



RTD Method a means for Hydrodynamic scale up of Pressurized Fluidized Bed Gasifier (PFBG)

J.S.Rao^{1*}, N.V.S.Ramani¹, H.J.Pant² and D.N.Reddy³

¹Corporate R&D Divisin, Bharat Heavy Electricals Limited, Hyderabad, India.

²Isotope Applications Division, Bhabha Atomic Research Centre, Mumbai, India.

³J.N.T.University, Hyderabad, India.

ARTICLE INFO

Article history:

Received: 28 February 2013;

Received in revised form:

5 April 2013;

Accepted: 11 April 2013;

Keywords

Fluidized bed gasifier,
Residence time distribution,
Mean residence time,
Scale up Radiotracer technique,
Lanthanum-140,
Gamma distribution model,
Bypassing.

ABSTRACT

One of the most challenging problems encountered by a fluidized-bed designer is assessing how changes in bed geometry and operating conditions affect the gasifier performance while scaling up to demonstration / commercial size. Typically, commercial gasifier designs are based on operating experience from small pilot plants. A cold model of a gasifier represents an inexpensive and convenient platform for conducting detailed hydrodynamic studies that would otherwise be impossible in the hostile high pressure and temperature environment of fluidized bed gasifier. A perspex three dimensional semicircular cold model test rig of ID 940mm which is hydro dynamically scale down model of a demonstration plant of 168 TPD pressurized fluidized bed gasification (PFBG) plant is established and hydro dynamic parameters viz Froude no., bubble rise velocity, and bubble diameter are presented which are used for further scale up. Besides performance of the gasification process involves knowledge of dynamics of two phases viz. solid (coal) and gaseous for scale-up of the gasifier. The measurement of mean residence time (MRT) and degree of axial mixing of solid phase is required for evaluation of PFBG. The paper presents the residence time distribution (RTD) studies carried out in a pilot scale hot model of PFBG of 200 mm dia and verified in a hydro dynamically similar cold model. The coal particles labeled by radio tracer Lanthanum -140 was used to measure RTD by collimated scintillating detectors located at ash extraction points at the bottom and gas outlet at the top of the gasifier. The measured RTD data of coal / ash particles were treated and normalized for arriving at the mean residence time (MRT). The treated RTD data were simulated using gamma distribution model and found that model predicated MRTs of cold and hot model tests were in good agreement. The paper suggest the parameters which assist to minimize the bypassing of the coal particles in the gasifier thus improving the carbon conversion efficiency and hence enable scale-up of the PFBG.

© 2013 Elixir All rights reserved.

Introduction

Coal is the mainstay fuel in India for power generation. With total reserves around 220 billion tones, of which 84.4 billion tones are proven, domestic coal will continue to be the cheap and assured energy source for the next century and beyond. Indian coals have high ash content (35-50 %) with ash fusion temperature of around 1400 °C. But most of the coals have low sulphur content. Increasingly higher ash content in coals available for power plants necessitates use of newer technologies to efficiently use such domestic coals. Increasing use of coals for power generation necessitates development of advanced coal conversion technologies (CCT) to achieve higher overall efficiency thus reducing emission of CO₂ and other pollutants as well as prolonging the life of the country's coal reserves. Further, environmental regulations that may come in force in future will mandate power plants to drastically slash emission levels of gaseous and particulate pollutants.

The IGCC technology is recognized as the most efficient and environment friendly among CCT's viz : Atmospheric / Circulating Fluidized bed Combustion (AFBC/CFBC) , Pressurized Fluidized Bed Combustion (PFBC) based combined

cycle , Pulverised Coal (PC) plants using Super Critical steam parameters and Integrated Gasification Combined Cycle (IGCC) plants for 21st century. Several studies on selection of gasification technology for IGCC, made by national level committees consisting of CSIR, CEA, TIFAC, CII, NTPC etc recommended the choice of air blown PFBG technology as the best option for Indian high ash coals as compared to moving bed and entrained bed processes.

Since construction and operation of demonstration PFBG plant are costly and time consuming, it may not be feasible to undertake an expensive development testing program for establishing the PFBG process technology. Further large fluidized beds are difficult to instrument to observe detailed behaviour. A perspex three dimensional semicircular test rig reactor of ID 940 mm which is hydro dynamically scale down model of a demonstration plant of 168 tpd PFBG plant is constructed and used to conduct cold model studies. The hydrodynamic parameters viz Froude no., bubble rise velocity, and bubble diameter are presented.

Besides, dynamic similarity has been the main criteria in scaling up the higher size PFBG. The Residence Time

Distribution (RTD) is an important parameter for studying the dynamics of solid phase in both cold and hot models. In the present work the mean residence time (MRT) and degree of mixing is derived from the treated RTD curves of pilot scale hot model of 200 mm dia PFBG and verified by hydrodynamically similar cylindrical cold model. The data is useful in improving the efficiency of PFBG process by increasing the carbon conversion as well as for scale up of the process.

The present experimental study is carried out in two parts. The first part involves establishing hydrodynamic parameters of cold model of 940 mm ID semi circular test rig which is scale down of 168 tpd demonstration scale PFBG. The second part of experimental study involves carrying out RTD studies using Radiotracer technique for dynamic simulations of PFBG. The measured RTD curves are treated and normalized to derive MRT of 200 mm PFBG in cold and hot model test rigs. The experimental RTD values have been validated using simulated RTD using gamma distribution model.

