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"AN ADVANCED MODELLING OF BUBBLING FLUIDIZED BED FOR ADDRESSING HYDRO DYNAMIC TYPE OF FLOW INTERMS OF MATERIAL BALANCE FOR GASES AND SOLIDS TOWARDS SUSTAINABLE DEVELOPMENT COMPRISING THE BALANCED OPTIMIZATION ORIENTED MODELLING ASPECTS OF BUBBLING SIZE''

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ABSTRACT

Generally, it is predicted that fluidized beds are used to have better performance in terms of combustion chemistry and reasonable chemical kinetics for the combustion process in order to satisfy the requirements of reliable quantity of fuel along with its favouring resistance for any application with less environmental pollution. In this contact, a modest attempt has been made for modeling, bubbling fluidized bed for representing hydrodynamic type of flow in term of material balance for gases and solids towards sustainable development by holding very less environmental pollution. The formation of "phthalic anhydride '' is highly exothermic, and even with the most careful design, the heat removal from packed bed reactors can become unchangeable, leading to temperature runaways, melts downs and even explosions.

The invention of the fluidized bed with its suspended and rapidly mixing solids completely overcomes this critical situation. This is because, the rapid mixing of solids, and the large heat sink as solids will only allow the bed temperature to change very slowly and it can be easily controlled. Another critical situation in this contact is that, catalyst formulation has been very successful in creating better and better catalyst, these that give higher and higher rates of reaction. The catalyst volumetric efficiency is usually kept with "Thiele modulus" and it is noticed that by using smaller and smaller particles, as overall reaction rate constant is made higher and higher. This leads to use suspended solids and also it is noted that with these very

KEYWORDS:

Bubbling Fluidized Bed,
hydrodynamic flow,
chemical kinetics,
material balance,
"phthalic anhydride",
Thiele modulus,
Voids fraction in gas solid systems,
gas velocity,
bubble size,
wakes,
less environment pollution,
K-L model for BFB,
cloud,
emulsion,
minimum fluidization.

effective catalysts, the required residence time of reactant gas becomes very small, a fact records for a large 30 m higher reactor. The expressions developed/modeled slows that, if ε_{mf} - void fractions in gas solid systems at minimum fluidizing condition are known, then m^3 of wake per m^3 of bubble can be estimated and then the superficial gas velocity in bed u_0 and u_{mf} gas velocity at minimum fluidized condition can be measured and finally all the flow quantities and reginal volumes can be determined in terms of size of the bubble. The use of this model is to calculate chemical reactor behavior as a direct method. The special feature of this model is that it's one parameter can be tested against what is measured and what is observed. Basically hydro dynamic models rests may develop either large bubbles or small bubbles depending on bed diameter, distribution design, baffle arrangements etc, therefore, bubble size must enter as the primary parameter in the model. Hence the consequence of this factor is that models which donot allow for different bubble sizes at given imposed bed condition certainly cannot be adequate. Therefore it is predicted that the feasibility of this type of model for bubbling fluidized bed for representing hydrodynamic types of flow holds reasonable validity and identity towards agreeable sustainable development because of its uniqueness exhaustively.

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1.INTRODUCTION

Generally, hydrodynamic flow type of models can be developed to represent the bubbling fluidized bed, based on the size of the bubble and minimum fluidizing condition. The development of K-L model in this context includes, pass an excess of gas upward through a bed of fine particles. Basically with a large enough bed diameters, one can get a freely budding bed of rapid bubbles. In order to simplifying the complications, the following assumptions are made.

i) All the bubbles are spherical in nature, all of the same size d_b and it follows the Davidson model ofcourse, and one can neglect the upflow of gas through the cloud.Hence, and the bed contains bubbles surrounded by thin clouds rising through a emulsion, because the cloud volume is small as compared to that of the bubble. This is the regime where$u_b \gg u_e$.

ii) The emulsion stays at minimum fluidization conditions hence the relative gas solid velocity stay as constant in the emulsion.

