Preliminary Studies on the Operating Flow Conditions of a Designed and Fabricated Fluid Catalytic Cracking (FCC) Reactor Systems

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Abstract
The philosophies on the circulating fluidized bed (CFB) riser reactors continue to be subject of investigation because of its many importance and wider phenomenon in fluidisation technology. The application of CFB as a parent cycle to fluid catalytic cracking (FCC) reactor in the oil and gas refining cannot be over-emphasised, because CFB offers critical advantages in the combustion of solid fuels, having higher fuel flexibility and the challenges to controlling the combustion temperature is very effective. Therefore, CFB process allows for near isothermal operation which is a resound practice that is good to FCC systems. FCC units operation involves conversion of high molecular weight gas oils (heavy gasoil) or residuum stocks from the crude distillation unit (CDU) into more useful, lighter and higher octane hydrocarbon products within a few second.
In this study, the CFB rig comprises of the feed injection system, a cylindrical riser of 4.0m height with inner diameter of 0.06 m, regeneration-disengager unit were modelled and constructed and operated under recycle flow situation. The developed CFB riser rig was made of galvanized steel which has high temperature and corrosive resistivity. A classical (mathematical) modelling methodology was used to study the hydrodynamics of the FCC riser process operating conditions. The bed porosity, maximum ΔP, solid mass concentration and minimum fluidizing velocity were studied with respect to the riser height. It was found that the minimum fluidisation, U_{mf} was 398 km/s at a solid-holdup (mass concentration), G_s of 7581 kg/m^{2}s, while the optimum pressure, P_{opt} was obtained to be 1.20kN/m^2.

Key words: *CFB, FCC, minimum fluidization, mass concentration and optimum pressure.*

1.0 Introduction

In the late seventies, the studies on circulating fluidized bed (CFB) reactors has been a subject of research in order to continuously enhance the performance of industrial processes such as CFB combustion and fluid-catalytic cracking (FCC) processes. The CFB is an advantageous alternative for combustion of solid fuels, because the fuel flexibility is high and the good possibility to control the combustion temperature. Due to the significances of CFB to the industry and their complex fluid dynamics, more researches on CFBs are being reported in the open literature.

The performance of a CFB boiler is influenced by the mixing of gas and particles. A high mixing rate contributes to an effective distribution of reactants, whereas an insufficient mixing can lead to hydrocarbon and CO-emissions. Therefore, an adequate understanding of the mixing behaviour is important to ensure a high combustion efficiency and emission control. Knowledge of the mixing characteristics is also useful for validation of computer simulations of CFB risers. It is obvious that the study of solid concentration and particle velocity in a CFB riser system remains of great interest in the technology of FCC unit improvement (Idris and Burn, 2008). This is because FCC units have been considered the most successful applications of CFBs all over the world in recent decades. Records show that about 46% of the global gasoline productions come either directly from FCC units or indirectly from combination with downstream units, such as alkylation. The feed injection system is by far the most critical breakthrough of modern FCC reactor design. Four recent developments have made the feed injection system increasingly important, they are:

(a) Due to the development of a highly active zeolite FCC catalyst, the reaction time has been shortened to a few seconds in the modern riser reactor.

(b) The regenerator temperature is getting higher to achieve more complete catalyst regeneration. The typical modern riser top temperature is in the range of 510 to 566ºC, but typical regenerated catalyst temperature is much higher, in the range of 677 to 760ºC. Feed injection reduces thermal cracking reactions by cooling off the lower riser quickly through fast mixing and vaporization of the feed.

(c) Feed injection reduces thermal cracking reactions by cooling off the lower riser quickly through fast mixing and vaporization of the feed. The typical mixing zone length from feed injection nozzle is given as 1.2 to 1.5m; gasoil vapourisation temperature is 733 K, while the gasoil heat of vapourisation 155 kJ/kg (Souza et al., 2011).

(d) Lastly, FCC feedstock is getting heavier, and this makes feed vaporization more difficult (Chen et al., 2006).

Globally, human activities remain on the increase, which has caused the ozone layer depletion and in turn global warming as a result of the emission of certain hydrocarbons (HCs) and their sister pollutants from the petroleum industry, these possess serious threat to the survival of life on the planet earth. An excellent performance of a CFB riser and regeneration systems are influenced by the quality of mixing of the constituted feed into the injection units. A good mixing rate contributes to an effective distribution of reactants, whereas an insufficient mixing can lead to hydrocarbon and CO-emissions, these pose a serious threat. Therefore, an adequate understanding of the mixing behaviour is important to ensure a high combustion efficiency and emission control.

The objective of this study was the application of AutoCAD14 in the design and the fabrication of prototype CFB/FCC process systems with the catalyst injection at perpendicularly/and inclined to the feed. Secondly, to apply classical modelling (relevant equations form the classical physics) approaches to study the flow dynamics on the CFB process along the riser length.

2.0 Background Studies on Fluidisation

When a packed bed of particles is subjected to a sufficient high upward flow of fluid (gas or liquid and/or solid) the weight of the particles is supported by the drag force exerted by the fluid on the particles and the particles become freely suspended or fluidised or fluid-like flow (Jahnig et al., 1980). The behaviour of fluidized suspension is similar in many aspects to that of a pure liquid. Mass transfer and heat transfer rates between particles and submerged objects (e.g. heat exchanger tubes) is greatly enhanced in fluidized beds. In addition, rapid particle mixing allows uniformity in bed. As a result, fluidised bed are widely used for conducting gas solid reactions (coal combustion), gas solid catalytic reactions (catalytic cracking of petroleum), etc. Several applications also utilize liquid fluidized beds e.g. in bioreactors (Smith and Harriot, 1985). The introduction of gas-solid system for coal gasification by Winkler in 1920s was a major breakthrough, in fluidisation technology as regards to first fluidise bed reactor.