Hydrodynamic Scaling

It is essential that designers have the capability to reliably scale up pilot plant performance to commercially viable levels. Hydrodynamic scaling provides a rational approach to address this scale-up issue (1,2,3,4). For example, bubbles in bubbling fluidized beds play a central role in solids mixing and in gas flow patterns through the bed. Yet it is virtually impossible to measure the characteristics of the bubbles in a hot gasifier. Therefore, hydrodynamic scaling makes it possible to conduct fast and inexpensive tests, in the laboratory, to evaluate the effects varying bed geometry and operating conditions on bed hydrodynamics.

On the basis of full scaling parameters, the cold model reactor has dimensions that are too smaller than those of the hot gasifier. Hence, it is desirable to identify a reduced set of scaling relationships that permits a scale factor to be supplied as a free parameter rather than determined by the scaling parameters themselves. Glicksman et al [5,6,7] also proposed a simplification to the full set of scaling parameters, which allow the scale factor for the bed dimension to be chosen independently. The simplification is based on the reduction on the number of dimensionless groups when either viscous or inertial effects dominate the fluid particle drag. In both viscous and inertial limits the scaling parameters reduce to groups listed below are referred as the simplified set of scaling parameters.

$$d_p^3 \rho_g (\rho_p - \rho_g) g / \mu_g^2, \rho_g / \rho_s, U_o^2 / gD, U_o / U_{mf}, L/D, \phi_s, PSD, RTD \dots [1]$$

where ϕ_s and PSD are included to match ϵ_{mf} between two fluidized beds and RTD is residence time distribution of coal particles within the gasifier/reactor.

Different coal particles take different time to come out of the reactor. The time the particles have spent in the reactor is called the residence time and the distribution of the various particles coming out the reactor with respect to time is called the residence time distribution (RTD).

A perspex three dimensional semi circular test rig reactor of ID 940 mm hydrodynamically scale down model of a demonstration size 168TPD-PFBG plant is shown in Fig.1. The simplified scaling parameters of 168 tpd PFBG hot and cold models is given at table -1.

It is found that Iso-propylene material with a density of 925 kg/m³ for cold model tests is hydrodynamically similar to bed material used in hot model test rig. The scale factor of two thirds is chosen and accordingly the bed diameter, expanded bed

height and other hydrodynamic parameters have been fixed for the cold model test rig. A comparison of hydrodynamic parameters of 168 TPD- PFBG hot and cold model is given at table 2.

Flat Plate Distributor:

Many types of distributors have been developed to improve the gas distribution in a fluidized bed. The choice of distributor is governed by the process and the operating conditions. The multi-orifice plate is the simplest gas distributor used in industries. Its ease of construction and maintenance makes it a common choice. Further the bubble-cap type of orifice have been specially designed to prevent the back-flow of solids through the distributor. In a fluidized bed the distributor plate must be designed to offer uniform fluidization through the bed cross section. Uniform fluidization is achieved only if the distributor plate imposes a resistance to the total flow sufficient to over-come the fluids inherent resistance to rearranging and redistributing itself. Hence for proceeding with design and sizing of a distributor, the pressure drop across the distributor has to be assumed. The usual practice is to assume the distributor loss as a fraction of the bed pressure loss. Many investigations show that a ratio of distributor loss to bed pressure loss of about 0.3 to 0.4 has been recommended. The flat plate distributor is shown in Fig.2. with ID of about 940mm and is provided with 10 rows, each row consisting of 16 equispaced holes of diameter 5.7 mm.

Concave Distributor:

In a fluidized bed with a concave distributor, a different flow pattern is observed. In view of different bed heights on inclined plates lead to different pressure drops across the bed more gas tends flowing through the region where bed material layer is thinner. Less gas passes through the central zone where bed material layer is thicker. As a result, acted by gravity, buoyancy and drag force, the particles in the central zone move down wards and the particles at both sides between the central zone and wall zone flow upwards rapidly. Once reaching the bed surface, most of them turn to central zone and then move downwards, the others turn to the wall zone and move downwards along the wall.

The conical distributor for the cold model is designed as shown in Fig 3 and is provided with 11 rows, each row consisting of 14 equi spaced holes of diameter 5.7 mm arranged in a zigzag manner at a pitch of 40 mm between each row. The measured hydrodynamic parameters of semi circular test rig using conical distributor are presented as (a) Froude no versus bubble rise velocity and (b) Bubble diameter versus bubble rise velocity graphically at Fig. 4 & 5 respectively.