Each bubble drags ups a wake of solids behind it. This generates a circulation of solids in the bed, up flow behind the bubbles, and downflow everywhere else in the bed .if this down flow of solids are quicker enough, there gas up flow in the emulsion is impeded , can actually stop and even reverse itself. Such down flow of gas has been observed and recorded and it occurs, when,$u_0 > (3\ to\ 11)u_{mf}$. Any up flow of gas in the emulsion can be neglected.

Let,

u_0 →Superficial gas velocity in the bed ($m^3 of\, gas/m^2 of\, bed.s$)

d→diameter (m)

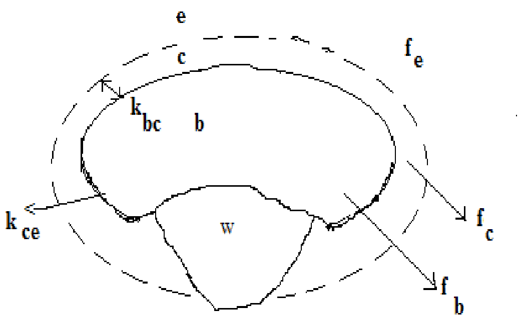
ε →fraction of voids in the bed

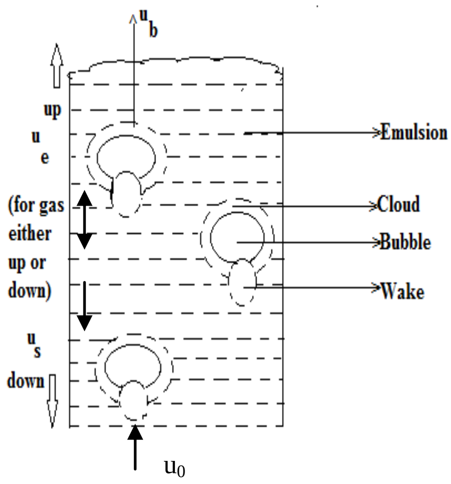
b, c, e,w→bubble, cloud, emulsion,wake

m → packed bed

m_f →minimum fluidization

f →bubbling fluidized bed conditions.





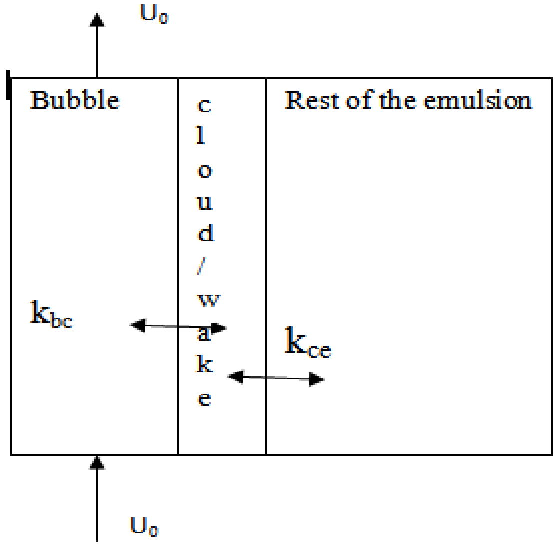
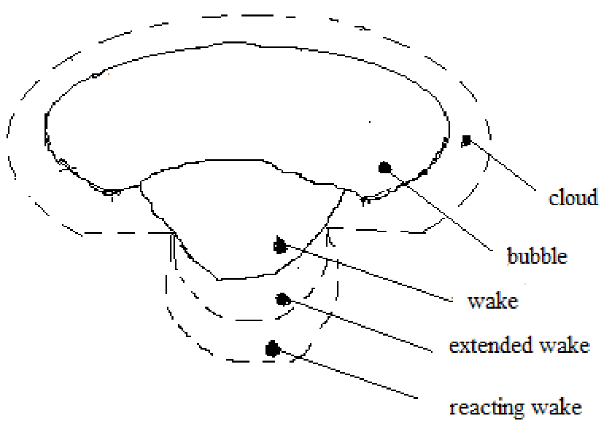


Figure-1. –models /symbols used to describe the K-L bubbling gas fluidized bed

In short, given $u_{mf}, \varepsilon_{mf}, u_0, \propto$ and the effective bubble size in the bed d_b, this model exhibits all the other properties of the bed such as flows, region volumes, interchange rates, and consequently the reaction behaviour. One can also propose the following structure for modelling purpose.



