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Project report on

CHARACTERISTICS OF A BINARY MIXTURE IN A PROMOTED SOLID FLUIDIZED BED-A STATISTICAL APPROACH

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CERTIFICATE

This is to certify that Pooja malik and ratheesh gopal have completed their project work "CHARACTERISTICS OF BINARY MIXTURES IN A PROMOTED GAS-SOLID FLUIDIZED BED-A Statistical approach" under my guidance and supervision during the session 2004-2005. This is in partial fulfillment for the Award of BACHELOR IN TECHNOLOGY degree in CHEMICAL ENGINEERING under the Department of Chemical Engineering, National Institute of Technology, Rourkela.

Place:

Date:

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ACKNOWLEDGEMENT

We are greatly indebted tour project guide Prof. G. K. Roy, Head of the Chemical Engineering Dept. NIT, Rourkela for his king cooperation and guidance. Without his counsel and support this project work could not have been completed.

We are also thankful to Asst. Prof. S. K. Aggarwal, Chemical Engineering Department, NIT, Rourkela, for his kind co-operation throughout the project.

Place: Rourkela
Date: 21/01/2005

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INTRODUCTION

Objective:- To study the characteristics of a binary mixture in a promoted gas solid fluidized bed using both dimensional and statistical approach.

Fluidization is the operation by which fine solids are transformed into a fluid like state through contact with a gas or a liquid. Extensive use of fluidization began in the catalytic cracking reactors in the petroleum industry. The chief advantages of fluidization are that it ensures contact of the fluid with all parts of the solid particles, prevents segregation of the solid by thoroughly agitating the bed aand minimizes temperature variations even in a large reactor, again by virtue of the vigorous agitation some uses of fluidized beds are in

(1) chemical reactions

- (a) catalytic
- (b) non-catalytic

(2) Physical contacting

- (a) Heat transfet
- (b) Solids mixing
- (c) Drying
- (d) Size enlargement
- (e) Size reduction
- (f) Gas mixing
- (g) Classification
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- (i) Heat treatment
- (j) Coating

LITERATURE REVIEW

Interest in fluidized beds has continued unabated in recent years, spurred on by new applications for fine particles and an ever-growing role for circulating fluidized beds and other high velocity systems. A large number of research papers, reports and other literature continue to appear presenting advances in fluidization, new applications and improvements in technology.

HYDRODYNAMIC REGIMES AND TRANSITIONS

Some advances have been made in recent years in characterizing different flow regimes and in predicting the transitions between them. It is frequently useful to apply the analogy with gas-liquid two-phase flow in seeking to understand hydrodynamic regimes in gas-solid systems (Grace 1986).

Minirnum fiuidization

The lowest transition condition, the superficial velocity at minimum fluidization Umf, continues to be the subject of some investigation despite the considerable volume of previous work on the subject. Based on a balance of pressure drops required to support the weight minus buoyancy acting on the particles at the point of minimum fluidization and the well-known Ergun equation, most equations are of the form

$$Remf = \sqrt{(C_1^2 + C_2 Ar)} \qquad \text{----------} \qquad (1)$$

where Remf and Ar are the Reynolds and Archimedes numbers given by

$$Rent = \rho_g dp U_{mf} / \mu_g$$

$$Ar = \rho_g \Delta\rho g dp^3 / \mu_g$$

where Ap = (pp - PG). Here pp, PG, dp,/2 G and g refer to particle and gas density, particle diameter, gas viscosity and gravity respectively. New pairs of values of (G, C2) have been proposed (Lucas et al. 1986; Adanez & Abanades 1991) which are particle-shape dependent and species dependent, adding to the large number of pairs already in the literature as summarized by the latter authors.

Much higher values of gas flow than predicted by purely hydrodynamic approaches are required when the temperature is raised to a point where agglomeration begins to occur (Yamazaki et al. 1986; Davies et al. 1989)• Even when agglomeration is not a factor, great care is needed to measure U~r at high temperature, in particular to ensure temperature uniformity (Flamant et al. 1991)•

Minimum bubbling

Group A powders show an appreciable bubble-free velocity range between Umf and the minimum bubbling velocity Umb- While there continue to be contrary views, supported by instability theory based purely on hydrodynamic considerations (e.g. Foscolo & Gibilaro 1987; Gibilaro et al. 1988)

HYDRODYNAMICS OF GAS-SOLID FLUIDIZATION

Batchelor 1988; Foscolo 1989), most fluidization researchers believe that interparticle forces play an important role for group A powders throughout the bubble-free range and in determining Umb. Hence, complex rheological behaviour and non-hydrodynamic factors appear to play significant roles, making predictions difficult (Homsy et al. 1992). There is also further evidence (Jacob & Weimer 1987; Foscolo et al. 1989) that both Umb and emb, the voidage at the minimum bubbling condition, increase with increasing system pressure. The condition for distinguishing group A powders from group B particles, i.e. the maximum particle diameter for Umb to be appreciably greater than Umf, can be approximated (Grace 1986) by the empirical relation

$$dp < 101\{ \mu_g^2/\rho_g\Delta\rho g\}(\Delta\rho/ \rho_g)^{-0.425} \quad \text{----------} \quad (4)$$

Onset of turbulent fluidization

As interest in the higher velocity regimes of fluidization has blossomed, spurred by applications of turbulent and circulating fluidized beds, there has been increased investigation of the onset of these regimes. Since early work by Yerushalmi & Cankurt (1979), the transition to turbulent fluidization has usually been characterized by the superficial velocity Uc at which the amplitude of pressure fluctuations reaches a maximum, or by the superficial velocity Uk at which the amplitude of pressure fluctuations levels off with increasing superficial gas velocity U. As summarized by Brereton & Grace (1992), there has been wide variation in the manner in which the experimental pressure fluctuations have been measured and analysed, some workers preferring absolute or dimensional values, while others employ differential values and/or normalize to give dimensionless values. Uk is not a well defined parameter since it depends, among other factors, on the solids return system employed (Bi & Grace 1995), and so its use is not recommended. The method used to determine Uc significantly affects the result (Bi & Grace 1995). Values in the extensive literature based on absolute pressure

fluctuation data and bed expansion measurements are well represented by an equation due to Cai et al. (1989),

