeneous and heterogeneous particles as compared to the single-sized particles under similar fluidization conditions.

Moreover, the distributor design, usage of different types of promotes and the different system variables have a synergistic effect on the bed fluctuation ratio. Furthermore, a suitable promoter and less open area of the distributor results in shortening the bubble development and slugging, thereby reducing the fluctuation of the particles and in-turn helps in achieving better fluidization quality.

Prediction of pressure drop

The variation of local pressure drop is owing to the formation of the channels in the fluidized bed, specifically at incipient fluidization conditions and just above the minimum fluidization conditions the channelling depends on the stable conditions [25]. The sufficiently charge pressure drop across the distributor causes the flow through the distributor undistributed across the cross-section in the bed, thereby ensuring the uniform fluidization with stable operation [65]. Moreover, the increase in local velocity results in an increase in the total pressure drop across both the distributor and the bed, which further hinder the channel formation [66]. Thus, for the design of the multi orifice distributor, the ratio of the distributor and the total pressure drop becomes a vital criterion.

Hiby [67] reported that, near to the minimum fluidization conditions, to attain a stable fluidization condition, a minimum of 30% of the total bed pressure need to be attained through the distributor. Further, he concluded with the facts that the desired ratio of distributor-to-bed pressure drop increases soberly with particle size. Qureshi and Creasy [68] in his extensive review found that for the multi-orifice distributor plate the pressure drop ratios was found to lie between 0.11 to 1.0. Moreover, to attain a stability of a fluidized bed (0.14 to 0.22), a smallest ratio is desirable for the range of range of Galileo number 1 to 104 [69]. Further, the modified value for this is suggested as 0.21 [70]. Moreover, the pressure drop ratio will be in the range of 0.05 to 1.0 as suggested by Siegel [69] and Whitehead and Dent [71]. Furthermore, for coarse and fine particles the ratio of distributor to bed pressure drop were obtained as 0.24 and 0.12 respectively [72].

For conventional gas-solid fluidized bed, pressure drop Equation 5 as given below

$$(5) \qquad \frac{\Delta p_b}{L} = \rho_s(1 - \epsilon)g$$

Padhi *et al.* [73] created correlations for the case of various configurations and shapes of immersed surfaces inside gas-solid fluidized beds. These correlations have also been predicted in gas-solid fluidized beds with varying design parameters, operating conditions, and gas physical properties.

$$(6)$$

$$\Delta p = 956.07 \left(\frac{h_s}{D}\right)^{0.876} (y)^{0.876} \left(\frac{G_f}{G_{mf}}\right)^{-1.5049}$$

Where y is the twist ratio of the internals

A correlation for bed pressure drop was developed by Mathew et al [25] using vertical internals and depicted as Equation 6.

(7)

$$\frac{\Delta p}{\rho_f u^2} = 6762.3 \, Fr^{-0.96} \left(\frac{D_C}{d_p}\right)^{-0.12} \left(\frac{h_s}{D}\right)^{1.24} \left(\frac{w}{s}\right)^{-0.14}$$

Where w is internal width and s is the internal spacing

Dimensional analysis approach was used by Taofeeq [74] for the development of correlation for bed pressure drop (Equation 7).

(8)

$$\frac{\Delta p}{\rho_g u^2} = 357.9523 \, (Fr)^{-1.0229}(Ar)^{0.7782} \left(\frac{D_i}{D_c}\right)^{-0.0802}$$

Moreover, Kumar [53] developed the correlations for distributor-to-bed pressure drop ratioby using dimensional analysis and predicted the following correlations for different type of bed which is summarized as **Table 6**.