2.MATERIAL BALANCE FOR GAS /SOLIDS:

From Kunii and Levenspiel [1999], a material balance for the bed material implies that,

$$u_{br} = 0.711 \times \sqrt{g d_b}$$

m/sec ….rise velocity of a single bubble in a bed otherwise at u_{mf} $------(1)$

g→ acceleration due to gravity =9.8 m/sec^2.

$$u_b = [u_0 - u_{mf} + u_{br}]m/s$$

rise velocity of bubbles in a bubbling in a bubbling bed$-----------(2)$

δ = Bed Fraction in bubbles

$$= \left[\frac{m^3 of bubble}{m^3 of bed}\right] = \left[\frac{volume of bubble}{volume of bed}\right] -----------(3)$$

$$\delta = \left[\frac{u_0 - u_{mf}}{u_b}\right] = \left[1 - \frac{u_{br}}{u_b}\right]$$

And for $u_b \gg u_{mf}$, we can use $\delta \cong \frac{u_0}{u_b}$

$$H_m(1-\varepsilon_m) = H_{mf}(1-\varepsilon_{mf}) = H_f(1-\varepsilon_f)$$

$$1 - \delta = \left[\frac{1-\varepsilon_f}{1-\varepsilon_{mf}}\right] = \left[\frac{H_{mf}}{H_m}\right]$$

$$u_s = \left[\frac{\propto \delta u_b}{1 - \delta - \propto \delta}\right] m/s \rightarrow Downflow of emulsion solids ----- (4)$$

$$u_e = \left\{\left[\frac{u_{mf}}{\varepsilon_{mf}}\right] - u_s\right\} m/s Rise velocity of emulsion gases ------(5)$$

(can be positive or negative).

by using Davidson's theoretical expression for bubble cloud circulation and "Higbie theory'' for cloud emulsion diffusion, the interchange of gas between bubble and the cloud is then found to be

$$K_{bc} = \left\{4.50\left[\frac{u_{mf}}{d_b}\right] + 5.85\left(\frac{D^{1/2} g^{1/4}}{d_b^{5/4}}\right)\right\}$$

$$= \frac{(interchange volume between \text{"b" and "c" or "c" and "b"})/s}{(volume of bubble)}, s^{-1} --- (6)$$

and between cloud-wake and emulsion,

$$k_{ce} = 6.77\left[\frac{\varepsilon_{mf} D u_{br}}{d_b^3}\right]^{1/2} = \left[\frac{interchange volume/s}{volume of bubble}\right], s^{-1} ---------(7)$$

$$f_b = 0.001 \sim 0.01 = \left[\frac{volume of solids in bubbles}{volume of bed}\right] ------------ (8)$$

(approximate estimation from the experiment)

$$f_c = \delta(1-\varepsilon_{mf})\left[\frac{3u_{mf}/\varepsilon_{mf}}{u_{br}-u_{mf}/\varepsilon_{mf}}+\propto\right]$$

$$= \left[\frac{volumeofsolidsincloudandwake}{volumeofbed}\right] ----------------(9)$$

$$f_e = [(1-\varepsilon_f)-f_c-f_b]$$

$$= [(1-\varepsilon_{mf})(1-\delta)-f_c-f_b] ----------------(10)$$

$$= \left[\frac{volumeofsolidsintherestoftheemulsion}{volumeofbed}\right]$$

$$\therefore f_{total} = f_b + f_c + f_e = (1-\varepsilon_f) -------------- (11)$$

∴Phase distribution co-efficient

$$H_{BFB} = H_f = \frac{W}{\rho_s}.A.(1-\varepsilon_f) ---------------- (12)$$

3.CONTRIBUTION TO CATALYTIC REACTIONS

The mandatory questionable assumptions that we have made are,

i. The flow of gas through the cloud is neglected, since the cloud volume is very small for fast bubble

ii. The flow of gas either up on down is neglected through the emulsion, since the flow is much smaller than the flow through the bubbles.