$Re_e = \rho_g dp U_c/\mu_g = 0.57Ar^{0.46}$ ---------- (5)

while differential pressure fluctuation data are well represented by

$Re_c = \rho_g dp U_c/\mu_g = 1.24Ar^{0.45}$ ---------- (6)

(Bi & Grace 1995). These equations improve on several expressions available in the literature of similar form. Since the differential pressure fluctuation data are more indictive of local conditions, [6] is recommended. Transition data based on visual observations tend to give smaller values of the transition velocity and to be highly subjective.
The transition to turbulent fluidization corresponds to breakdown of bubbling or slugging due to rapid coalescence and splitting beyond a certain point (Cai et al. 1990; Bi et al. 1996). Accordingly, the transition may be affected by such factors as the presence of baffles (Andersson et al. 1989) and particle size distribution (Sun & Grace 1992), which affect the growth and breakup of bubbles. Transition can occur quite sharply for group A powders where the transition is from bubbling to turbulent, or much more gradually for group B or D particles where slugging occurs first and the transition involves intermittent periods of slug-like and turbulent character (Brereton & Grace 1992). Caution is needed when studying the transition to avoid an apparent transition caused by gradual emptying of the interval between two pressure taps over which differential pressure fluctuation measurements are being taken (Rhodes & Geldart 1986a).

Transition to fast fluidization

Bi et al. (1995) showed that the transition from turbulent to fast fluidization corresponds to a critical superficial velocity Use which corresponds to the onset of significant particle entrainment Dimensionless regime diagram corresponding to Bi et al. (1995) extended from that of Grace (1986). From the riser. Except for columns of small size, Uso represents an equipment-independent property of the particulate material. Values of Use are well correlated by

$U_{se} = 1.53°5(gdp\Delta\rho/\rho_g)^{0.5}$ ---------- (7)

which makes use of > V_x (the terminal velocity) for group A and B particles (especially the former), while U_{se} is essentially equal to vv for group D particles. For U > Use, the flow pattern of circulating fluidized beds depends on the solids circulation rate as well as U.

INTERPARTICLE FORCES AND INFLUENCE OF PARTICLE SIZE DISTRIBUTION

There are continuing attempts to extend fluidization to finer and finer particles, into the range of cohesive particles found in group C of the Geldart classification. At the same time, more attention is being given to effects of interparticle forces for both group C and group A powders.

A useful review of cohesive forces affecting fluidization has been published by Visser (1989). In some cases, these cohesive forces may be counteracted sufficiently by external means to allow fluidization to proceed, e.g. by an acoustic field (Chirone et al. 1993). It has also been found that some fine powders will, beyond a certain minimum superficial velocity, spontaneously form agglomerates which are large and stable enough to fluidize like a group A powder (e.g. see Brooks & Fitzgerald 1986; Li et al. 1990). More work is required to understand this behaviour and to devise practical techniques for extending fluidization to finer particles. When bubbling fluidization does occur, Clift (1993) and Clift& Rafailidis (1993) have shown that interparticle stresses play a key role in bubble wakes while, for group A, B and D solids, interparticle stresses play a secondary role elsewhere and can often be neglected to a first approximation. In other words, the motion of bubbles, once formed, is insensitive to the rheological properties of the dense phase. Clift (1993) showed, on the other hand, that to explain the pre-bubbling differences between group A and B powders, one must consider interparticle forces, in particular the elasticity of the particulate phase which is critically dependent on particle-particle contacts. This is a very active area for future research. For many years industrial operators of catalytic fluidized bed reactors have recognized that it is important to maintain a significant proportion of 'fines', i.e. particles with diameters much smaller than the mean, within fluidized bed reactors in order to optimize reactor performance. The extent of the influence of fines and the underlying causes have become better understood in recent years.

Both Yates & Newton (1986) and Pell & Jordan (1988) demonstrated that addition of catalyst fines to fluidized bed reactors causes significant improvements in conversion, even when the added fines are catalytically inactive. In these two studies, the mean particle size, as well as the size distribution, were altered as fines were added. Sun & Grace (1990) performed experiments where the mean particle size [defined a s dp --- 1/Y~(xi/dpi), where xi = mass fraction of size dpi] was held constant, while three different size distributions were investigated. A wide size distribution always gave higher conversions (i.e. better gas-solid contacting) than a narrow size distribution, with a bimodal distribution showing intermediate results. In each of the cases above, the authors employed group A particles. Somewhat similar results were obtained by Kono & Soltani-Ahmadi (1990) by adding inert fines to

group B particles undergoing a gas-solid reaction. Addition of finer particles usually led to higher conversions, as well as lower pressure fluctuations, suggesting smaller bubbles. Geldart & Buczek (1989) found that bed expansion, entrainment and collapse test de-aeration times all increased significantly when the finest 3% by mass of particles (mean size 32/~m) of FCC powder was removed and replaced by small quantities of ultrafine particles of various types. The extent of the increase was greater when the size of the added ultrafine material was diminished. A reduction in bubble size appeared to be responsible for the observed influence. The improved performance of fluidized bed reactors with wide particle size distributions and appreciable quantities of fines is due to a number of factors (Grace & Sun 1991):
(a) In the bubbling regime, voids tend to be smaller with wide particle size distributions (Hatate et al. 1988; Geldart & Buczek 1989; Soltani-Ahmadi 1989; Sun & Grace 1992), probably associated with lower effective dense phase viscosities (Khoe et al. 1991).
(b) There are more particles dispersed inside the dilute phase when there are fines present (Sun & Grace 1990 and 1992), because fines spend more time inside voids when their terminal settling velocities are of similar magnitude to the relative through flow velocity of gas inside the void (Grace & Sun 1990).
(c) Wide size distributions trigger earlier transition to the turbulent fluidization regime where gas-solid contacting is better than in the bubbling regime. These findings are primarily for group A powders. Work is required to see whether they are also applicable to group B and D solids.

BUBBLING BED HYDRODYNAMICS

Much has been written about bubbling in fluidized beds over the last three and a half decades. A good understanding of the bubble hydrodynamics is necessary to understand bubble-related phenomena such as solids mixing and segregation, reaction conversion, heat and mass transfer, erosion of heat transfer tubes and particle entrainment in beds operated in the bubbling regime. In this review, recent studies related to bubbling or slugging fluidized beds are briefly discussed.
Bubble size, velocity, shapes and flow patterns are of key interest in bubbling hydrodynamics. These properties have been extensively measured experimentally by various methods. The experimental methods and findings have been summarized in several review articles (Davidson et al. 1985; Geldart 1986; Cheremisinoff 1986).