Table 6: Distributor to bed pressure drop ratio for different type of bed

Different type of bed	Distributor to bed pressure drop ratio ($\frac{\Delta p_d}{\Delta p_b}$)
Un-promoted bed	$129 \times 10^{-2}(G_{mf})^{1.14}\left(\frac{\rho_s}{\rho_f}\right)^{0.48}\left(\frac{h_s}{D_c}\right)^{-1.02}\left(\frac{A_{do}}{A_c}\right)^{-1.83}\left(\frac{d_p}{d_o}\right)^{0.89}$
Rod Promoter bed	$866 \times 10^{-3}(G_{mrf})^{1.26}\left(\frac{\rho_s}{\rho_f}\right)^{0.53}\left(\frac{D_e}{D_c}\right)^{-0.21}\left(\frac{A_{do}}{A_c}\right)^{-2.01}\left(\frac{d_p}{d_o}\right)^{0.94}\left(\frac{h_s}{D_c}\right)^{-1.18}$
Disk promoter bed	$941 \times 10^{-3}(G_{mrf})^{1.16}\left(\frac{\rho_s}{\rho_f}\right)^{0.51}\left(\frac{t}{D_c}\right)^{-0.12}\left(\frac{D_k}{D_c}\right)^{0.22}\left(\frac{A_{do}}{A_c}\right)^{-1.9}\left(\frac{d_p}{d_o}\right)^{0.93}\left(\frac{h_s}{D_c}\right)^{-1.06}$
Blade promoter bed	$131 \times 10^{-2}(G_{mrf})^{1.17}\left(\frac{\rho_s}{\rho_f}\right)^{0.48}\left(\frac{d_p}{d_o}\right)^{0.92}\left(\frac{h_s}{D_c}\right)^{-1.04}\left(\frac{A_{do}}{A_c}\right)^{-1.87}$

Furthermore, the various system parameters of un-promoted and promoted batch gas-solid fluidized bed were correlated to predict the bed pressure drop ratio [64] and represented as **Table 7**.

Table 7: Pressure drop ratio for different type of bed

| Different type of bed	Pressure drop ratio ($\frac{\Delta p}{\rho_f u^2}$)
Un-promoted bed	$\frac{124}{10000}\left(\frac{H_s}{D_c}\right)^{2.8}\left(\frac{H_e}{D_c}\right)^{-1.51}\left(\frac{\rho_s}{\rho_f}\right)^{0.66}\left(\frac{D_p}{D_c}\right)^{-1.45}$
co-axial rod type promoter bed	$\frac{3}{100}\left(\frac{H_s}{D_c}\right)^{2.19}\left(\frac{H_e}{D_c}\right)^{-2.15}\left(\frac{\rho_s}{\rho_f}\right)^{1.02}\left(\frac{D_p}{D_c}\right)^{-1.3}$
co-axial disk type promoter bed	$\frac{1842}{1000}\left(\frac{H_s}{D_c}\right)^{1.86}\left(\frac{H_e}{D_c}\right)^{-0.93}\left(\frac{\rho_s}{\rho_f}\right)^{0.34}\left(\frac{D_p}{D_c}\right)^{-0.57}$