In fact, we can consider the “emulsion as stagnant”. Moreover, many general expressions can be developed for beds, where bubbles have thick clouds (not too large and fast bubbles), or where flow through the emulsion is significant [u_0 is closer to u_{mf}],hence where $u_0 = (1-2u_{mf})$. However for fast bubble, seriously bubbling fine particle beds, the assumption already made are seems to be reasonable.

4.FIRST ORDER REACTION:

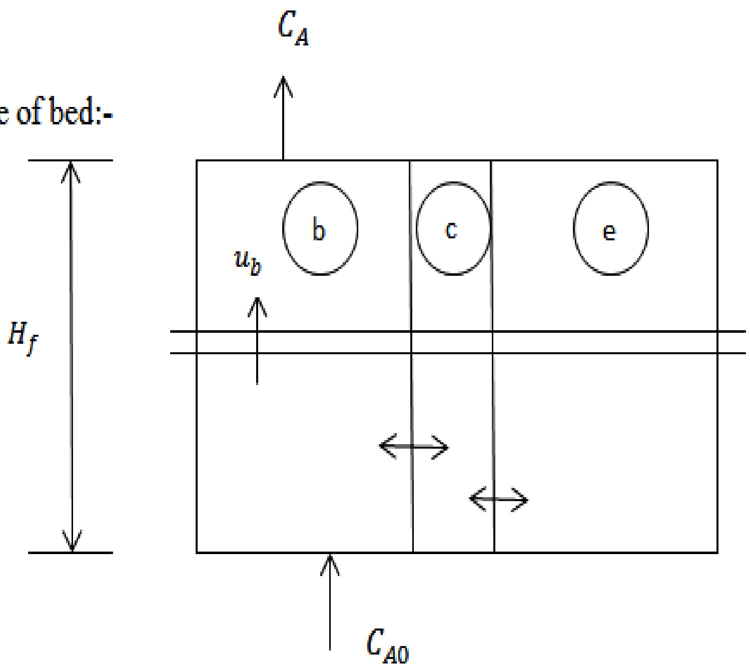
Let the reaction be:-

$$A \to R, \qquad r_A''' = k'''C_A \quad , \qquad \frac{mol}{m^3.S.S} ------------$$

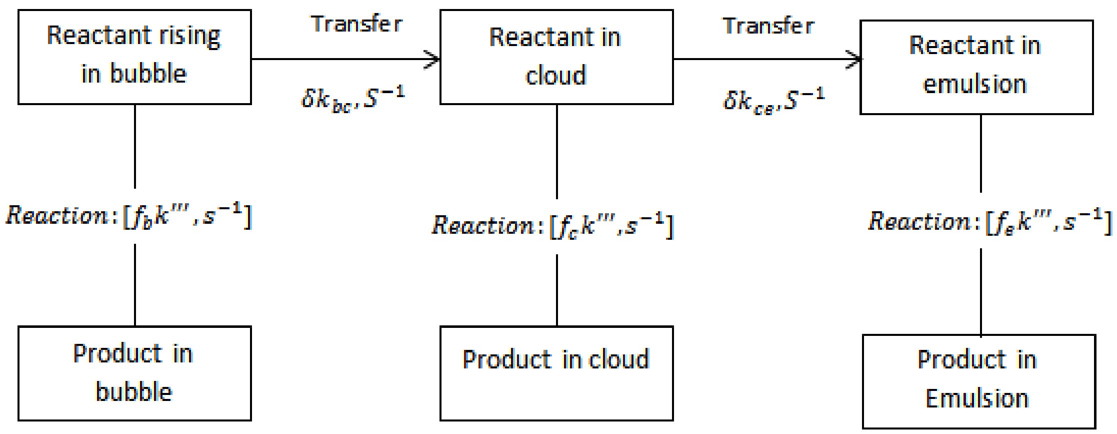
(13)