2.1 Fluidized beds

Fluidized beds are reactors in which fluidization of particulate solids takes place. There are several types and geometries of fluidized beds but most of them have some key components: a plenum, a distributor, a bed region, and a freeboard region. The plenum is where the fluid enters the bed. Fluid next passes through a distributor or aeration plate, which uniformly distributes the fluid at the base of the bed. The particulate solid is located above the distributor in the bed region. Finally, located above the bed chamber is the freeboard region, which contains particles that have been ejected from the bed. Figure 1 depicts the schematic diagram of a typical fluidized bed riser system (Crowe, 2006).

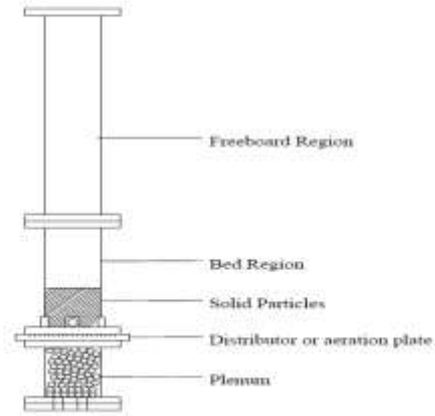

Figure 1: Schematic diagram of a fluidized bed (Crowe, 2006)

FCC converts low value heavy components of crude oil into a variety of high value lighter products. The second category is the gas-phase reaction using solids as heat carriers. In this reaction both the reactants and the products are in the same phase (gaseous) but solids are required to produce or carry the heat needed for the reaction. The third category is the gas-solid reaction, where reactants and products are gases and solids, with the option of being gases or a combination of gas and solids. Combustion and gasification are examples of processes using this type of reaction (Escurado and Heindel, 2011), (Squires et al., 1986).

Finally, the last category is where no chemical reactions occur, e.g. ceramic and pharmaceutical industries etc. Fluidized bed drying applications are an example of this type and are used due to the fluidized bed high drying rates, high thermal efficiency and lower costs; they are commonly used among the chemical, food, ceramic, and pharmaceutical industries (Escurado and Heindel, 2011).

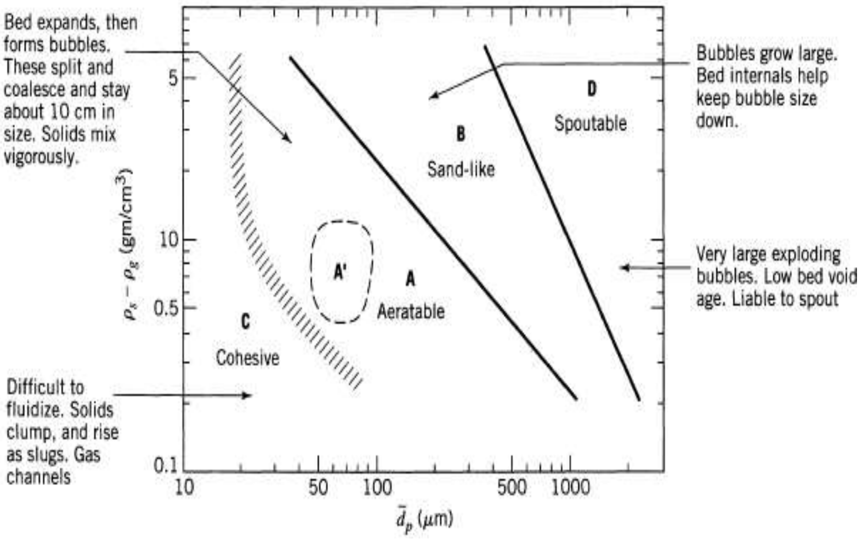

Figure 2: Geldart classification of solids (Levenspiel, 2007)

2.1.1 Behaviour of suspended solids

Geldart (1973) and Geldart and Abrahamson (1978) simply classification of solids which we now call the Geldart classification, thus Geldart A, B, C, D. These are shown and described in Figure 2 above. In considering the distribution of solids in the vertical vessel, let *f* be the volume fraction of solids at height *z* of the riser vessel.

From Figure 3, at higher and higher gas velocities the solids spread throughout the vessel. According to Geldart classification, Group A solids are characterized by bed expansion, which forms bubbles, splits and coalesce as solids mix vigorously. Group B solids witness large bubble growth, only for bed internals to help keep bubble size down. The Group C solids have difficulty to fluidize, and gas channels; whereas, Group D witness very large exploding bubbles Kunii and Levenspiel (1999). Figure 3 represent the distribution of solids in the various contacting regimes (Levenspiel, 2007).