Bed dynamics with distributors

The alteration in the distributor design in the fluidization is an important parameter for the improvement of the fluidization quality. Thus, various investigators had pointed out the contribution of the distributors in the direction of enhancement of fluidization quality and their wide range applicability in the gas-solid fluidized bed. The usage of distributor causes a transverse distribution of the fluidizing medium at the base of the bed, thereby resulting in a uniform distribution through the entire cross-section of the bed. Further, the distributor also contributes towards the supporting of the de-fluidized bed, in-turn prevents the back-flow of the particles during downtime [75]. **Figure 3** summarizes the various types of distributor designs based on the direction of air at the entrance of distributor available in the literature. Similarly, **Figure 4** and **Figure 5** show the distributors having a different opening channel [54] and different type of distributor Design.
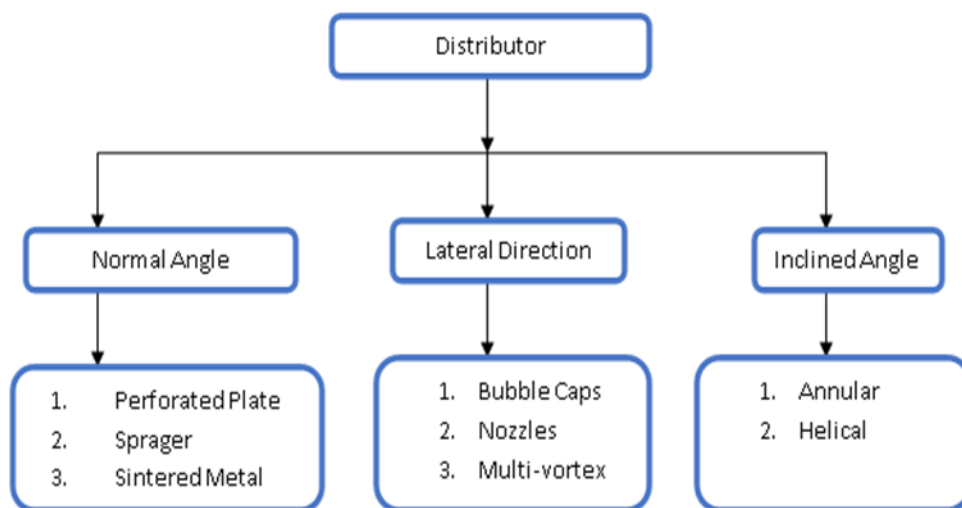
Figure 3. Primary distributor designs
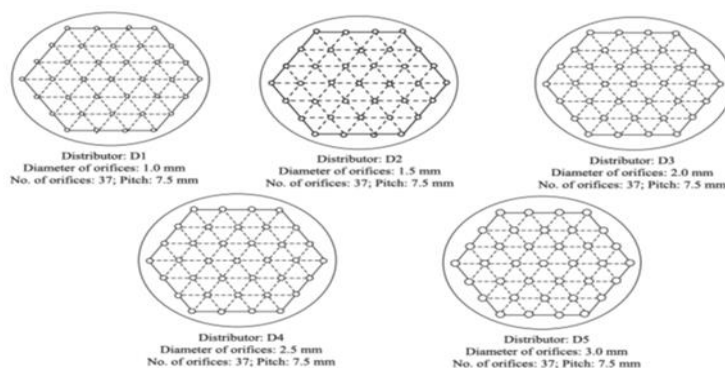
Figure 4. Distributors having a different opening channel (Source Awadhesh Kumar [53])
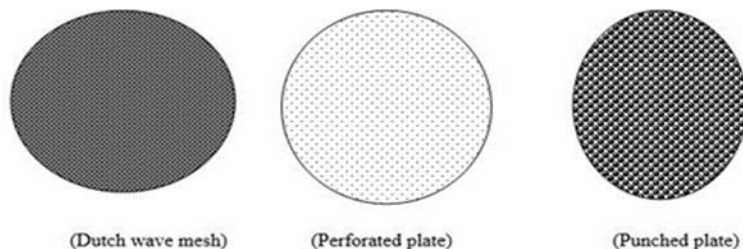
Figure 5. Different type distributor design

Bed behaviour qualitative

Ghose and Saha [76] demonstrated about the bubble formation which influence robustly by the type of distributor. The effect of distributors on a gas-solid fluidized bed has been further studied by Saxena *et al.* [70]. The design of distributor drives the distributor performance, thereby resulting in effective performance of the fluidized bed [8]. Thus, different distributors by varying the orifice diameters strewn in the annular and central zones with same open-area were used by Swain et al [77] and developed the following correlation for bed fluctuation ratio as Equation 9.

$$(9)$$

$$r = \frac{3316}{1000}\left(\frac{G_f}{G_{mf}}\right)^{0.60}\left(\frac{\rho_s}{\rho_f}\right)^{-0.23}\left(\frac{A_{do}}{A_c}\right)^{0.24}\left(\frac{d_p}{d_c}\right)^{-0.43}\left(\frac{h_s}{D_c}\right)^{-0.35}\left(\frac{A_A}{A_c}\right)^{-0.11}$$

The effect of the distributor area on the bed fluctuation ratio was studied by Kumar *et al.* [34] and predicted the correlation as depicted in Equation 10.

$$(10)$$

$$r = 1 + \frac{24}{100}G_R^{0.85}\left(\frac{\rho_s}{\rho_f}\right)^{0.63}\left(\frac{A_{do}}{A_c}\right)^{0.21}\left(\frac{d_p}{d_c}\right)^{0.41}\left(\frac{h_s}{D_c}\right)^{-0.36}$$

Further, aspect ratio along with the distributor type and operating velocity synergistically contribute towards quality of fluidization [78]. The effect of the usage of gas distributor on the segregation of the particles in a high-density fluidized bed was investigated by Luo et al [79]. They concluded that for the improvement in the performance of the fluidization, a higher pressure drop across the distributor is desirable. A well-built interface between the fluidized bed and air supply in the form of pressure waves were observed by Sasic *et al.* [80]. Moreover, they observed a clear interaction for low-air distributor pressure drop, whereas the fluctuation in pressure in the assembly is independent of the bed fluctuation for high air-distributor pressure drop. Mohanty *et al.* [81] had used distributor plates of varying open area (6%, 8%, 10% and 12%) and observed that a 10% open area in the distributor plate results in decreasing the bed pressure drop.