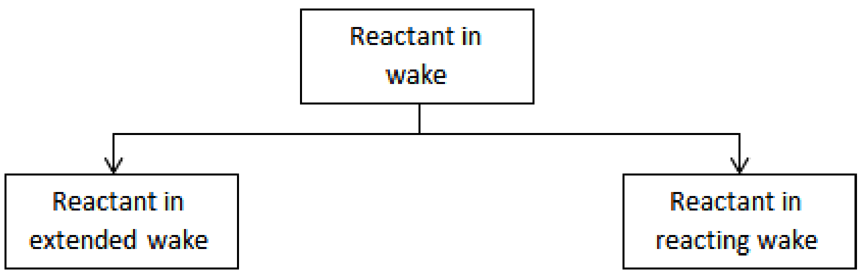
$$where\ ,\qquad k''' \rightarrow {m^3}/{m^3.S.S}$$

A slice of bed:-



For any segment of bed, we can have,





$$f_w = 0.003 \sim 0.02 = \left[\frac{volume\ of\ solidin\ the\ wakes}{volume\ of\ bed}\right] - - - - - - - - -(14)$$

(approximate estimation from the experiment)

According for these five resistances in series -parallel, eliminating cloud and emulsion concentrations, and integrating from the bottom to the bed implies that,

$$l_n = \frac{C_{AO}}{C_A} = [K^{'''}.\tau^{'''}] - - - - - - - - - - (15)$$

$$= \left[\frac{f_b k^{'''} + \cfrac{1}{\cfrac{1}{\delta k_{bc}} + \cfrac{1}{f_c k^{'''} + \cfrac{1}{\cfrac{1}{\delta k_{ce}} + \cfrac{1}{f_e k^{'''}}}}}}{f_{total}}\right] * \frac{f_{toatl}\,.H_{BFB}}{u_0}$$

where K‴ →Effective rate constant for the fluidized bed. $\left(m^3/_{m^3.s.s}\right)$

$$\tau^{'''} \to \left[\frac{f_{total}\ .H_{BFB}}{u_0}\right](m^3 s.s/m^3) - - - - - - - - - - (16)$$

We can also find that the average gas composition seen by solids is approximately expressed as

c_A, bathing the solids $= \left[\frac{C_{A0}-C_A}{K^{'''}\tau^{'''}}\right] = \left[\frac{C_{A0}X_A V_0}{K^{'''}V_S}\right] = \left[\frac{C_{A0}X_A V_0}{K'W}\right] - - - - - -(17)$

Vs→of solids done (w/ρ_s)

This quantity is important for non-catalytic gas /solid reactions, because it is this C_A that the solids see and react with. As far as the packed bed reactions are concerned, by assuming that the plug flow$K_{bc} \to \infty, K_{ce\to}\infty$, then equation (15) reduces to,

$$ln\frac{c_{A0}}{c_{Ap}} = k^{'''}.T^{'''} = \left[\frac{k^{'''}.H_p(1-\varepsilon_p)}{u_0}\right] = k'\tau' = \left[\frac{k'w}{u_0.A_t}\right] - - - - - - - - -(18)$$

(For plug flow)

$$c_{Ap} = \left[\frac{C_{A0}-C_{Ap}}{k^{'''}T^{'''}}\right] -(19)$$

(For plug flow)

By comparing equation (15) and (18) with equation (19), it shows that, Fluidized bed can be treated as a plug flow reactor, if K''' is replaced by k''' .