SOLIDS MIXING, SEGREGATION, PARTICLE MOTION AND EROSION

A good understanding of solids mixing behaviour is important in the design of physical and chemical processes in bubbling fluidized beds. Gas mixing, though related to solids mixing (Bellgardt et al. 1987), Solids mixing has been reviewed

by Potter (1971), van Deemter (1985), Fan et al. (1990) and Kunii & Levenspiel (1991). It is well recognized that solids mixing is directly related to bubble flow phenomena. The combined effects of gross circulation caused by drift and wake transport and small scale local-mixing in bubble wakes leads to favourable axial mixing. The extent of lateral solids mixing is much less favourable, particularly in shallow fluid beds (height to diameter ratio H/D < 0.25) where the influence of the axial wake transport is weakest. Particle segregation due to differences in particle size, density or shape, though commonly considered separately, can be treated as a subset of mixing processes with inclusion of preferential particle settling effects. In recent years, greater attention has been directed to the study of individual particle motion in fluid beds.

Solids mixing

A number of mathematical models have been proposed to predict mixing behaviour in bubbling fluidized beds. The need for a more realistic model based on relevant hydrodynamic parameters of bubbling beds was stressed by van Deemter (1985). A one-dimensional diffusion model has been demonstrated to be inadequate in describing axial mixing behaviour due to observed cycling in the concentration responses (de Groot 1967; Lim et al. 1993), especia!ly for large particle systems. Attempts have also been made to model axial solids mixing as a stochastic process (Fox & Fan 1987). However, lack of suitable experimental data for model verification has restricted the applicability of this approach. However, Fan et al. (1986) showed that lateral mixing can best be based on a stochastic diffusive model in which particle motion is characterized by both diffusive and convective components. The counter-current back-mixing (CCBM) model originally proposed by van Deemter (1961) and refined and generalized by Gwyn et al. (1970) has gained greater acceptance due to its good representation of the transport process in a bubbling bed. The model depicts the bed as a multiple phase system, with an upward flow of gas and wake phases and a downward flow in the dense phase. Exchange occurs between these phases. Mass balances over the individual phases were represented by a system of hyperbolic partial differential equations. Some of the main features of the CCBM model have been adopted by various researchers. For example, Sitnai (1981) used this approach to model solids mixing in a fluidized bed containing horizontal tubes by including a fastmoving, narrow downflow of solids at the wall. Numerical solutions through numerical inversion by Laplace transforms and a 'cinematic' approach have been provided by Lakshmanan & Potter (1990). This method was shown to be more computationally efficient and robust than the former for impulse and pulse inputs. Verification of the CCBM model has been impaired until recentlyby a lack of experim ental data on mixing and independent measurement of relevant bubble hydrodynamic parameters. Kozanoglu & Levy (1991) developed a CCBM model which further divides the

wake phase into four different compartments. Solids exchange between different phases is allowed except for the innermost wake phase which was taken to be stagnant. This multiple-layer wake region derived from an experimental/theoretical study by Kocatulum et al. (1992), where a strong particle velocity gradient in the wake was shown to exist. The nearly stagnant zone in the wake was believed to cause particles to be transported through large vertical distance. The solids exchange parameters between phases were adapted from the same study. The model provided a reasonable prediction of the axial concentration profile in comparison with experimental data.

Transient concentration responses measured by Lim et al. (1993) have been interpreted using a three-phase CCBM model similar to that of Gywn et al. (1970). Methods of predicting the wake exchange coefficient kw, available in the literature (Yoshida & Kunii 1968; Chiba & Kobayashi 1977), were found to be inadequate. The wake exchange, estimated from fitting to experimental data, was found to be weakly influenced by the minimum fluidization velocity, contrary to the proposed correlations. The solids exchange coefficient was found to be related to bubble size with kw = Aw/db, where Aw was of the order of 0.03 to 0.15 m/s for Umf from 0.068 to 0.35 m/s. This finding was further supported by Basesme & Levy (1992), where the wake exchange coefficient was measured in a two-dimensional bed using tracer displacement techniques and the data analysis method described by Chiba and Kobayashi (1977). The model of Chiba and Kobayashi (1977) agreed well with the experimental data over only a very narrow range of Umr and db. The model developed by Kocatulum et al. (1992) overpredicted the exchange coefficient by a factor of two to close to an order of magnitude, depending on the range of Umr and bubble size. The discrepancy was attributed to the theoretical analysis of Kocatulum et al. (1992) and the well-mixed assumptions in obtaining exchange coefficients from tracer measurements. The assumption of spherical or circular bubble wakes may also cause errors. The inability of the steady state model to account for the instabilities that lead to periodic wake shedding should also be taken into consideration. Hoffman et al. (1993) also pointed out that the proportionality of the solids exchange parameter to the minimum fluidization velocity did not hold. A better fit of the experimental data was realized when the dependency on Umf was dropped. The CCBM model is probably the best existing model to represent mixing in bubbling beds despite the fact that other possible mixing mechanisms, such as solids splashing at the bed surface and turbulent mixing near the distributor, are not accounted for. Although the solids convective component used in the model is well established, the key limitation which prevents application of this model with greater confidence is the absence of reliable solids exchange coefficients. The corresponding minimum fluidization velocities varied from 0.004 to 0.35 m/s, while almost all of the exchange coefficient values are within the range estimated by Lim et al. (1993). Quantitative understanding of solids mixing in the wake region remains poor; more work is clearly needed to develop a more reliable prediction for the exchange coefficient.

Promoted bed:-

The use of a suitable promoter and proper gas distributor can improve fluidization quality with better gas-solid contact through minimization of channeling and slugging and limit the size of bubbles and their growth. This results in ultimate reduction of bed fluctuation to a considerable extent and there by limiting the size of the equipment. A number of investigators have stressed the use of promoters to improve fluidization quality and to increase the range of applicability of gas-solid fluidized beds. Balakrishnan and Rao studied the effect of horizontal screen disc baffle on fluidized bed pressure drop and minimum fluidizing velocity. Horizontal baffles in reactors were used by Lewis, et al and Massimilla and Johnstone for the hydrogenation of ethylene and ozidation of ammonia, respectively. Kav, et al, carried out investigations with horizontal perforated disk on hydrogen chloride conversion and pressure fluctuations. Yong, et al, reported the effect of performance of gas-solid beds. Dutta and suciu investigated qualitatively the effect of perforated plate, wire mesh, angle iron grid and some other type of baffles in breaking bubbles in fluidized beds.