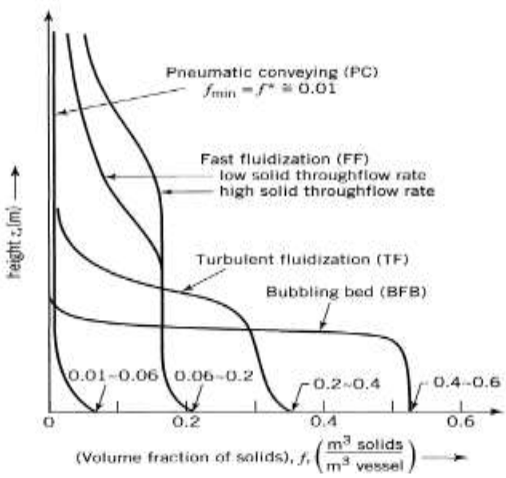

Figure 3: Distribution of solids in the various contacting regimes (Levenspiel, 2007)

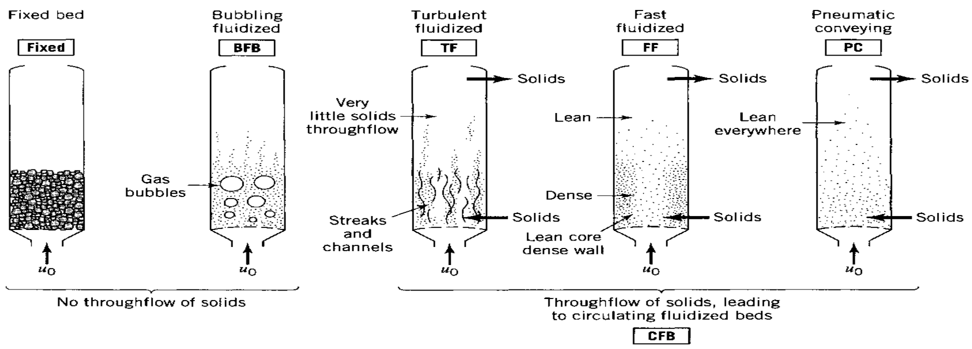

Figure 4: Fluidization regimes in a gas-solid fluidized bed (Kunii and Levenspiel, 1999)

2.1.2 Fluidization regimes

Yang (2003) considered at least six different fluidization regimes for gas-solid fluidized beds: fixed bed (FB), bubbling fluidization (BF), slugging fluidization (SF), turbulent fluidization (TF), fast fluidization (FF), and pneumatic conveying (PC). In the FB regime the air flowing across the particle does not have enough velocity to move the particles. As the superficial gas velocity (Ug) increases, the system reaches the bubbling fluidization regime.

Yang (2003) considered that the slugging regime appears in beds where the bed height (H) over the bed diameter (D) is larger than 2. This requirement ensures that bubbles have enough time to coalesce in bigger bubbles called slugs, when the bubbles grow to 2/3 of the bed diameter the system enters to a slugging regime. Finally, the pneumatic conveying regime is reached when the superficial gas velocity is much higher than the transport velocity; this regime is characterized by the particle being transported out of the bed in a dilute phase. Figure 4 depict the fluidisation regime in a gas-solid flow fluidised bed (Crowe, 2006).

2.1.3 Gas hold up

In fluidized bed hydrodynamics, gas holdup is one very important parameter that characterizes the fluidization quality, homogenous mixing, and process efficiency in a fluidization system, and is defined as the volume fraction of gas present within the bed material. Using an optical probe, (Zhu et al., 2008) determined the solid concentration (the inverse of gas holdup) in a gas-solid system for bubbling and turbulent fluidization regimes. Results show that the turbulent regime solid concentrations are not uniform in the axial and radial direction, showing a non-uniformity of the flow structure. Results showed that increasing the static bed height produced an increase in the solid concentrations mainly in the central region of the bed, while the wall region had no significant changes. This phenomenon is attributed to the increased presence of bubbles in the material as the bed height is increased (Zhu et al., 2008), (McCabe et al., 1980) and (McCabe et al., 1985).

2.1.4 Minimum fluidization velocity (U_{mf})

The minimum fluidization velocity (U_{mf}) is the point of transition between a fixed bed regime and a bubbling regime in a fluidized bed. Minimum fluidization velocity is one of the most important normalized parameters when characterizing the hydrodynamics in a fluidized bed. The point of transition between a fixed bed regime and a bubbling regime is denoted by a constant pressure line in a plot of pressure vs. superficial gas velocity; this point marked the minimum fluidization velocity. In the voidage method, the minimum fluidization velocity is determined when the voidage inside the bed starts to increase due to bed expansion as the superficial gas velocity is increased. However, this method is not commonly used because it is much more complicated to locate the point where bed expansion starts (Escudero and

Heindel, 2010). Sau et al. (2007) determined the minimum fluidization velocity for a gas-solid system in a tapered fluidized bed (conical fluidized bed) and studied the effects that bed geometry, specifically the tapered angle, had on the minimum fluidization velocity. They used three different angles (4.61^0, 7.47^0, and 9.52^0) to observe their effects on minimum fluidization velocity. Results showed that as the tapered angle increased, Umf also increased, which implied a dependence of the minimum fluidization velocity on the geometry of the fluidized bed. The pressure at the base of a fluid (due to fluid weight) comes from the static component of Bernoulli's equation. Where L is fluid height, g is the acceleration due to gravity and ρ is the fluid density. Thus:

$$\Delta P = L[\rho_s - \rho]g[1 - \varepsilon] \tag{1}$$

Where ρ_s is the true solid density (kgm^{-3}) and ε is the bed voidage. This developed to Darcy's law as represented in Equation (2):

$$\frac{\Delta P}{L} = \left[\frac{K}{\mu}\right]\left[\frac{dV}{dt}\right]\left[\frac{1}{A}\right] \tag{2}$$

Where V is the volume of fluid flowing in time, t; K is The Kozeny Constant or bed permeability. Note: K=f (ε); and remembering that fluidization occurs when the bed weight (per unit area) equals the fluid drag given in Equations (3) - (4):

$$U_{mf} = \left[\frac{K}{\mu}\right][\rho_s - \rho]g[1 - \varepsilon] \tag{3}$$

Where U_{mf} is the minimum fluidizing velocity. Note that the U_{mf} is a superficial velocity (not interstitial velocity). For spherical particles, an alternative equation for U_{mf} is given by:

$$U_{mf} = \frac{[\rho_s - \rho]g\varepsilon^3 xSV^2}{180[1-\varepsilon]\mu} \tag{4}$$