5. COMMENTS/INFERENCES

The fire terms in bricked of the performance equation (15),represents the complex serious –parallel resistance to mass transfer and reaction or for very fast reaction [High k” value],very little “A” gets as far as the emulsion and the first two terms are influencing for slow reaction, the latter terms becomes important.

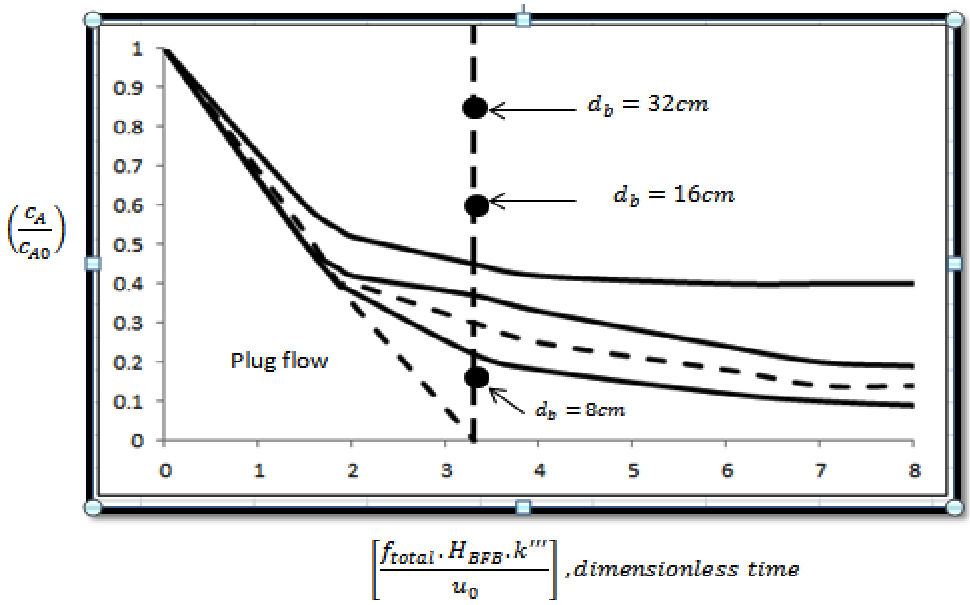


Fig-2:-Performance of a fluidized bed as a function of bubble size [comparison with plug flow and mixed flow prediction].

Since the bubble size is the one quantity which governs all the rate quantities with the exception of k'''. We can plot the performance of a fluidized bed as a function of d_b ,diameter of the bubble as shown in the figure. It is noted that large diameter of the bubble gives poor performance, because of extensive by passing of bubble gas, and that the performance the bed can drop considerably below mixed flow.For multiple reactions, the effect of this flow is much more important. Therefore ,for reactions in series, the lowering in the amount of intermediate formed can be,and usually it is quite drastic.

1.Hydrodynamic Flow Models:

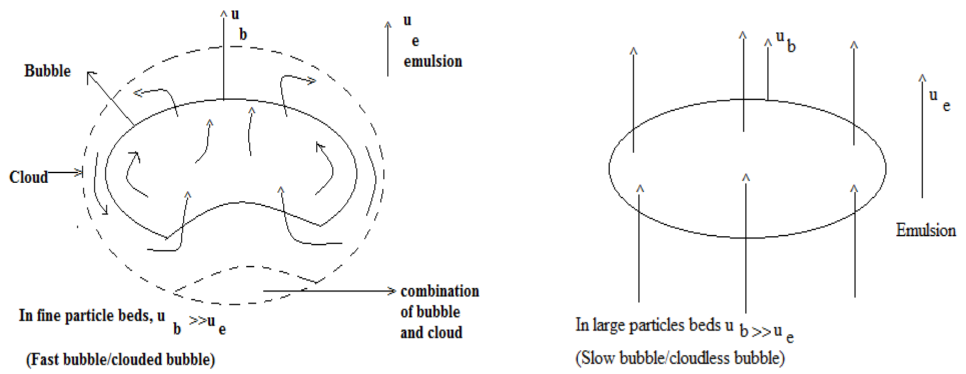


Figure 3.Extremes of gas flow in the vicinity of rising gas bubbles in BFBs.