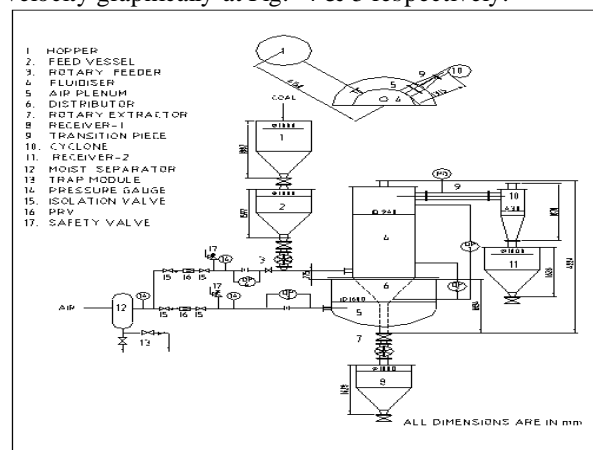


Fig.1: General Arrangement of Semi circular Test Rig

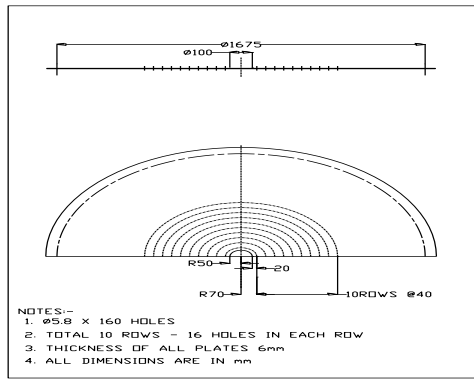


Fig. 2 . Flat Plate Distributor of Semi circular Test Rig

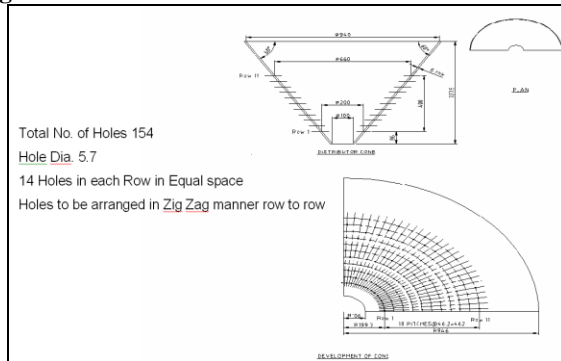


Fig 3 Schematic drawing of conical distributor

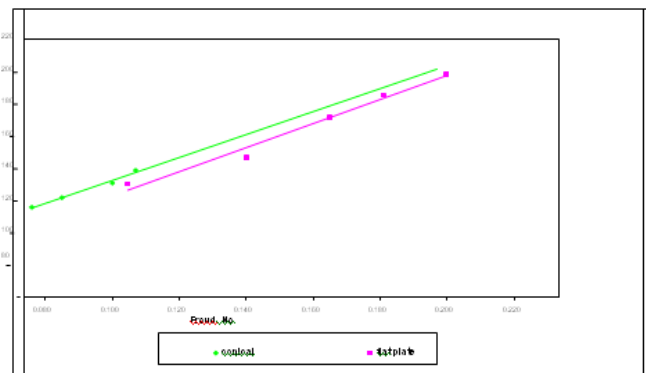


Fig. 4. Variation of Froude Number Vs Bubble Rise Velocity

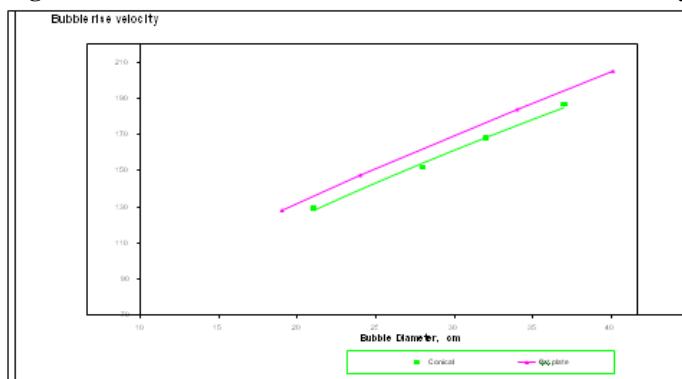


Fig. 5. Variation of Bubble Diameter with Bubble Rise Velocity

3.0 RTD Studies In PFBG Process

The use of radioisotope techniques in industrial process investigations is increasing steadily as more and more potential applications of process efficiency improvements and scale up are identified [8,9,10,11,12,13,14]. RTD is an important characteristic of continuous flow systems and provides vital

information such as MRT of process fluid, degree of axial mixing which is used to improve the process efficiency.

The concept of RTD can be applied to PFBG process as well. The knowledge of dynamics of two phases coal and fuel gas produced by chemical reactions in the bed is important to assess the performance of the gasifier as well as for scale up of the process. The main objective of the present work is to emphasize the use of RTD measurement in the gasifier design and scale up of dynamics to demonstration / commercial size PFBG.

A pilot- scale of 200 mm dia PFBG system has been set up to study gasification performance of high ash Indian coals for power generation. The simplified scaling parameters for 200 mm PFBG hot and cold models is given at Table .3.

The readers may refer to paper of Pant et al [15] for detailed description of the test rig as shown in Fig.6. Initially the gasifier is filled with a known quantity of coal particles (< 4 mm). The coal properties used as fluidizing material are given in Table 4. The comparison of operating parameters of 200 mm PFBG hot and cold models is given at Table -5.

The coal is continuously fed into the gasifier using a coal feeding system consisting of a hopper of storing capacity of about 120 kgs of sub bituminous coal and rotary valves. The rotary valves are used to measure the coal feed that is transported pneumatically into the gasifier. The gasifier consists of air plenum and conical distributor assembly made of stainless steel material. The conical distributor has 72 no of holes of 2.5 mm dia. The gasification process produces various gases which flows upward in the freeboard section and passes through the cyclone system. The separated fines in the cyclone are fed back to the gasifier using recycle system and fuel gas is fed to the wet gas cleaning system.

3.1 Radio tracer Preparation :

Earlier work of Pant et. al. [16] was carried out to establish the feasibility of adopting radiotracer technique for measurement of RTD of the coal particles in the PFBG. Also, Pant et al investigated two different radiotracers i.e. gold-198 and lanthanum-140 and concluded that both are equally suitable to trace the coal particles in gasifiers. For the present scale up study, lanthanum-140 radio isotope was selected to be used as a tracer, as it has strong affinity to get adsorbed on solid particles. Lanthanum-140 was produced by irradiating lanthanum oxide powder (La_2O_3) in DHURVA reactor at Bhabha Atomic Research Centre. The irradiated target was processed to produce lanthanum chloride (LaCl_3). About 1 mCi (37 MBq) activity of lanthanum-140 was taken from mother solution and diluted in about 300 ml of distilled water. The crushed coal of about 100 gms quantity is mixed with the diluted solution and stirred for about 5 -10 minutes using a glass rod for each experimental run. The coal soaked in the radioactive solution was left for about half an hour and was decanned and the coal particles were dried using an electrical heater till about 600-700 micro curie (220-260 MBq) activity got adsorbed on the coal particles and was used as tracer in each run.