Kar and Roy used co-axial road and co-axial disk type promoters for their studies on fluidization quality and developed the following correlations for bed fluctuation ratio.

$$. r = 0.004(h_s/D_c)^{0.15}(dp/D_c)^{-0.29}(\rho_s/\rho_f)^{0.29(}(G_f - G_{mf})/G_{f.})^{0.3}$$

for bed with co-axial rod type promoter

$$. r = 0.87(h_s/D_c)0.04(dp/D_c)^{-0.04}(\rho_s/\rho_f)^{0.02}((G_f - G_{mf})/G_{f.})^{0.04}$$

for bed with co-axial disk promoter

PRESENT WORK:-

In the present work we are conducting experiments to study the hydrodynamic characteristics namely minimum fluidization velocity, pressure drop at minimum fluidization velocity, bed expansion ratio etc. In this study based on statistical design is made in order to bring out the interaction effects of variables, which would not be found otherwise by conventional experimentation and to explicitly find out the effects of each of the variables quantitatively on the response. In addition, the number of experiments required is far less compared to the conventional experiments. The variables affecting fluctuation ratio at minimum fluidization velocity are flow rate of fluid, static bed height

and equivalent diameter. Thus total number of experiments required at two levels for the three variables is 8 for response to fluctuation ratio. Each experiment is repeated 3 times and the average of three values is response value

SCOPE OF EXPERIMENT:-
(factorial design analysis)

Sr. Nos.	Name of the variable	Variable (general symbol)	Factorial (design symbol)	Minimum level (-1)	Maximum level (+1)	Magnitude of variable
1.	Flow rate of fluid (Kg/m^2hr)	$(G– G_{mf})/ G_{mf}$	X_1	0.25	0.75	0.25, 0.5, 0.75
2.	Static bed height (cm)	h_s	X_2	$8*10^{-2}$	$20*10^{-2}$	$(8, 12, 16, 20) *10^{-2}$
4.	Equivalent diameter (cm)	D_{eq}	X_3	$2.42*10^{-2}$	$3.988*10^{-2}$	$(2.42, 2.666, 3.183, 3.988)*10^{-2}$

EXPERIMENTAL SET UP:-

A schematic diagram of experimental setup is shown in the figure. The experiments have been conducted in pyrex glass column of 5.256 cm internal diameter and 3mm thick. The pressure drop across the bed is measured by manometer with CCl_4 as manometric liquid. A calming section is provided below the distributor plate for the uniform distribution of air compressed air has been used as the fluidizing medium. Four rod types promoters with 5, 9, 13 and 17 rods are used. Dolomite material of size -10 + 12 and -12 + 14 is studied exhaustively.

EXPERIMENTAL PROCEDURE:-

The binary mixture of dolomite of size -10 + 12 and -12 + 14 was used which was sieved and then mixed in the desired ratio. Then the test column was charged with required amount of dolomite of desired particle size depending upon the static bed height

For a particular run, data for bed pressure drop and expansion with varying flow rate have been noted and the same bave been repeated for different bed materials of varying particle size, initial bed height and promoters

THEORETICAL MODEL:-

The pressure drop equation for a packed bed of non-spherical particles is given by Ergun's equation i.e

$$\Delta P/L = \{150\, V_0 \mu (1-Є)^2\}/ (Ø_s^2 D_s^2 Є^3) + \{1.75 V_0^2 \rho (1-Є)\}/ Ø_s D_s Є^3\}$$

The first term is called Kozney-carmen equation and is applicable for flow through beds at particle Reynold's number upto 1.0. and the second term called burke-plummer equations applicable Re>1000.

$Ø_s$ is the sphericity defined as the surface-volume ratio for a sphere of diameter D_p divided by the surface volume ratio for the particle where nominal size is D_p.

$Ø_s = (6/D_p)/(S_p/V_p)$. It ranges from 0.6 to 0.95

For non spherical particles

$D_p = 6(1 – Є)/ Ø_s S$

The modified form of Ergun's equation for the binary mixture will be :

$$\Delta P/L = [\{150\, V_0 \mu (1-Є)^2\}/(Ø_{sa} D_{sa} + Ø_{sb} D_{sb})^2 Є^3] + [\{1.75 V_0^2 \rho (1-Є)\}/(Ø_{sa} D_{sa} + Ø_{sb} D_{sb}) Є^3$$

Minimum voidage to be determined by passing fluid up through the bed and noting the bed height at incipient the bed and nothing the bed height at incipient particles motion or minimum fluidization

$$Є_{mf} = (1-W) / L_{mf} A(\rho_s – \rho_{f)})$$

A correlation is to be developed for fluctuation ratio depending upon h_s , D_d, $(G_f – G_{mf})/G_{f.}$

STUDY OF DIFFERENT SIZES OF DOLOMITE PARTICLES

Mesh size	**Average diameter**
-8 + 12	1.7
-10 + 12	1.55
-12 + 14	1.29
-18 + 25	0.725
-30 + 25	0.55

For a binary mixture D_{pi}/D_{p2} should be less that 1.3. So, binary mixture used in the experiment is as follows

D_{pi}	D_{p2}	D_{pi}/D_{p2}
1.7	1.29	1.3178
1.55	1.29	1.2015
0.725	0.55	1.318
1.7	1.55	1.096

For binary mixture of size (-10 + 12) and (-12 + 14), (-8 + 12) and (-10 + 12) and for binary mixture of (-8 + 12) and (-12 + 14), (-18 + 25) and (- 30 + 25), the ratio of mixing is 50:5

STUDY OF DYNAMIC BEHAVIOUR OF HOMOGENEOUS MIZTURE IN A PROMOTED BED

Fluctuation ratio (r) = maximum expanded bed height (H_{fmax})/ minimum expanded bed height(H_{fmin})