$$\left[\frac{u_e}{u_{br}}\right] = \left[\frac{thickness\ of\ the\ cloud}{Diameter\ of\ the\ bubble}\right]$$

Two developments are of particular importance in this context. The first one is Davidson's remarkable theoretical development and experimental verification of the flow in the vicinity of a single rising bubble in a fluidized bed, which is otherwise at minimum fluidizing condition. It is found that the gas behavior in the vicinity of the bubble depends only on the bubble size, and that the gas behavior in the vicinity of the bubble depends only on the relative velocity of the rising bubble and of gas rising in the emulsion u_e and of course a completely different behavior was noticed. For catalytic reactions, we are only interested in fine particle beds, so that the large particle extreme can be neglected. For the fine particle bed, gas circulates within the bubble plus a thin cloud surrounding the bubble. Hence the bubble gas forms a vortex ring and stays segregated from the rest of the gas in the bed.

As an example, if the bubble rises 25 times as fast as the emulsions gas [not all that uncommon because this ratio is 100 on more in some industrial operation], then the cloud thickness is just 2% of the bubble diameter .The second finding on single bubble is that, every rising gas bubble drags behind it a wake of solids, and it is designated by " $\propto$ " .

Where

$\propto=$

$\left[\frac{volume\ \ of\ wake}{volume\ \ of\ bubble}\right]$ ("$\propto$" $varies\ between$ 0.2 and 2.0 $depending\ on\ research\ investigation$)

Gas /Solid Contacting Regions

To develop the scenario of contacting regime, we can consider solids of size d_p in a bed of cross – sectional area "A", which is fed gas at a superficial gas velocityu_0. In order to simply the equation, it is necessary to define two dimensionless quantities, ie,

$$d_p^* = d_p\left[\frac{\rho_g\left(\rho_s-\rho_g\right)g}{\mu^2}\right]^{1/3}$$

$$u^* = u\left[\frac{\rho_g^2}{\mu(\rho_s-\rho_g)g}\right]^{1/3} = (Re_p)d_p^*$$

Minimum Fluidizing Velocity

The solids will be suspended, when the pressure drop exceeds the weight of solids. This happens, when the gas velocity exceeds the minimum fluidizing velocity u_{mf} and it is expressed as,

$$150\left[1-\varepsilon_{mf}\right]u_{mf}^* + 1.75(u_{mf}^*)^2 d_p^* = \varepsilon_{mf}\,(d_p^*)^2$$

Terminal Velocities (Ut)

Individual particles are below out the bed, when the gas velocity exceeds, then it is referred to "terminal velocity" (u^t).

Terminal velocity for spherical particles is expressed as;

$$u_t^* = \left[\frac{18}{\left(d_p^*\right)^2} + \frac{(0.591)}{\left(d_p^*\right)^{1/2}}\right]^{-1}$$

and for irregular shaped particles of sphericityφ_s ,

$$u_t^* = \left[\frac{18}{\left(d_p^*\right)^2} + \frac{(2.335-1.744)\varphi_s}{\left(d_p^*\right)^{1/2}}\right]^{-1}$$

When the particle sphericity"φ_s" is defined as the ratio between the surface of the sphere and surface of a particle at the same volume.