3.2 Gamma Distribution model:

In order to simulate the experimentally measured RTD data the simple tanks-in-series model [17,18] was used. The mathematical model assumes that the system under investigation consists of a series of N number of well-mixed stirred tanks each of volume V. However the model has drawback that the value of N should be an integer and the experimentally measured data did not fit with model. This drawback was overcome by using

gamma function model which assumes that a flow system consists of N equal sized tanks of volume V and a fractional compartment of volume V_f . The response equation for The RTD function $E(t)$ can be written as [2]

$$E(t) = C(t) / \int_0^{\infty} C(t) dt \quad (2)$$

where, $C_i(t_i)$ is tracer concentration in terms of counts/minute at time interval t_i . The experimental RTD is calculated from the count rate distribution at the outlet of the system in cps or cpm. The zeroth moment of normalized RTD function ($E(t)$) gives area under the curve, which is equal to unity. The first moment of the

N is 1 for well-mixed system or continuously stirred tank reactor (CSTR) and tends to ∞ for a plug flow system (PFR). $\Gamma(N)$ is called gamma function [18,19] and is given as [6]:

$$\Gamma(N) = \int_0^{\infty} e^{-x} x^{N-1} dx \quad (6)$$

Where $x = V_f/V$ and N tank number is positive and need not to be an integer. The tank each is of volume V and a fractional compartment of volume V_f . The main use of gamma model is to fit small deviations from the exponential distribution of a single stirred tank. If $N < 1$, then this implies that the system behaves as a well-mixed system with some amount of bypassing of the process fluid. After developing or selecting a suitable model to simulate the measured RTD data, next step is to fit the model simulated RTD to the experimentally measured RTD data. The model simulated RTD curves were fitted to the experimental data using least square curve-fitting method [20]. The quality of the fit is judged by choosing the model parameters to minimize the sum of the squares of the differences between the experimental, $E(t)$ and model simulated or predicted curves, $E_m(t, N, \tau_m)$ in the form of RMS given as [7].

$$RMS = \left\{ \frac{1}{n} \int_0^{\infty} [E(t) - E_m(t, N, \tau_m)]^2 dt \right\}^{0.5} = \text{Minimum} \quad (7)$$

where, RMS is root mean square 'n' is number of data points.

The theoretical MRT (\bar{t}_T) of the material in a closed system is given as [8]:

$$\bar{t}_T = \frac{M}{Q} \quad (8)$$

Where M : weight of bed material and Q is flow rate at steady state condition of gasifier. The results of \bar{t}_T were calculated based on equation (4) for cold and hot model and are given in Table.6.

Results and Discussions

In experimental test carried out at cold condition, the tracer concentration curves were recorded by detectors mounted at the inlet, bottom and outlet of the gasifier using DAS system. The response of detector D1 shows an instantaneous increase in tracer concentration soon after injection and subsequently become constant. The tracer concentration curve monitored by detector D1 at coal feed inlet is a sharp pulse of narrow width. The tracer curve monitored at distributor outlet by detector D2 and gasifier outlet at top by detector D3 could be considered as an impulse for analysis. The concentration curves monitored by detector D1 at the coal feed inlet and by detector D2 at the distributor outlet were considered for detailed flow analysis. The experimentally measured and normalized RTD curves at the distributor outlet (D2) during the five different runs of both cold and hot conditions are shown in Fig. 7.

The model parameters i.e. tanks number (N) and mean residence time corresponding to best fit i.e. minimum RMS value are tabulated in Table 7. The Impulse responses of RTD functions of Gamma function model for different values of tank number are shown in Fig. 8. The gamma distribution model as described above was adopted to simulate the treated RTDs recorded at the bottom of the gasifier by detector D2. The model predicted RTD corresponding to the minimum RMS and experimentally measured RTD curves are compared as shown in Figs. 9 to 13. The values of theoretical MRTs, experimental

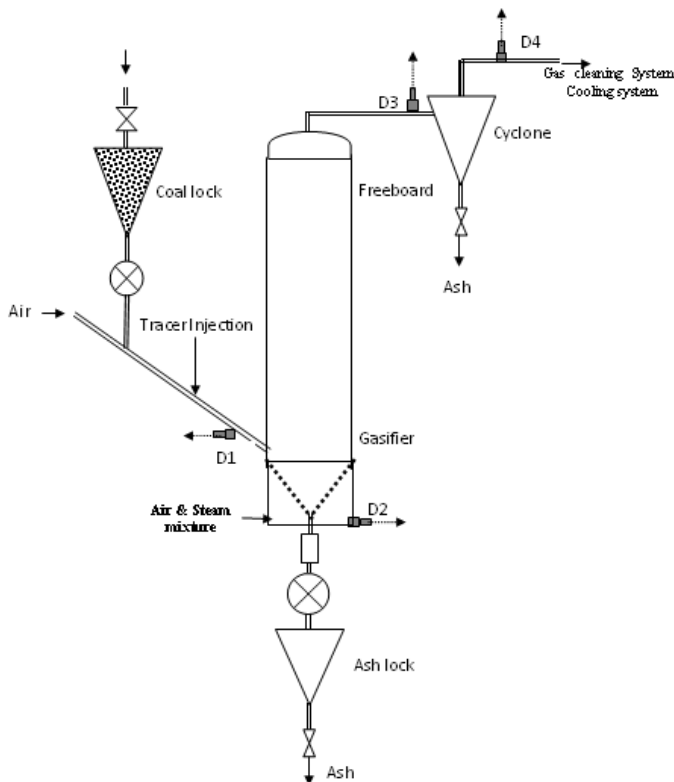


Fig .6. Schematic Diagram of 200 mm Test Rig
RTD curve, (M_1) gives MRT of the processes given by [3]