Where *xSV* is the Sauter mean diameter for the particle flow distribution

2.2 CFB riser reactors

CFB allows for near isothermal operation via the relatively large heat capacity of the recirculated solids provides outstanding gas-solids contacting because of the slip factor (φ) between the two phases. The slip factor is defined as „the ratio of interstitial gas velocity over the solids velocity". In fully developed zone of a tall riser, the overall slip is approximately 2 (Berruti et al., 1995). Three distinct hydrodynamic regimes in CFB riser are the *turbulent* bed, *fast fluidized* and the *pneumatic conveying* regimes (Levenspiel, 2007). In Figure 5, the surface of the dense bed fades and solids are found increasingly in the lean region above the dense bed. The concentration of solids in the upper lean region can be reasonable represented by an exponential decay function which starts from the value in the lower region f_d and fall to $f*$ the limiting value in an infinitely high vessel. This is the value for pneumatic conveying (Levenspiel, 2007).

2.3 FCCU process description

FCCU is one of the most important conversion processes used in the refining industry. During the catalysed cracking reactions, the heavier molecules are cracked to lighter components e.g. liquefied petroleum gas (LPG), dry gas, gasoline, diesel etc. and coke formation. When sufficient heat, especially in the presence of a catalyst, is applied to paraffin in hydrocarbon, it breaks into two or more fragments and one of the components is always an olefin. The size of the fragment produced usually includes all the possibilities so that the product is a mixture. The three sections of FCC unit are: reactor and regenerator, main fractionation and gas concentration section. This study focuses on the reactor and regeneration section.

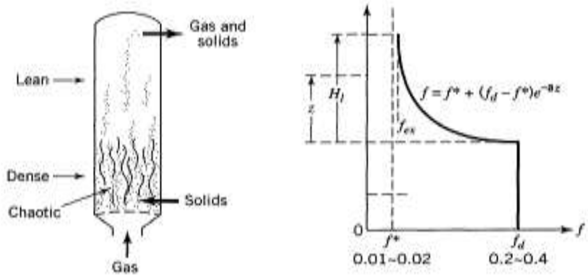

Figure 5: The turbulent bed and its solid distribution (Levenspiel, 2007)

2.3.1 Reactor section

This consists of the riser system, disengager and the regenerator. Fresh feed of vacuum gas oil (VGO) and heavy gas oil (HGO) and recycled cycle oil are injected into the riser bottom; these are known as combined feed. The feed enter the riser together with a controlled amount of regenerated catalyst, as shown in Figure 6.

The hot regenerated catalyst vaporizes the feed and the resultant vapours entrain the catalyst upwards through the riser with a minimum back mixing into the disengager section. The gas oil cracks immediately when it contacts the catalyst. In the disengager, an abrupt separation of the cracked vapour and catalyst takes place by the use of cyclones.

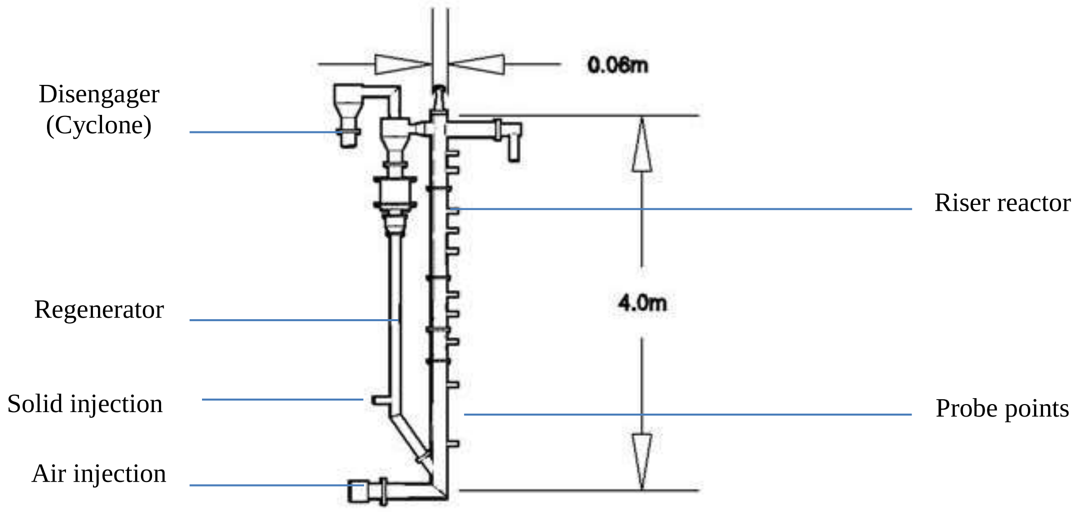

Figure 6: AUTOCAD14 Design of inclined CFB process injection system

The reactor effluent is directed to the FCC main fractionator for resolution into gaseous light olefin co-products, FCC gasoline and cycle stocks. During the cracking reaction, a carbonaceous by products is deposited on the circulating catalyst. This material called 'coke' is continuously burnt off the catalyst in the regenerator. The energy carried by the hot regenerated catalyst is used to satisfy the thermal requirement of the cracking section of the unit (Idris et al., 2007).