Zhong *et al.* [82] discovered that increasing the spouting gas velocity while keeping the fluidizing gas velocity constant results in a spouted bed characteristic, whereas increasing the fluidizing gas velocity while keeping the spouting gas velocity constant results in a fluidized bed characteristic.

Awadhesh Kumar *et al.* [14] had developed the model equation and state that if all the coefficient parmeters are positive then all the respective ratio's are increases.

$$(12)$$

Bed expansion ratio (R)= 1.795 + 0.389 X_1+ 0.141 X_2 -0.144 X_3+ 0.112 X_4+ 0.069 $X_1 X_2$ -0.070 $X_1 X_3$

$$(13)$$

Bed fluctuation ratio (r)= 1.618 + 0.285 X_1+ 0.094 X_2 -0.106 X_3+ 0.102 X_4+ 0.043 $X_1 X_2$-0.049 $X_1 X_3$

$$(14)$$

G_{mf} = 1824.75 + 1242.75 X_2 – 55.25 X_4 – 37.25X_2 X_4

Sobrino *et al.* [83] had developed a correlation for the pressure drop across the distributor using bubble cap and perforated gas distributors in a turbulent fluidized bed with group A particles is presented as Equation 15-17.

$$(15)$$

Pressure drop across the distributor = $\Delta p_d = \sum k \frac{1}{2}\rho V_{base}^2$

Perforated plate pressure drop

$$(16)$$

$$\Delta p_{p,d} = \frac{1}{2}\rho_g U^2 \left[K_{p,orifice}\left(\frac{A}{A_{p,orifice}}\right)^2 + K_{mesh}\left(\frac{A}{A_{mesh}}\right)^2 \right]$$

Bubble-cap distributor pressure drop

$$(17)$$

$$\Delta p_{c,d} = \frac{1}{2}\rho_g U^2 \left[K_{c,orifice}\left(\frac{A}{A_{c,orifice}}\right)^2 + K_{cap}\left(\frac{A}{A_{cap}}\right)^2 \right]$$

As observed by Ghaly *et al.* [84] during investigation on a pilot scale fluidized bed that with increase in static bed height the pressure drop ratio ($\Delta PD/\Delta PB$) decreases. Further they concluded that a >1.25Umf in the superficial gas velocity is desirable for achieving better mixing, enhanced fluidization quality and non-slugging bed.

A novel method of air distribution, introduction of a porous sponge as secondary distribution layer, was used by Fan *et al.* [85] for coal separation. They observed that such novel air distribution can cause a uniform distribution of air throughout the bed and the process becomes energy intensive. Moreover, the rate of coalescence of bubbles tend to decrease, thereby resulting in better separation of the particles and in-turn reduces the formation of partial channelling. In their investigation, they optimized the various system parameters viz. gas velocity, static bed height and time of separation using Response Surface Methodology (RSM) and developed a quadratic model for ash segregation degree (S) as Equation 18.

$$(18)$$
$$S = 2.88894 + 0.12552\,NH - 5.54564N^2 - 0.000013\,T^2 - 0.00073\,H^2$$

Conclusion

By reviewing the different author's investigation, the following conclusions were drawn.

1. The usage of suitably designed promoter and appropriate gas distributor tend to exhibit superior fluidization quality through inhibiting the bubble intensification, coalescence, their growth and thereby minimizing the slugging. These advantages resulted will lead to substantial diminution of the expanded bed height, thereby limiting the height of the column, which in-turn reduces the fabrication cost.

2. The various correlations developed for the cylindrical conduits remain un-applicable for non- columnar conduits. However, suitable modifications were made in the correlation so as to make it useful for the non-columnar conduits.

3. The distributor fabricated with lessen open area along with the introduction of promoter possess a synergistic effect on the quality of fluidization through limiting the TDH.

4. The vertical internals, being suitable aid for achieving better hydrodynamics characteristics in terms of enhanced gas-solid mixing. Moreover, this may lead to an amplified gas residence time, and thereby increasing interaction between gas and solid, which in-turn results in enhanced rate of heat and mass transfer.