$$M_1 = \bar{t} = \int_0^{\infty} t E(t) dt \quad (3)$$

The tracer concentration curves measured at the bottom of the gasifier (D2) were treated and analyzed using RTD analysis software developed by IAEA, Austria, [17]. The treated curves for different runs are shown in Fig.7. After applying corrections of background subtraction, tail correction and zero shifting, mean residence time of tracer concentration curves were determined using [4]

$$\tau_{Exp} = \frac{\int_0^t t_i C_i(t_i) dt}{\int_0^t C_i(t_i) dt} \quad (4)$$

The response equation of above can be written in dimensionless form as [5]:

$$E(\theta) = \frac{N^N \theta^{N-1} e^{-N\theta}}{\Gamma(N)} \quad (5)$$

where, θ is dimensionless time (t/τ), $E(\theta)$ is called dimensionless residence time distribution function. The value of

MRTs and model predicted MRTs were found to be in good agreement with each other.

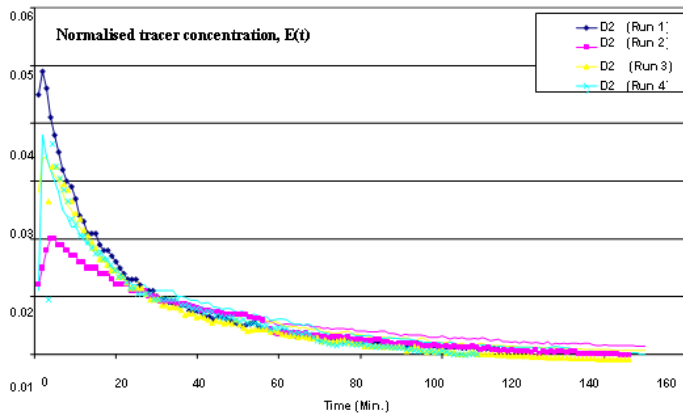


Fig. 7: Experimentally measured Residence Time Distributions

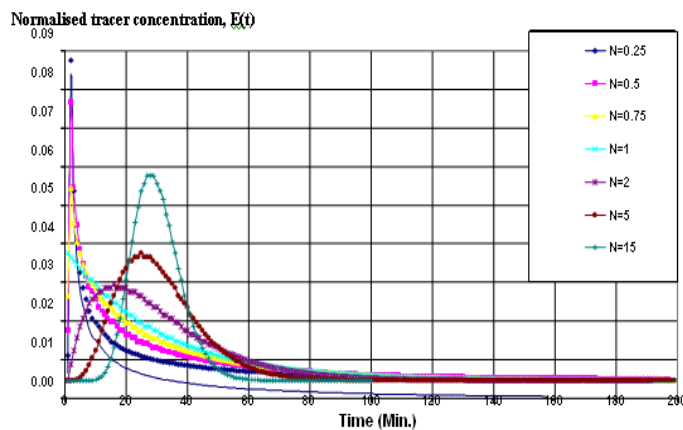


Fig. 8: Impulse responses of gamma function model for different tank number

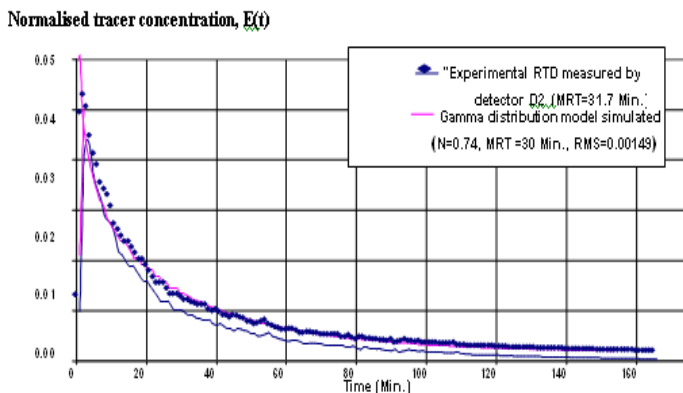


Fig. 9: Experimental and model simulated RTDs (Run 1)

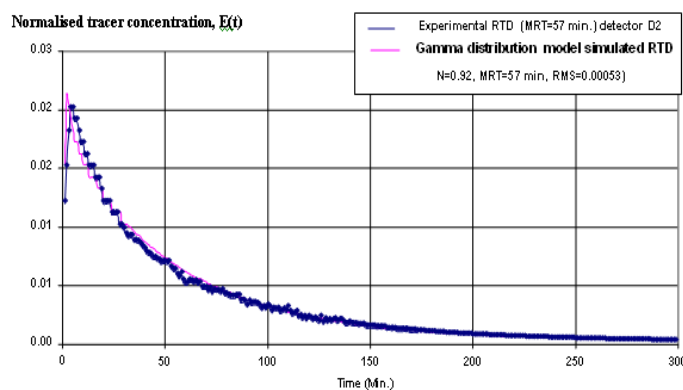


Fig. 10: Experimental and model simulated RTDs (Run 2)