2.3.2 Regenerator section

Depending on the specific application, the regenerator may be operated at conditions that achieve a complete or partial internal combustion of CO to CO_2 or alternatively, CO may be converted to CO_2 in an external CO boiler. If internal conversion of CO to CO_2 is practiced, the sensible heat of the flue gas can be recovered in a waste heat boiler. Flue gas is directed through cyclones separator to minimize catalyst entrainment prior to discharge from the regenerator. Figure 7 is the constructed CFB riser-regenerator systems.

Figure 7: 2-sections pictorial representation of CFB/FCC riser-regenerator systems awaiting installations

Figure 8 depicts the panel controller devices for the CFB riser-regenerator systems operations. The CFB rig can be found at the technology hall of Chemical Engineering Department, Faculty of Engineering, University of Maiduguri, Nigeria.

Figure 8: Panel controller for the CFB riser-regenerator systems awaiting installations

3.0 Methodology

3.1 Equipment and material selections

The choice of suitable materials of construction is fundamental to the success of product or equipment design. In order to ensure safety, durability and optimize performance, it is necessary to consider certain properties of materials such as physical properties, chemical properties as well as thermodynamic properties. AutoCAD14 was used in the design of the CFB riser reactor as shown in Figure 6.

3.2 Material used for constructing the CFB system

Galvanized steel was used for the construction of the CFB process injection system. The cylindrical riser is 4.0 m height and inner diameter of 0.06 m. Galvanized steel has good machining property, ductility, formability, weld ability, and high temperature resistivity. It has a matte-gray appearance, and resists corrosion up to one hundred (100) times better than uncoated steel (Figures 7 and 8).

3.3 Model equations

The modelling equations essentially relate the maximum pressure drop (ΔP_{max}) in a riser reactor, minimum fluidizing velocity (U_{mf}), solid mass concentration (Gs) and bed voidage (ε_{pf}) to the bed height (L), φ is a viscosity property and U_{go} is the initial gas velocity. To determine the relationship between the solid mass flux (solid mass concentration), G_s and the riser height h, the Equation (5) is given below:

$$G_s = \frac{\rho_s U_{go}}{\varphi}\left[\frac{1}{\varepsilon_{pf}} - 1\right] \tag{5}$$

Similarly, the maximum possible ΔP in the riser, with respect to the bed voidage fraction (bed porosity, ε) along the height is expressed using Equation (6) as given below:

$$\Delta P_{r,max} = \rho_s(1 - \varepsilon_{dense})H_{riser} \tag{6}$$

The in relation to the bed porosity with respect to riser height was shown in Equation (4) above.

3.4 Standards used in the CFB/FCC Design Operating Conditions

The standard conditions of FCC operation used are presented in Table 1 below.

Table 1: FCC design operating conditions used

S/No.	Parameters	Units	Values
1.	Operating pressure for the FCC conditions, P	kPa	284
2.	Temperature ranges in the riser, T	^{0}C	510 – 565.6
3.	Temperature ranges in the regenerator, T	^{0}C	676.7 – 732.6
4.	Particle (solid) density, ρ	kg/m^3	1400
5.	Bed voidage ranges, ε	dimensionless	0.95 – 0.99
6.	Maximum riser height, L	m	4.0
7.	Ideal air density, ρ_{air}	kg/m^3	1.2
8.	Fluid viscosity, μ	Ns/m^2	1.81x10^{-5}

4.0 Results and Discussion

The modelling results is represented in Table 2 below

Table 2: Determined modelling results using classical approach

L(m)	**Gs (kg/m^2.s)**	**ρ(kg/m^3)**	**ΔP (N/m^2)**	**ε_{pf}**	**U_{mf} (m/s)**
0.0	7581.325	1.20	0.00	0.950	397.366
0.4	7581.318	1.20	278.569	0.951	397.366
0.9	7581.329	1.20	601.198	0.953	397.366
1.4	7581.325	1.20	795.913	0.960	397.366
2.1	7581.319	1.20	1074.482	0.964	397.366
2.8	7581.321	1.20	**1193.869**	0.970	397.366
3.2	7581.325	1.20	909.614	0.975	397.366
3.6	7581.325	1.20	1023.316	0.980	397.366
4.0	7581.229	1.20	586.509	0.990	**397.366**

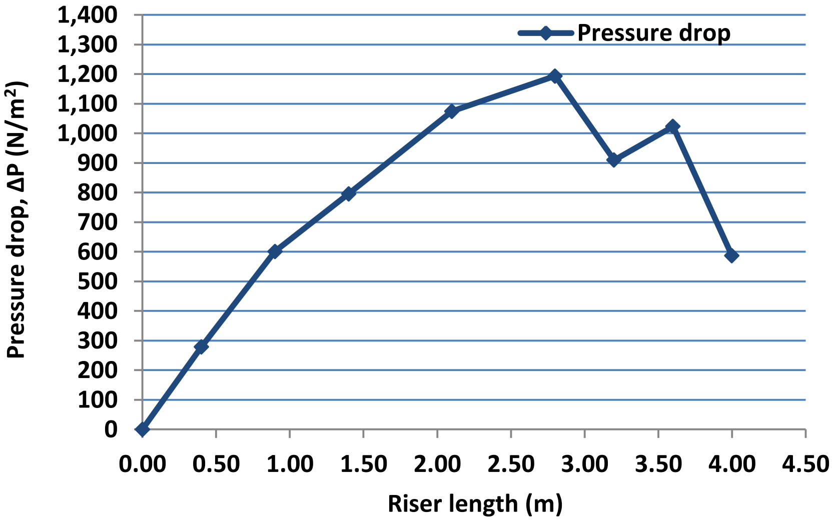

Figure 9: Profile of maximum pressure drop with respect to riser height (m)

Table 1 show that there is a corresponding *relation* between G_s and U_{mf}. This behaviour is one the basic philosophy behind the clarification on the flow dynamics for gas-solid flow phenomenon.