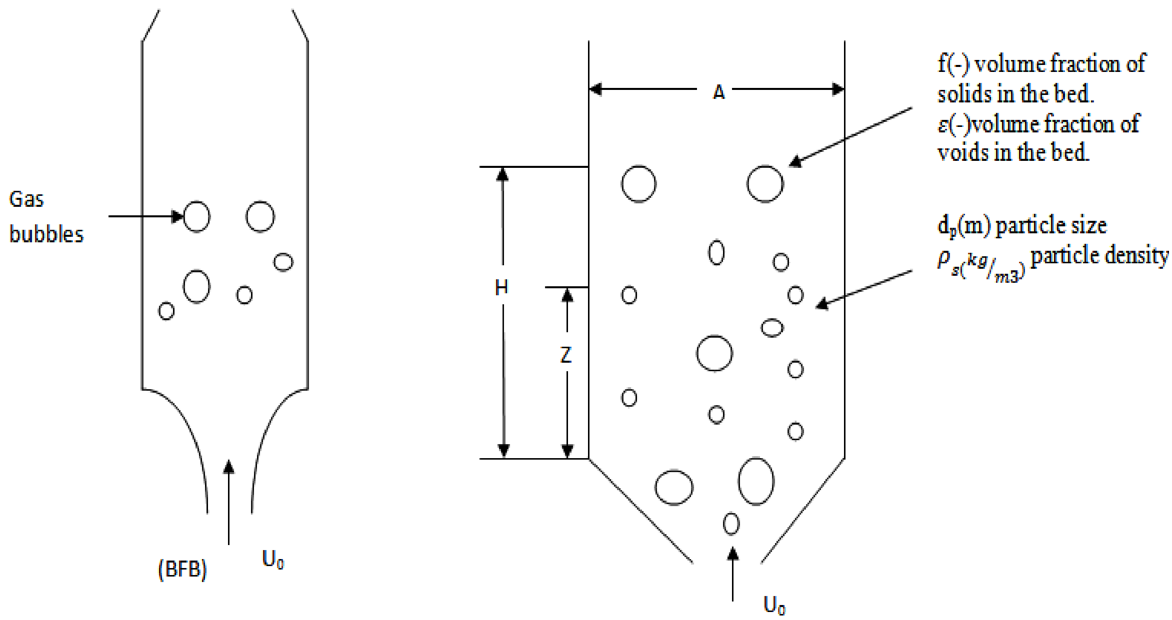
φ_s = particle sphericity $=\left[\frac{surface\ \ of\ a\ sphere}{surface\ \ of\ a\ particle}\right]$

For fine particles, we can evaluate the size by screen analysis, which gives d_{scr}. Actually there is no relationship between $d_{p.}$

Pressure Drop Concentration Issues

- $d_p = \varphi_s . d_{scr}$ =>for irregular particles with no seeming longer or shorter dimension
- $d_p \cong d_{scr}$ =>for irregular particles with one some what longer dimension but with length ratio not greater than 2:1[for example eggs]
- $d_p = \varphi_s{}^2 . d_{scr}$ =>for irregular particles with one shorter dimension but with length ratio not less than 1:2[for example pillows]

Although a single particle will be entertained by a stream of gas flowing than (u_t), this finding does not extend to a fluidized bed of particles. In BFB, the gas velocity can be many times greater than ((u_t), with very little carryover of solids. Therefore, the single particles terminal velocity is not very useful in estimating, when entertainment of solids will become appreciable.



CONCLUSION

- Expressions developed in this model shows that if ε_{mf} is known, then we can estimate α and we can measure u_{mf} and u_0 , then all the flow quantities and regional volumes can be determined interms of one parameter, i.e., the bubble size.
- The use of this model to calculate the chemical reactor behaviour is direct one and the special feature of this model is that its one parameter can be tested against what is measured and what is observed
- Various other hydrodynamic models have been proposed recently, using other combination of assumption such as, charging bubble size with height in the bed, negligible bubble- cloud resistance, negligible cloud –emulsion resistance, and non spherical bubbles.
- Models which do not allow for different bubbles sizes at given imposed bed conditions certainly cannot be adequate.
- It is also noticed that this type of model can be tested, and it can be shown to be not fair enough model, because of its bubble size and it can be compared with observations, so that we come to a conclusion about the fairness of this model towards sustainable development by considering extended wake and reacting wake.

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