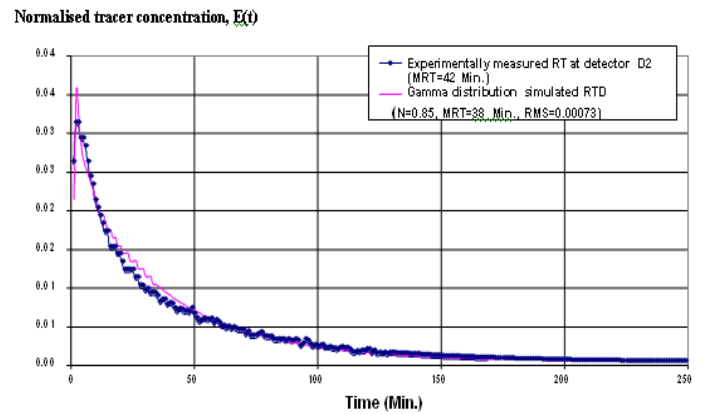


Fig. 11: Experimental and model simulated RTDs (Run 3)

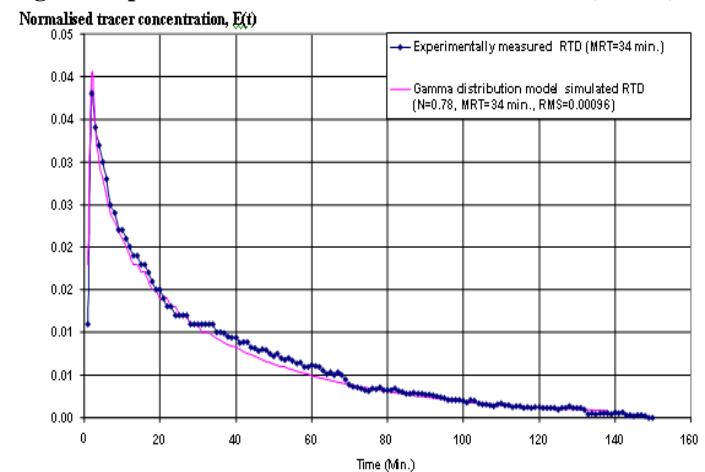


Fig. 12: Experimental and model simulated RTDs (Run 4)

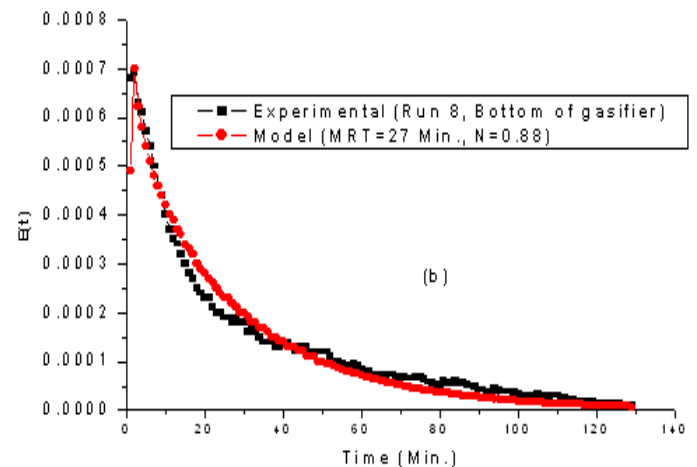


Fig. 13: Experimental and model simulated RTDs (Run 5)

Conclusions

a) For fluidized bed reactors, the bubbles are the prime motive agents for both gas and solids displacement, a detailed verification should involve comparison of indirect measurements of bubble characteristics viz bubble diameter and bubble rise velocity in the demonstration hot bed as well as the scaled down cold model. Experiments using scaled models of bubbling beds have been carried out using the scaling relationships presented. The hydro dynamic parameters viz Froude no, bubble rise velocity and bubble diameter of semi circular dia of 940 mm cold model were measured and presented graphically. The results of 168 tpd PFBG hot bed hydrodynamic behaviour was in close agreement with that of semi circular cold model.

Table 1: Comparison Of Simplified Scaling Parameters Of 168 TPD PFBG and Cold Model

PARAMETERS	168TPD-PFBG HOT MODEL	2/3 rd COLD MODEL
ρ_s / ρ_g	750	750
U_o^2 / gD	0.051	0.051
U_o / U_{mf}	2.96	2.96
ϕ_s	0.63	0.65
D / d_p	1420	1420

Table 2: Comparison Of Hydrodynamic Parameters Of 168 TPD PFBG Hot And Cold Model

PARAMETERS	168TPD-PFBG	D MODEL	2/3 rd COLD MODEL
Temperature (K)	1273	303	303
Pressure (atm)	13	1.0	1.0
Gas Density (kg/m ³), ρ_g	3.26	1.2	1.2
Solid Density (kg/m ³), ρ_s	2248	925	925
Air Flow in distributor (kg/hr)	12733	5745	1025
Minimum Fluidization Velocity (m/s), U_{mf}	0.226	0.284	0.231
Operating Velocity (m/s), U_o	0.84	1.1	0.68
Diameter of Bed (m)	1.4	1.4	0.94
Particle Diameter (mm)	0.99	0.99	0.662
Orifice Diameter (mm)	5.7	5.7	5.7
Number of orifices in distributor, N	370	678	154
Orifice velocity (m/s) U_{or}	38.6	66	54
Pressure Drop in bed (kg/m ²), ΔP_b	2898	1187	795
Pressure Drop in distributor (kg/m ²), ΔP_d	1159	475	318
Static bed height (m), H_{static}	1.75	1.79	1.20
Expanded bed height (m), H_f	2.47	2.51	1.66