Expansion ratio(R) = average expanded bed height / initial static bed height

Table 1

Static bed height = 8 cms

Proportion = 50:50

Room temperature = 38 ^{o}C

Promoter used = 5 rods

G (m^3/hr	G (Kg/hrm^2)	h (cm off hg)	ΔP (Kg/ms^2	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuatio ratio (r)	Expansio ratio (R)
3	1788.9	4.3	672.82	-	-	-	-
4	2385.2	5.5	860.6	-	-	-	-
5	2981.6	6.4	1001.4	-	-	-	-
6	3577.9	7.0	1095.3	-	-	-	-
7	4174.2	7.2	1126.6	-	-	-	-
8	4770.5	7.5	1173.5	-	-	-	-
9	5366.8	7.7	1204.8	11.3	10.5	1.10	1.36
10	5963.1	7.5	1173.5	14.8	12.5	1.18	1.71
11	6559.4	7.5	1173.5	16.7	13.5	1.23	1.89
11.5	6857.6	7.5	1173.5	19.2	15.2	1.26	2.15
12	7155.8	7.5	1173.5	25.0	15.8	1.58	2.55
13	7752.1	7.5	1173.5	26.7	16.0	1.67	2.67
14	8348.4	7.5	1173.5	27.9	16.8	1.81	2.96

Table 2

Static bed height = 8 cms
Proportion = 50:50
Room temperature = 38 ^{o}C
Promoter used = 9 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	3.2	500.7	-	-	-	-
4	2385.2	4.3	672.8	-	-	-	-
5	2981.6	5.5	860.6	-	-	-	-
6	3577.9	6.1	954.5	-	-	-	-
7	4174.2	6.7	1236.1	-	-	-	-
8	4770.5	7.9	1048.3	-	-	-	-
9	5366.8	6.5	1017.05	1.66	11	1.06	1.42
10	5963.1	6.5	1017.05	2.65	11.5	1.1	1.51
11	6559.4	6.5	1017.05	13.7	12	1.14	1.61
11.5	6857.6	6.5	1017.05	16.2	13.7	1.18	1.87
12	7155.8	6.5	1017.05	18.8	14.9	1.26	2.11
13	7752.1	6.5	1017.05	24.8	16.1	1.54	2.56
14	8348.4	6.5	1017.05	30.6	17.5	1.75	3.0

Table 3

Static bed height = 8cms
Proportion = 50:50
Room temperature = 38 ^{o}C
Promoter used = 13 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	2.3	359.88	-	-	-	-
4	2385.2	4.0	625.88	-	-	-	-
5	2981.6	4.1	641.52	-	-	-	-
6	3577.9	5.5	860.6	-	-	-	-
7	4174.2	5.7	891.88	-	-	-	-
8	4770.5	6.4	1001.4	-	-	-	-
9	5366.8	7.1	1110.93	-	-	-	-
10	5963.1	7.5	1173.52	-	-	-	-
11	6559.4	7.5	1173.52	11.7	11.2	1.04	1.43
12	7155.8	7.5	1173.52	14.25	12.5	1.14	1.67
13	7752.1	7.5	1173.52	16.00	13.3	1.2	1.83
14	8348.4	7.5	1173.52	21.2	15.6	1.36	2.3
15	8944.7	7.5	1173.52	27.05	16.8	1.61	2.74

Table 4

Static bed height = 8cms
Proportion = 50:50
Room temperature = 38 ^{o}C
Promoter used = 17 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	2.5	391.2	-	-	-	-
4	2385.2	3.6	563.3	-	-	-	-
5	2981.6	4.8	751.05	-	-	-	-
6	3577.9	6.5	1017.05	-	-	-	-
7	4174.2	7.0	1095.3	-	-	-	-
8	4770.5	7.2	1126.6	-	-	-	-
9	5366.8	7.3	1142.3	-	-	-	-
10	5963.1	7.3	1142.3	-	-	-	-
11	6559.4	7.3	1142.3	11	10.8	1.02	1.36
12	7155.8	7.3	1142.3	12.7	11.7	1.09	1.525
13	7752.1	7.3	1142.3	15.9	13.5	1.8	1.84
14	8348.4	7.3	1142.3	18.8	14.5	1.3	2.08
15	8944.7	7.3	1142.3	24.2	15.6	1.55	2.5

Table 5

Static bed height = 12cms
Proportion = 50:50
Room temperature = 35.5 ^{o}C
Promoter used = 5 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	4.3	672.82	-	-	-	-
4	2385.2	5.9	923.2	-	-	-	-
5	2981.6	7.7	1204.8	-	-	-	-
6	3577.9	8.6	1345.6	-	-	-	-
7	4174.2	9.1	1423.9	-	-	-	-
8	4770.5	9.6	1502.1	-	-	-	-
9	5366.8	9.7	1517.7	-	-	-	-
10	5963.1	9.9	1549.05	14.0	12.7	1.10	1.11
11	6559.4	9.9	1549.05	15.5	13.5	1.15	1.21
12	7155.8	9.9	1549.05	18.1	15.2	1.19	1.39
13	7752.1	9.9	1549.05	24.9	16.1	1.35	1.71
14	8348.4	9.9	1549.05	26.9	16.8	1.61	1.82
15	8944.7	9.9	1549.05	30.1	17.3	1.79	1.97

Table 6

Static bed height = 16cms
Proportion = 50:50
Room temperature = 35.5 ^{o}C
Promoter used = 5 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	5.6	876.23	-	-	-	-
4	2385.2	7.8	1220.5	-	-	-	-
5	2981.6	10.4	1627.3	-	-	-	-
6	3577.9	11.6	1815.05	-	-	-	-
7	4174.2	12.53	1960.6	-	-	-	-
8	4770.5	13.1	2049.7	-	-	-	-
9	5366.8	13.2	2065.4	-	-	-	-
10	5963.1	13.2	2065.4	17.9	17.2	1.04	1.1
11	6559.4	13.2	2065.4	19.2	8.1	1.06	1.17
12	7155.8	13.2	2065.4	20.8	18.9	1.10	1.24
13	7752.1	13.2	2065.4	25.35	19.5	1.30	1.4
14	8348.4	13.2	2065.4	31.7	20.3	1.56	1.65
15	8944.7	13.2	2065.4	37.6	22.1	1.7	1.87