5. The distributor with optimum cross-sectional area, limiting the static bed height and average particle size possess an improved fluidization quality.

6. The bubble behaviour in terms of infringement, limiting the size can be improved through the introduction of rod and disc type of promoters. Moreover, the rod type promoter, being more effective, contributes towards decreasing the bed fluctuation ratio and increasing the bed expansion ratio.

7. The rotation of the promoter using a stirrer exhibit a superior fluidization quality through achieving a better particle mixing, rather than the stationary promoter. Moreover, a rotating rod promoter is better suitable than the disc promoter when mixing is concern. However, the rotating or non-rotating disc promoter exhibits better for segregation.

8. During the usage of different promoter better mixing was obtained under higher air mass velocity, however, when segregation is concerned, lower air mass velocity need to be maintained.

Conflict of Interest

The authors declare that this article content has no conflict of interest.

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Nomenclature:

$A =$ Erguns constant

$A_n =$ Cross-sectional area of non-column bedm^2

$Ar =$ Archimedes number$= \frac{g dp_{sm}{}^3 \rho_{sm}{}^2}{\mu_f{}^2}$

$B =$ Erguns constant

$dp_{sm} =$ Average particle diameter of ternery mixture

$dh =$ Differential bed height , m

$dP =$ Pressure drop through bed height dh , N/m^2

$D_{0=}$ Bottom diameter of tapered bed , m

$D_c =$ Equivalant column diameter , m

$d_p =$ Particle diameter , m

$D_t =$ Diameter of column , m

$Fr =$ Frouds number

F=Force exerted by the fluidizing medium on particle in the bed , N

G= Effective weight of the material in the medium , N

$G_{mf} =$ Superficial fluid mass velocity at minimum fluidiation , Kg/hrm^2

$G_f =$ Flow rate of fluid at fluidization condition , $\frac{m^3}{hr}$

$G_{mf} =$ Flow rate of fluid at minimum fluidization condition , $\frac{m^3}{hr}$

g = Gravitational acceleration , 9.81 m/s^2

$g_c =$ Gravitational constant

H= Height of the bed , m

$h_{max} =$ Maximum height of fluidized bed, L

$h_{min} =$ Minimum height fluidization bed , L

$h_s =$ Initial static bed height , L

$H_f =$ Height of expanded bed, m

$H_s =$ Height of initial static bed, m

K = Constant

m = Correlation factor , dimensionless

r = Bed Fluctation ratio (dimensionless)

R = Bed expantion ratio (dimensionless)

$Re_c =$ Critical Reynolds number

U = Superficial fluid velocity , $\frac{m}{sec}$

$U_0 =$ Incipient fluidizing velocity , $\frac{m}{sec}$

$U_{vf} =$ Superficial velocity of gas where both components in a binary mixture are fluidized, m/sec

$U_{pc} =$ Minimum fluidization velocity of packed components in single component fluidized beds, m/sec

$U_f =$ Minimum fluidization velocity of fluid in single component fluidized beds , $\frac{m}{sec}$

$U_{mf} =$ Superfecial minimum fluidization velocity , $\frac{m}{s}$

$U_{mb} =$ Minimum bubbling velocity , ms^{-1}

$U_{ms} =$ Minimum slugging velocity , , ms^{-1}

$V_{mf} =$ Minimum fluidization velocity

$V_0 =$ Fixed bed the superficial velocity

$V_t =$ Particle terminal velocity

X_f: = Volume fraction of fines in the mixture

Greek word

$\Delta P_{mf} =$ Pressure drop at minimum fluidization velocity , $\frac{N}{m^2}$

$\Delta P =$ Pressure drop through particle bed , $\frac{N}{m^2}$

$\emptyset_s =$ Sphericity of particle

$\rho_p =$ Density of particle , $\frac{kg}{m^3}$

$\rho_f =$ Density of air , $\frac{kg}{m^3}$

$\varepsilon =$ Porosity of the fluidizing bed

$\epsilon_0 =$ Voidage at packed condition

$\mu =$ Viscosity of gas phase

$\rho_f =$ Density of fluid phase , $\frac{kg}{m^3}$

$\rho_s =$ Density of a particle , $\frac{kg}{m^3}$

$\emptyset_s =$ Sphericity of a particle

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