Table-3: Comparison of simplified scaling parameters for 200 mm PFBG hot and cold models

Scaling Parameters	Hot Model	Cold Model
ρ_s / ρ_g	1929	2163
U_o^2 / gD	1.564	1.566
U_o / U_{mf}	13.48	13.48
L/D	Geometrically similar	Geometrically similar
ϕ_s	0.82	0.85
PSD	Matched	Matched

Table 4 : Properties of coal

Property	Hot model values
Bulk Density	815 Kg/m ³
Particle density	1680 Kg/m ³
Composition of Coal, (% Wt)	Carbon (C): 38.3, Hydrogen (H ₂): 2.4, Sulphur (S): 0.3, Nitrogen (N ₂): 0.71, Oxygen (O ₂): 9.6.
Type of particle according to Geldart Classification	Group D

Table-5: Comparison of operating parameters of 200 mm PFBG Hot and Cold models

Parameter	Hot Model	Cold Model
Temperature, T (K)	1323	308
Pressure, P (kg/cm ²) (a)	3	1.3
Dynamic viscosity, μ_g (kg/m-sec)	49.18E-06	18.875E-06
Particle density, ρ_s (kg/m ³)	1502.65	2596.2
Gas density, ρ_g (kg/m ³)	0.78	1.2
Sphericity (ϕ_s)	0.82	0.85
Minimum fluidization velocity (U_{mf}), m/sec	0.131	0.1
Superficial velocity (U_o), m/sec	1.752	1.24
Reactor diameter (D), m	0.2	0.2

Table 6: Results of model simulation of Residence time distribution (RTD)

Run No.	Pressure (Atm)	Q _{Feed} (kg/hr)	Q _{Extraction} (kg/hr)	H _{Bed} (m)	W _{Bed} (kg)	Q _{Air} (kg/hr)	Q _{Steam} (kg/hr)	T (°C)	\bar{t}_T (Min)
1	1.1	15.2	13.3	0.17	6.3	115	0	Ambient	30.0
2	1.1	7.6	7.2	0.2	7.6	115	0	Ambient	60.0
3	1.1	10.2	8.7	0.2	7.6	115	0	Ambient	45.0
4	1.2	23.3	4.7	0.1	4.61	55	2	900-1000	30.0
5	1.2	28.96	13.8	0.21	7.6	60	3	900-1000	22.0

Table 7: Results of measured and predicted MRT and RMS using Gamma Distribution Model

Run No.	Q _{Feed} (kg/hr)	\bar{t}_T (Min)	\bar{t}_E (Min)	Gamma distribution model		
				τ_m (Min)	N	RMS
1	15.2	30.0	29.0	30.0	0.74	0.00149
2	7.6	60.0	57.0	57.0	0.92	0.00053
3	10.2	45.0	42.0	48	0.85	0.00073
4	23.3	30.0	34.0	34.0	0.78	0.00096
5	28.96	23.0	21.0	22.0	0.8	0.0008

b) The Pressurised Fluidised Bed Gasifier (PFBG) system is designed to behave as a well mixed flow system for coal and any deviation from the well mixed flow condition will deteriorate the performance and efficiency of the gasification system. Radiotracer technique was employed to measure the RTD of the coal in the gasifier. The results of treated RTD curves generated in a 200mm pressurized fluidised bed gasification test rig both in cold and hot conditions are presented. The model simulated MRT values predicted by gamma distribution model are close to the measured values.

c) The gamma distribution model was found suitable to describe the dynamics of the coal particles in the gasifier. The values of tank number (N) estimated for all the five runs were found to be less than one indicating bypassing of the small fraction of coal particles. This fraction of process material is observed to leave the gasifier immediately after entering into the system resulting in production of un burnt carbon at the outlet of gasifier. This implies that the fraction of the coal which leaves bypassing the system partially gets converted, thus reducing the carbon conversion efficiency of the gasifier. Hence the gasifier does not behave as a well mixed system as desired and the carbon conversion efficiency of process can be improved by repeating the RTD tests with varying gasifier parameters. Hence the RTD technique can be used to minimize the bypassing of the coal particles by conducting / repeating the experimentation with different average coal particle size, transport velocity and changing the location of coal feed above distributor till the targeted carbon conversion efficiency gets maximized thus achieving the gasifier design efficiency.

Acknowledgments:

The authors would like to thank the members of coal research group at BHEL for their suggestions in preparation of the paper.

References

- [1] Jones, T. Glicksman, L.R., "An Experimental Investigation of Gas Flow in a Scale Model of a Fluidized Bed Combustor", Powder Technology, 45, (1986), 201-203.
- [2] M.T. Nicastro and L.R.Glicksman "Experimental Verification of Scaling Relationships For Fluidized Bed". Ch.Eng.Sci., vol.39., 1984.
- [3] Jochim Werther "Scale-Up Modeling for Fluidized Bed Reactors", Ch.Eng.Sci., vol.47. NO.9-11, pp.2457-2462, 1992.

- [4] R.DiFelice, S.Rapagna, P.U.Foscolo and L.G.Gibilaro "Cold Modeling Studies of Fluidized Bed Reactors"- Ch.Eng.Sci. Vol.47, No. 9-11, pp. 2233-2238, 1992.