Table 7

Static bed height = 16cms
Proportion = 50:50
Room temperature = 35.5 ^{o}C
Promoter used = 5 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	7.4	1157.9	-	-	-	-
4	2385.2	9.9	1549.05	-	-	-	-
5	2981.6	12.8	2002.8	-	-	-	-
6	3577.9	14.7	2300	-	-	-	-
7	4174.2	15.7	2456.6	-	-	-	-
8	4770.5	16.8	2628.7	-	-	-	-
9	5366.8	16.8	2628.7	-	-	-	-
10	5963.1	16.8	2628.7	-	-	-	-
11	6559.4	16.8	2628.7	-	-	-	-
12	7155.8	16.8	2628.7	23.2	22.5	1.03	1.14
13	7752.1	16.8	2628.7	28.4	23.7	1.2	1.3
14	8348.4	16.8	2628.7	34.1	26.2	1.3	1.5
15	8944.7	16.8	2628.7	45.2	27.5	1.64	1.815

Table 8

Static bed height = 20cms
Proportion = 50:50
Room temperature = 35.5 ^{o}C
Promoter used = 9 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	6.1	954.5	-	-	-	-
4	2385.2	7.7	1204.8	-	-	-	-
5	2981.6	9.8	1533.4	-	-	-	-
6	3577.9	13.2	2065.4	-	-	-	-
7	4174.2	14.7	2300	-	-	-	-
8	4770.5	15.5	2425.3	-	-	-	-
9	5366.8	15.8	2472.2	-	-	-	-
10	5963.1	15.8	2472.2	-	-	-	-
11	6559.4	15.8	2472.2	-	-	-	-
12	7155.8	15.8	2472.2	22.4	22	1.02	1.11
13	7752.1	15.8	2472.2	25.6	22.5	1.14	1.2
14	8348.4	15.8	2472.2	30	24	1.25	1.35
14.5	8646.5	15.8	2472.2	33.4	25.3	1.32	1.47
15	8944.7	15.8	2472.2	39.1	26.1	1.5	1.63
16	9541.01	15.8	2472.2	51.3	28.5	1.8	1.99

Table 9

Static bed height = 20cms
Proportion = 50:50
Room temperature = 35.5 ^{o}C
Promoter used = 13 rods

G (m^3/hr)	**G (Kg/hrm^2)**	**h (cms off hg)**	**ΔP (Kg/ms^2)**	**Maximum bed height(H_{fmaz}) cms**	**Minimum bed height(H_{fmini}) cms**	**Fluctuation ratio (r)**	**Expansion ratio (R)**
3	1788.9	5.0	782.35	-	-	-	-
4	2981.6	6.9	1079.64	-	-	-	-
5	2385.2	8.9	1392.6	-	-	-	-
6	3577.9	11.7	1830.7	-	-	-	-
7	4174.2	14.5	2268.81	-	-	-	-
8	4770.5	15.4	2409.6	-	-	-	-
9	5366.8	18.6	2910.3	-	-	-	-
10	5963.1	20.3	3176.3	-	-	-	-
11	6559.4	19.8	3098.1	-	-	-	-
12	7155.8	19.8	3098.1	-	-	-	-
13	7752.1	19.8	3098.1	23.2	21.7	1.07	1.12
13.5	8050.2	19.8	3098.1	25.7	23.4	1.1	1.23
14	8646.5	19.8	3098.1	30	25	1.2	1.37
15	8944.7	19.8	3098.1	40.15	27.5	1.16	1.69

Table 10

Static bed height = 20cms
Proportion = 50:50
Room temperature = 35.5 ^{o}C
Promoter used = 17 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	6	938.82	-	-	-	-
4	2385.2	8.4	1314.3	-	-	-	-
5	2981.6	10.9	1705.52	-	-	-	-
6	3577.9	14.2	2221.9	-	-	-	-
7	4174.2	17.5	2738.2	-	-	-	-
8	4770.5	19.7	3082.4	-	-	-	-
9	5366.8	18.3	2863.4	-	-	-	-
10	5963.1	18.3	2863.4	-	-	-	-
11	6559.4	18.3	2863.4	-	-	-	-
12	7155.8	18.3	2863.4	-	-	-	-
13	7752.1	18.3	2863.4	21.4	21	1.02	1.06
14	8348.4	18.3	2863.4	25.8	23	1.12	1.26
15	8946.5	18.3	2863.4	33.6	24.5	1.39	1.45
16	9541.01	18.3	2863.4	33.6	26.3	1.52	1.66

Table 11

Static bed height = 20cms
Proportion = 50:50
Particle average diameter(-18+25 and -25+30) = 0.625
Promoter used = 5 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	7.5	1173.52	10.4	10.2	1.02	1.3
4	2385.2	7.9	1236.1	11.9	11.1	1.07	1,4
5	2981.6	8.5	1330	13.3	11.5	1.15	1.53
6	3577.9	8.5	1330	14.8	12.3	1.2	1.7
7	4174.2	8.5	1330	16.0	12.6	1.27	1.8
8	4770.5	8.5	1330	17.9	13.3	1.35	1,95
9	5366.8	8.5	1330	20.3	13.9	1.46	2.2
10	5963.1	8.5	1330	25.3	15.5	1.63	2.55

Table 10

Static bed height = 8 cms
Proportion = 50:50
Particle average diameter(-8+12 and -12+14)=1.495mm
Promoter used = 5 rods

G (m^3/hr)	G (Kg/hrm^2)	h (cms off hg)	ΔP (Kg/ms^2)	Maximum bed height(H_{fmaz}) cms	Minimum bed height(H_{fmini}) cms	Fluctuation ratio (r)	Expansion ratio (R)
3	1788.9	2.9	453.8	-	-	-	-
4	2385.2	3.4	531.9	-	-	-	-
5	2981.6	3.9	610.2	-	-	-	-
6	3577.9	4.4	688.5	-	-	-	-
7	4174.2	4.8	751.05	-	-	-	-
8	4770.5	5.1	797.9	-	-	-	-
9	5366.8	6.0	938.8	-	-	-	-
10	5963.1	6.7	1048.3	-	-	-	-
12	7155.8	7.0	1095.3	-	-	-	-
14	8348.4	7.0	1095.3	10.3	9.2	1.03	1.11
16	9541.01	7.0	1095.3	12.2	9.6	1.12	1.22
18	10733.6	7.0	1095.3	13.9	10,1	1,16	1.29
20	11926.3	7.0	1095.3	16.7	11.3	1.32	1.825
22	13118.9	7.0	1095.3	19.2	11.9	1.48	1.75
24	14311.5	7.0	1095.3	23.7	13.0	1.62	2.29