The detail relationship between riser bed heights, solids mass concentration, minimum fluidizing velocity, pressure drop (ΔP) and bed voidage (porosity) has been presented in this section, because of the hydrodynamic importance of gas-solid flow phenomena in the CFB riser systems. From Figure 9, it is observed that the CFB riser systems can be divided into two zones namely; (i) a dense zone at the bottom and (ii) a dilute zone at the top of the riser. This is in line with the report of Zhu et al. (2008), Idris and Burn (2008) and Idris (2008).

As can be seen from the classical predicted results shown above, the void fraction (solid fraction or voidage) in the riser, when operating in the S-shaped profile (dilute phase) of the pneumatic conveying regime is a function of minimum fluidizing velocity (superficial gas velocity). Thus, the maximum pressure that can be generated in a riser of fixed height is also a function of Umf when the riser is completely filled with a dense bed, (Aberuagba et al., (2005) as shown in Figure 10. The results of this profile indicate that solids mass concentration, Gs (kg/m^{2}s) does not show any significant change with height. The G_s was uniformly high at the bottom (dense region) of the bed, slightly declined and almost became stable at riser top. This is due to high concentration of solids at both the bottom and top of the riser. The ΔP is as a result of the porosity of the bed; the higher the porosity of a bed, the magnitude of the drop in pressure and vice versa (David, 2010).

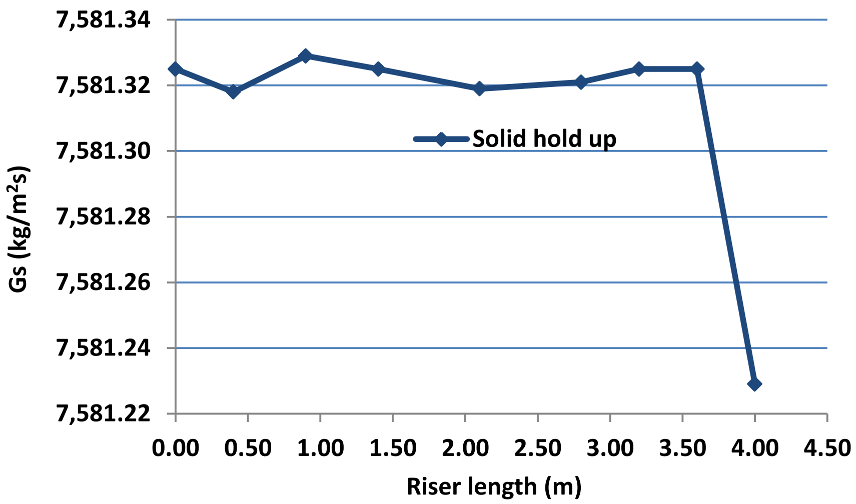

Figure 10: Profile of solid mass concentration with respect to riser height (m)

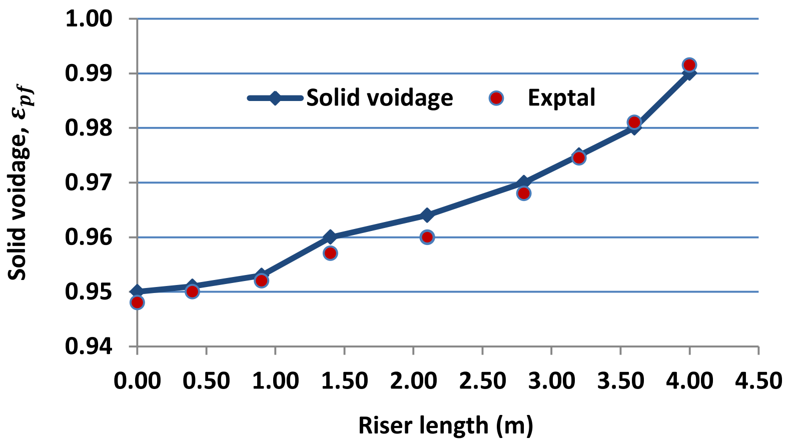

Figure 11: Profile of bed voidage with respect to riser height (m)

Figure 11 depicts the profiles of solid voidage, εpf along the riser length. Practically, the figure has shown that the voidage increases within a value of 0.95 to 0.99 along the riser length. This is in consonance with the fact that when we have quality mixing of the fluidising gas-solid, and having a resound solid concentration at a minimum pressure range. However, keeping a moderate pressure range provides also a moderate solid holdup. Therefore, Figure 11 would be a base-guide to further give some ideas on the discrete studies in gas-solid hydrodynamics which were captured in CFD today (Idris, 2008; Idris and Burn, 2008).