- [5] L.R. Glicksman "Scaling Relationships For Fluidized Beds", Ch.Eng.Sci., vol. 39, No. 9, pp. 12373-12379, 1984.

- [6] L.R.Glicksman, T.Yule and A.Dymess and R.Carson. "Scaling the Hydrodynamics of Fluidized Bed Combustor With Cold Models: Experimental configuration",

- [7] Peter E.G.Gogolek and John R.Grace, "Fundamental Hydrodynamics Related to Pressurised Fluidized Bed Combustion" Prog.Energy.Combust.Sci., vol.21. pp.419-451, 1995.

- [8] Danckwerts, P.V., "Continuous flow systems, distribution of residence times", Chem. Eng. Sci. 2, pp.1-13 (1953).

- [9] Charlton, J.S. (Eds.), Radioisotope Tracer Techniques for Problem solving in Industrial Plants, Leonard Hill (1986).

- [10] Pant, H.J., Kundu, A. and Nigam, K.D.P., "Radiotracer Applications in Chemical Process Industry", Reviews in Chemical Engineering, 17, pp.165-252 (2001).

- [11] Thyn, J., Zitny, R., Jaroslav, K. and Cechak, T., Analysis and Diagnostics of Industrial Processes by Radiotracers and Radioisotope Sealed Sources, 1 & 2, CVUT, Praha, (2000)

- [12] Pant, H.J., Saroha, A.K. and Nigam, K.D.P., "Measurement of Liquid Holdup and Axial Dispersion in Trickle Bed Reactors Using Radiotracer Technique", Nukleonika 45, pp. 235-241 (2000).

- [13] Pant, H.J., Thyn, J., Zitny, R. and Bhatt, B.C., "Radioisotope Tracer Study in a Sludge Hygienization Research Irradiator (SHRI)", Applied Radiation and Isotopes 54, pp.1-10 (2000).

- [14] Pant, H.J., "Flow Rate Measurement in a Draft Tube Crystallizer By Means of a Neutrally Buoyant Sealed Radioactive Follower Technique", Applied Radiation and Isotopes 53, pp. 999-1004 (2000)

- [15] Pant H.J. et al, "Investigation of flow behaviour of coal particles in a pilot-scale fluidized bed gasifier (FBG) using radiotracer technique", Applied Radiation and Isotopes 67 (9), pp.1609-1615 (2009).

- [16] Geldart, D., "Types of Gas Fluidization", Powder Technology, 7 (5), pp. 285-292 (1973).

- [17]. International Atomic Energy Agency, Residence Time Distribution Software Analysis User's Manual", Computer Manual Series No.11, IAEA, Vienna, Austria, pp-218 (1996).

[18] Nauman, E.B. Buffham, B.A: Mixing in continuous flow system. John Wiley and Sons, NY, 1983.

[19] Buffham, B.A. and Gibilaro, L.G., "A Generalization of the Tanks in Series Mixing Model", *AIChE J.*, 14 (5), pp.805-806 (1968).

[20] Michelsen, M.L., "A Least-Squares Method for Residence Time Distribution Analysis", *Chem. Eng. J.*, 4, pp.171-179 (1972).

Nomenclature

A = Amount of radiotracer (MBq)

A_r = Archimedes number, $d_p^3 \rho_g (\rho_p - \rho_g) g / \mu_g^2$

$C(t)$ = Radiotracer concentration (Counts/unit time)

D = Reactor Diameter

d_p = Material particle size

$E(t)$ = Experimental RTD function

$E_m(t, \tau_m, N)$ = Theoretical RTD function

Fr = Froude number

g = Acceleration due to gravity

H_{Bed} = Bed height (m)

W_{Bed} = Bed weight (kg)

H_s = Static bed height

H_{mf} = Height of bed at minimum fluidization (m)

L = Height of bed

M = Weight of bed material (kg)

N = Number of tanks

ΔP_{mf} = Pressure drop at min. fluidizing velocity

ΔP_d = Pressure Drop in distributor (kg/m^2)

α = distributor apex angle

ρ_p, ρ_s = Density of the material used

ε_{mf} = Voidage of bed at minimum fluidization

ϕ_s = Sphericity of the particles

τ_m : Model predicted mean residence time (Min)

Γ : Gamma function

μ_g = Viscosity of the fluid

ρ_f = Density of the fluid

ΔP_b = Pressure Drop in bed (kg/m^2) ,

Q_{Feed} = Feed flow rate of coal particles (kg/hr)

$Q_{Extraction}$ = Extraction flow rate of coal particles (kg/hr)

Q_{Air} = Flow rate of air (kg/hr)

Q_{Steam} = Flow rate of air steam (kg/hr)

Re_{mf} = Reynolds No at minimum fluidizing velocity

RT = Residence time

t = time variable (second)

\bar{t}_E = Experimental mean residence time (min)

\bar{t}_T = Theoretical mean residence time (min)

T = Temperature ($^{\circ}\text{C}$)

U_o = Superficial Gas velocity (m/sec)

U_{mf} = Minimum Fluidization Velocity (m/sec)

V = Volume of tank (m^3)

V_f = volume of fractional tank (m^3)

X = Fraction of volume (m^3)

Abbreviations

BHEL Bharat Heavy Electricals Ltd

CCT Clean Coal Technology

IGCC Integrated Gasification Combined Cycle

PFBG Pressurized Fluidized Bed Gasification

RTD Residence Time Distribution

CSIR Council of Scientific and Industrial Research

TIFAC Technology Information Forecasting and Assessment Council

CII Confederation of Indian Industry

NTPC National Thermal Power Corporation