Table 10

Static bed height = 8 cms
Proportion = 50:50
Particle average diameter(-8+12 and -12+14)=1.495mm
Promoter used = 5 rods

G (m^3/hr)	**G (Kg/hrm^2)**	**h (cms off hg)**	**ΔP (Kg/ms^2)**	**Maximum bed height(H_{fmaz}) cms**	**Minimum bed height(H_{fmini}) cms**	**Fluctuation ratio (r)**	**Expansion ratio (R)**
4	2385.2	0.4	62.6	-	-	-	-
6	3577.9	0.7	109.6	-	-	-	-
8	4770.5	1.1	172.1	-	-	-	-
9	5366.8	1.6	250.3	-	-	-	-
10	5963.1	2.1	328.6	-	-	-	-
12	7155.8	2.8	438.1	-	-	-	-
14	8348.4	3.2	500.7	-	-	-	-
16	9541.01	3.7	578.9	-	-	-	-
18	10733.6	4.5	704.11	-	-	-	-
20	11926.3	5.6	876.2	-	-	-	-
22	13118.9	6.1	954.5	8.8	8.4	1.05	1.075
24	14311.5	6.9	1079.6	10.1	9	1.12	1.19
26	15504.2	6.6	1032.7	10.7	9.2	1.16	1.24
28	16696.8	6.6	1032.7	11.7	9.5	1.23	1.325
30	17889.4	6.6	1032.7	13.1	10.1	1.3	1.57
32	19082.62	6.6	1032.7	18	10.7	1.3	1.57
34	20274	6.6	1032.7	22.9	12.0	1.42	2.18
36	21467.3	6.6	1032.7	24	13.5	1.47	2.24
38	22659	6.6	1032.7	26.3	15.3	1.5	2.36

DEVELOPMENT OF CORRELATION

The model equations are assumed to be linear and the equations take a general form

$$Y = a_o + a_1X_1 + a_2X_2 + a_3X_3 + a_{12}X_1X_2 + a_{13}X_1X_2 + a_{23}X_2X_3 + a_{123}X_1X_2X_3 \quad \rightarrow (1)$$

i. The coefficients are calculated by Yate's technique

$$b_i = \sum \alpha_i y_i / N$$

where b_i is the coefficient, y_i is the response, α_i is the level of the variable and N is the total number of treatments.

ii. calculation of levels of variables

X_1: $(G_F - G_{mf})/ G_{mf}$ = (flow rate of fluidizing medium, $(G_F - G_{mf})/ G_{mf}$ -0.5)/ (0.5 – 0.25)

X_2: static bed height, h_s = (static bed height,(m) – 14)/(14-8)

X_3: equivalent diameter, D_{eq}= (equivalent diameter, (m) – 3.204)/ (3.204-2.42)

The equivalent data based on factorial design can be given as follows:-

Table 7.1

Yates std order	**Run no**	**X_1**	**X_2**	**X_3**	**Fluctuation ratio**
1	1	0.25	8	2.42	1.1
X_1	2	0.75	8	2.42	1.62
X_2	3	0.25	20	2.42	1.31
X_1 X_2	4	0.75	20	2.42	1.45
X_3	5	0.25	8	3.988	1.17
X_1 X_3	6	0.75	8	3.988	1.7
X_2 X_3	7	0.25	20	3.988	1.2
X_1 X_2 X_3	8	0.75	20	3.988	1.64

Table 7.2
Yate's technique

Yates std order	**Run no**	**X_1**	**X_2**	**X_3**	**$X_1 X_2$**	**$X_1 X_3$**	**$X_2 X_3$**	**$X_1X_2X_3$**
1	1	-	-	-	+	+	+	-
X_1	2	+	-	-	-	-	+	+
X_2	3	-	+	-	-	+	-	+
$X_1 X_2$	4	+	+	-	-	-	-	-
X_3	5	-	-	+	+	-	-	+
$X_1 X_3$	6	+	-	+	-	+	-	-
$X_2 X_3$	7	-	+	+	-	-	+	-
$X_1X_2X_3$	8	+	+	+	+	+	+	+

CALCULATION OF COEFFICIENTS

$b_i = \sum \alpha_i y_i /N$

$a_o = \sum r/8 = 11.19/8 = 1.399$

$a_1 = 1.63/8 = 0.204$

$a_2 = 0.01/8 = 0.00125$

$a_3 = 0.23/8$

$a_{12} = -2.05/8 = -0.256$

$a_{13} = 0.31/8 = 0.039$

$a_{23} = -0.07/8 = -0.00875$

$a_{123} = 0.29/8 = 0.036$

The following equation is obtained for the fluctuation ratio(r)

$r = 1.399 + 0.204X_1 + 0.00125X_2 + 0.029 X_3 – 0.256 X_1 X_2 + 0.039 X_1 X_3 – 0.00875 X_2 X_3 + 0.036 X_1X_2X_3$

the value of coefficients indicates the magnitude of the effect of the variable and the sign of the coefficient gives the direction of the effect of the variable .i.e. a positive coefficient indicating an increase in the value of the response with increase in the

value of the variable and a negative coefficient showing that the response decreases with increase in the value of the variable.

COMPARISION OF THEORETICAL VALUE

The correlation for minimum fluidization velocity developed by Kumar and Roy in case of bed with rod promoters

$$G'_{mf} = [0.000829 + 0.001(D_e/D_c)^{-0.48}]*[Ø_s^2 d_p^2 \rho_f(\rho_s - \rho_f)g/\mu]$$

The value calculated from the correlation have been compared with the experimental values as shown in table 8.1 and 8.2. they are found to be in good agreement with each other as deviation was between $\pm$ 10%

Also fluidization ratio for promoted bed by the correlation

$$r = 0.004(h_s/ D_e)^{0.15}(dp/ D_e)^{-0.29}(\rho_s/ \rho_g)^{0.29}((G_F - G_{mf})/ G_{mf})^{0.3}$$

The values are compared in table 8.3, 8.4 and 8.5

Table 8.1

Promoter	**G_{mf} (theo)**	**G_{mf} (expt)**	**% error**
5	5034.2	5366.8	6.60
9	5103.8	4770.5	5.15
13	5163.5	5366.8	3.93
17	5199.3	5366.8	3.22