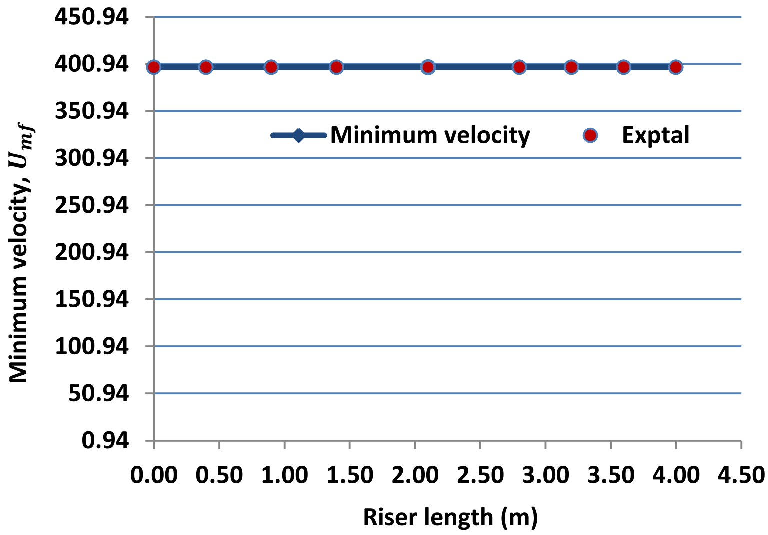

Figure 12: Profile of U_{mf} with respect to riser height (m)

In comparison of the calculated solid voidage with the experimental data, it was observed that there was a close argument between them, as shown in Figure 11. From Figure 12, the minimum fluidizing velocity, Umf remain constant with respect to the height (m) of the riser, while pressure drop (ΔP) rose steadily from the bottom but suddenly began to fall as we approach the peak. The Umf remain constant because the fluid is been considered and treated as a *mono-dispersed*. The ΔP increases across the bed length as the bed is fluidized and drop profusely as fluid proceed toward the riser top, Idris and Alan (2008).

The axial particle distribution in CFB is one of the most important phenomena that equally require close understanding in riser gas-solid hydrodynamics. The predicted ΔP shows an increase in magnitude with respect to increasing bed height along the CFB, are functions of the uniform density of the flowing fluid as depicted in Figure 13. A safe process operation using appropriate operating pressure is the best way to control the effect on gas-solid density in the riser systems. These effects are as result of bulk density and mass of the flowing material. With an increase in height, the solids flow becomes fully developed and moves up with a constant velocity. Ideally for a fully developed *single* flowing fluid, the length is usually from 1.20 m or thereabout. When the flowing gas-solid hydrodynamics are compared with the riser bed height, the results clearly showed that the material density has a define effect on minimum fluidization velocity, Umf. These results imply that a denser material requires more bed pressure force to equalize the gravity force of the bed in order to achieve fluidization (David, 2010; Idris et al., 2007; Escurado and Heindel, 2011).

The turbulent regime of solid concentration is a non-uniform flow in the axial direction, an indication of non-uniformity of the flow structure. This phenomenon is attributed to the increase presence of bubbles in the reactor as the bed height is raised. An interaction between turbulence and collisions could have been responsible for cluster formation and disintegration, and a lateral equilibrium with more solids in the annulus than in the core. The solids mean size in the downward flow was bigger than the average size in the total CFB loop operations. The graphs show symmetrical behaviour which could be attributed to the uniform injection of solids at the entry base of the riser. It also demonstrates that it is imperative to traverse the entire column when taking profile samples. The match between the data and the predicted profile is quite appreciative as depicted from Figures 11 to 13. Industrially, large solids mass fluxes are employed and are of interest to this study (Pugsley, 1995). Overall, the results achieved in this project work are in substantial agreement with previous work as presented in Zhu et al. (2008), and Idris and Burn (2008). There are, however, slight deviations which could be attributed to the limitations on the classical approach method adopted in this work.

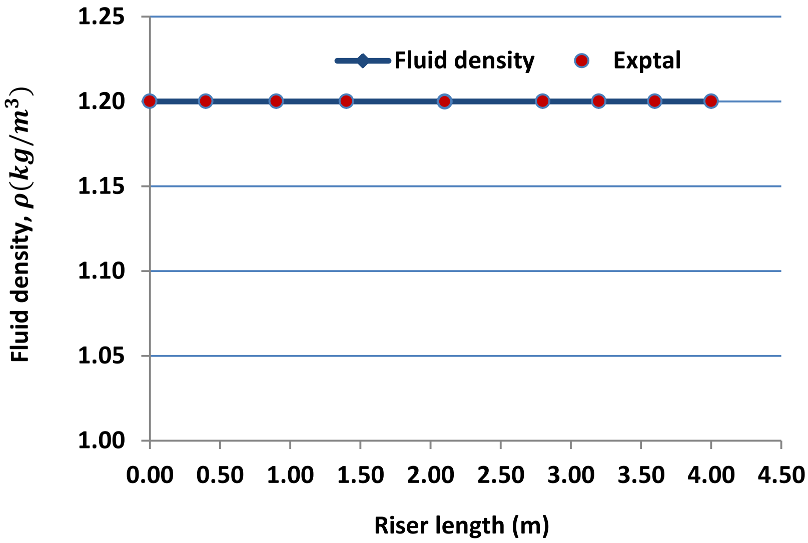

Figure 13: Profile of fluid density with respect to riser height (m)

5.0 Conclusions

Within the limit of the classical modelling methodology used in this study, the nature of CFB/FCC riser hydrodynamics was evaluated. The relationship between the solids mass flux, bed voidage (porosity), pressure drop and minimum fluidizing velocity profile and the riser height was attempted. However, the best operating conditions for FCC riser is the pneumatic regime, that is, near plug flow situation. In this study, a ‘plug flow’ condition of operation was targeted; but fortunately the downflow of gas and solids in the annulus caused the average solids fraction in the fully developed zone to be altered significantly from the ‘plug flow’ solids fraction. In addition to this limitation, the effect of the back-mixing of gas would lower conversion in a riser reactor for the fast regime reaction. The match between the experimental reported data and the predicted profile is quite reasonable within the limit. Industrially, large solids mass fluxes are employed and are of interest to this study. In conclusion, the bed porosity, maximum ΔP, solid mass concentration and minimum fluidizing velocity were studied with respect to the riser height. The minimum fluidisation, Umf was 398 km/s at a solid-holdup, Gs of 7581 kg/m^{2}s, while the optimum pressure, Popt was obtained at 1.20kN/m^2. The results achieved in this study are in agreement with previous work reported in open literature. However, slight deviations which could be attributed to the limitations form the CFB modelling methodology used, which can now been addressed using computational fluid dynamics (CFD).