Mean deviation = Standard deviation =

Table 8.2

Ø	**Particle dia**	**G_{mf} (theo)**	**G_{mf} (expt)**	**% error**
0.8	0.6375	1803.2	2385.2	
0.6	1.42	5043.2	5366.8	
0.6	1.495	5579.9	5963.1	

Mean deviation = Standard deviation =

The experimental valuyes ofminimum fluidization velocity are found to be within a standard deviation of 1

Table 8.3
For fluctuation ratio
Different bed height

Bed ht	$((G_F - G_{mf})/G_{mf})$	r(theo)	r(expt)	% deviation
8cms	0.125	1.22	1.1	9.83
	0.25	1.27	1.17	7.87
	0.375	1.31	1.24	5.34
	0.5	1.33	1.58	18.7
	0.625	1.36	1.67	22.7
	0.75	1.38	1.7	23.18
	0.875	1.4	1.81	29.2
12cms	0.125	1.23	1.1	10.5
	0.25	1.33	1.19	10.5
	0.375	1.38	1.6	15.9
	0.5	1.35	1.63	20.7
	0.625	1.38	1.7	23.1
	0.75	1.4	1.74	24.2
10cms	0.125	1.24	1.04	16.1
	0.25	1.3	1.06	18.46
	0.375	1.34	1.1	17.9
	0.5	1.36	1.3	4.41
	0.625	1.4	1.56	11.4
	0.75	1.42	1.7	19.7
20cms	0.125	1.25	1.03	17.6
	0.25	1.31	1.2	8.39
	0.375	1.34	1.3	2.98
	0.5	1.38	1.43	3.62
	0.625	1.41	1.53	8.5
	0.75	1.43	1.64	14.68

Table 8.4
For different promoters

Promoters	$((G_F-G_{mf})/G_{mf})$	r(theo)	r(expt)	% deviation
9 promoters	0.125	1.21	1.06	1.21
	0.25	1.26	1.14	9.52
	0.375	1.3	1.19	9.24
	0.5	1.34	1.26	5.97
	0.625	1.35	1.54	1.407
	0.75	1.37	1.75	27.7
13 promoters	0.125	1.21	1.04	14.05
	0.25	1.26	1.14	9.52
	0.375	1.29	1.2	6.9
	0.5	1.32	1.35	2.27
	0.625	1.36	1.58	16.1
	0.75	1.40	1.62	15.7
17 promoters	0.125	1.22	1.02	16.39
	0.25	1.25	1.1	12
	0.375	1.28	1.19	7.03
	0.5	1.31	1.3	0.76
	0.625	1.33	1.45	9.02
	0.75	1.35	1.62	20

Table 8.5
For different particle sizes

Particle size	$((G_F-G_{mf})/G_{mf})$	r(theo)	r(expt)	% deviation
0.6375	0.125	1.28	1.02	20.3
	0.25	1.34	1.07	20.15
	0.375	1.29	1.15	10.8
	0.5	1.43	1.2	16.08
	0.625	1.45	1.26	13.1
	0.75	1.48	1.35	8.78
1.495	0.125	1.22	1.12	8.19
	0.25	1.26	1.16	7.93
	0.375	1.3	1.21	7.43
	0.5	1.34	1.31	2.23
	0.625	1.38	1.51	9.42
	0.75	1.42	1.62	14.08
1.625	0.125	1.22	1.12	8.19
	0.25	1.26	1.16	7.93
	0.375	1.29	1.23	4.65
	0.5	1.32	1.3	1.51
	0.625	1.34	1.36	1.49
	0.75	1.36	1.42	4.41

RESULTS :-

(1) Experimental data have been collected in gas-solid promoted fluidized beds to study the different aspects of hydrodynamic characteristics

(2) It is observed that minimum fluidization velocity increases with increase in particle size.

(3) The pressure drop at minimum fluidization velocity increased with particle size and static bed height as expected

(4) A correlation for fluctuation ratio was developed using statistical approach which can be given as

$$r = 1.399 + 0.204X_1 + 0.00125X_2 + 0.029\ X_3 - 0.256\ X_1\ X_2 + 0.039\ X_1\ X_3 - 0.00875\ X_2\ X_3 + 0.036\ X_1X_2X_3$$

CONCLUSIONS:-

(1) Experimental work has been carried out to study the effect of flowrate of fluidizing medium static bed height, particle size on hydrodynamic parameters viz minimum fluidization velocity, pressure drop at minimum fluidization velocity, fluctuation ratio and expansion ratio for promoted gas-solid fluidized bed.

(2) Equation (based on factorial design of analysis) proposed for the prediction of the fluctuation ratio (r) bring out the interaction effects of different variables which is not possible from conventional experimentation.

(3) The number of experiments required for development of model equation from factorial design is considerably used in comparision to conventional experimentation

(4) The effect of different variables could be explicitly and quantitatively presented from the analysis of factorial design

(5) The equations proposed for the prediction of fluctuation ratio are mostly empirical in nature and applicable for air-dolomite system only. Hence, they have limitations with respect to their applicability beyond the range of experimental conditions. It is desired that the equations be developed based on conservation equations and the theory of fluid and particle mechanics making use of the information obtained from the statistically designed experiments

(6) The equation require that the variables be substituted in appropriate dimensions to make use of them to calculate responses

(7) The results from the equation on agree with experimental data with percent standard deviation.

NOMENCLATURE:-

dp: particle size, m
dpm: mean particle size, m
D_c: column diameter, m
D_e: equivalent column diameter, m
G_f: fluidizatin mass velocity, Kg/hrm^2
G_{mf}: minimum fluidization mass velocity in promoted beds, Kg/hrm^2
h_{max}: maximum height of fluidized bed, m
h_{min}: minimum height of fluidized bed, m
ΔP: pressure drop across static bed
r: bed fluctuation ratio
R: bed expansion ratio
g: acceleration due to gravity
ρ_s: density of solid, kg/m^3
ρ_g: density of gas, kg/m^3
μ: viscousity
Ø: sphericity
X_1: ratio of flow rate of fluidizing medium, $((G_F–G_{mf})/G_{mf})$
X_2: static bed height
X_3: equivalent diameter

SUBSCRIPTS

av: average
f: fluidized bed
mf: minimum fluidization condition

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