Recommendations

Since we have established the fact that approaching closer to *plug flow* regime would give the designer better control of product distribution, and allowed production of a larger fraction of desired products. Because increasing the production of desired product by 1% would increase the profit per reactor by $1 million to $2 million per year (Levenspiel, 2007). Secondly, hydrodynamic models alone cannot completely describe the riser reactor mechanism; hence the coupling of the hydrodynamic and kinetic model will go a long way in describing the model. More intensive studies using the application of CFD would yield better results in studying the hydrodynamics, heat transfer effect and kinetic reactions involved. More so, the flow pattern can be properly predicted and validated if the radial flow is also considered, hence hydrodynamic flow in both the axial and radial flow shall be considered in the subsequent study.

List of Symbols and Abbreviations

G	=	gravitational acceleration (m/s^2)
Gs	=	solid mass flux (kg/m^2s)
Hriser	=	height of riser (m)
T	=	temperature (K)
U_{go}	=	gas velocity (m/s)
ρ	=	gas (air) density (kg/m^3)
ρ_s	=	density of sand (kg/m^3)
φ	=	slip factor
μ	=	viscosity of air (Ns/m^2)
ε_{pf}	=	bed voidage at plug flow condition
$\Delta P_{r,max}$	=	maximum pressure drop (N/m^2)
xSV	=	Sauter mean diameter for the particle
U_{mf}	=	minimum fluidizing velocity (m/s)

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References

Berruti F, Chaouki J, Godfroy L, Pugsley T S, Patience G S (1995). 'Hydrodynamics of Circulating Fluidized Bed', pp. 2 – 7

Chen X, Hong W, Zhang M (2006), 'Hydrodynamic characteristics of spout-fluid bed: pressure drop and minimum spouting/spout fluidizing velocity', Chem. Eng. J. 118(1-2):37-46.

Crowe R (2006), 'Simulation of a Fluidized Bed Simulation of a Bubbling Fluidized Bed', Particle Technology and Fluidisation. Pp. 3 -10

David R (2010). 'Bed Height and Material Density Effects on Fluidized Bed Hydrodynamics', Iowa State University. Pp. 4 - 12.

Escudero D, Heindel TJ (2010), 'Bed height and material density effects on fluidized bed hydrodynamics', Chem. Eng. Sci. 66(16):3648-3655.

Geldart D (1973), 'Types of gas fluidization', Powder Technol. 7(5): 285-292.

Geldart D, Abrahamson AR (1978), 'Commercial experience with synthol CFB Reactor – in CFB Technology', Elsevier book company. Pp. 18 – 56.

Idris M. N. (2008). 'Hydrodynamic studies on an injection system as applicable to the modern circulating fluidized bed (CFB) riser using computational fluid dynamics (CFD)'. Int. Pet. Train. J. 5:1.

Idris M. N, Burn A., (2008), 'Hydrodynamics Studies on an Injection System as Applicable to Modern Circulating Fluidized Bed (CFB) Riser using Computational Fluid Dynamics (CFD)', AIChE centenary journal, pg. 36-46.

Idris M. N, Mahmud T, Gibbs B (2007), 'Application of Process hydrodynamics and design of a robust FCC Riser Reactor', Int. Pet. Train. J. 4:1.

Jahnig C E, Campbell D L, Martin H A (1980), 'History of fluidized solids development at EXXON. Fluidization', New York, Plenum Press pp. 3-24.

Kunii D and Levenspiel O (1999), 'Fluidisation Technology', Elsevier Book Company. Pp. 3 – 78.

Levenspiel O (2007), 'Chemical Reaction Engineering', (3rd edition), Prentice Hill publication, New York, pp. 447 - 470.

McCabe WL, Smith JC, Harriot P (1980). 'Unit Operations of Chemical Engineering'. McGraw Hill, New York.

McCabe WL, Smith JC, Harriot P (1985). 'Unit Operations for Chemical Engineering'. McGraw Hill Book Company

Mustafa A (2005), 'Simulation Studies of Gas-Solid in the Riser of a Circulating Fluidized Bed', Conference Sponsors: Advanced Energy Systems ISBN: 0-7918-4183-9 FBC2005-78014, pp. 201-207. Risers: A Review". Can. J. Chem. Eng. 73:579-602.

Sau et al (2007), 'Fluidisation characteristics of binary mixtures of biomass and sand quartz in an acoustic fluidised bed', Elsevier book company. Pp. 2 – 8.

Souza J. A, Vargas J. V. C, Ordonez J C, Martignoni W. P, von Meien O. F., (2011), 'Thermodynamic optimization of fluidized catalytic cracking (FCC) units'. Elsevier Int. J. Heat Mass Transfer 54:1187-1197.

Squires A. M, Kwauk M, Avidan A. A., (1986). 'Fluid Beds: At Last, Challenging Two Entrench ed. Practices.' Science 230(4732):1329-1337.

Yang W. C (2003). 'Fluidisation flow regime', Markel Dekker New York, USA. Pp. 18 - 24

Zhu et al. (2008). 'Hydrodynamics of Bubbling Fluidised Bed Reactor', Powdered Technol. J. pp. 19 - 37