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EXPERIMENTAL AND COMPUTATIONAL STUDY OF THE BED DYNAMICS OF SEMI -CYLINDRICAL GAS - SOLID FLUIDIZED BED

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ABSTRACT

With computational fluid dynamics (CFD) it is possible to get a detailed view of the flow behavior of the fluidized beds. A profound and fundamental understanding of bed dynamics such as bed pressure drop, bed expansion ratio, bed fluctuation ratio and minimum fluidization velocity of homogeneous binary mixtures has been made in a semi-cylindrical fluidized column for gas solid systems, resulting in a predictive model for fluidized beds. In the present work attempt has been made to study the effect of different system parameters (viz. size and density of the bed materials and initial static bed height) on the bed dynamics. The correlations for the bed expansion and bed fluctuations have been developed on the basis of dimensional analysis using these system parameters. Computational study has also been carried out using a commercial CFD package Fluent (Fluent Inc.). A multifluid Eulerian model incorporating the kinetic theory for solid particles was applied in order to simulate the gas–solid flow. CFD simulated bed pressure drop has been compared with the experimental bed pressure drops under different conditions for which the results show good agreements.

KEY WORDS

Computational Fluid Dynamics, multi-fluid Eulerian model, semi-cylindrical bed, gas-solid fluidization and bed dynamics

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INTRODUCTION

A very large part of the processes in the chemical processes industries concern gas solid flows. In order to obtain well operating gas solid contactors a through understanding of flow phenomena of gas solid flow is vital. A good qualitative understanding and an accurate quantitative description of fluid flow is necessary for the modeling of these devices. Accurate modeling of these fluidized beds is complicated. Despite their widespread application, much of the development and design of fluidized bed reactors has been complex empirical as the flow behavior of gas–solid flow in these systems makes flow modeling a challenging task.

The fundamental problem encountered in modeling hydrodynamics of a gas–solid fluidized bed is the motion of two phases where the interface is unknown and transient, and the interaction is understood only for a limited range of conditions (Gilbertson and Yates, 1996). With the advent of increased computational capabilities computational fluid dynamics, CFD is emerging as a very promising new tool in modeling hydrodynamics. While it is now a standard tool for single-phase flows, it is at the development stage for multiphase systems, such as fluidized beds. Work is required to make CFD suitable for fluidized bed reactor modeling and scale-up. CFD also holds great promise for multiphase flows, obtaining accurate solutions is much more challenging, not just because each of the phases must be treated separately, but, in addition, a number of new and difficult factors (such as drag, lift and inter particle phases etc.) come into play.

LITERATURE

A discrete particle model of a gas-fluidised bed has been developed by Hoomans et al. (1996). The two-dimensional motion of the individual, spherical particles was directly calculated from the forces acting on them, accounting for the interaction between the particles and the interstitial gas phase. Gera et al. (1998) have compared Lagrangian model and multifluid or Eulerian-Eulerian model in applying CFD modeling to gas-solid fluidized beds. They have also modeled the particle collisions by the hard sphere approach and soft sphere approach was used by Kobayashi et al. (2000). Kaneko et al. (1999) have examined the multi body collisions by the Discrete Element Method (DEM). Though the models based on a DEM allow the effects of various particle properties on the motion of fluid to be studied, it is computationally intensive. Due to these computational limitations, the Eulerian-Lagrangian model is normally limited to a relatively small numbers of particles. Therefore, the multifluid model is the preferred choice for simulating macroscopic hydrodynamics in Eulerian-Eulerian continuum modeling with the fluid and solid phases treated as interpenetrating continuum phases and it is the most commonly used approach for fluidized bed simulations (Pain et al., 2001). They have used certain averaging techniques and assumptions as required to obtain a momentum balance for the solids phase.

The drag force on a single sphere in a fluid has been well studied and empirically correlated (Clift et al., 1978 and Bird et al., 2002) for a wide range of particle Reynolds numbers. However, when a single particle moves in a dispersed two-phase mixture, the drag is affected by the presence of other particles. In the present study, calculation of the momentum exchange coefficient for the gas–solid systems has been done using Syamlal and O' Brien (1989) drag function. The solid-phase momentum equation contains an additional term to account for momentum exchange due to particle collisions. The absence of the stress term of the particle phase in the particulate momentum equation has led to different models adopting different closure methods, including the kinetic theory model (Sinclair and Jackson, 1989; Gidaspow, 1994; Hrenya and Sinclair, 1997). In granular flow, particle velocity fluctuations about the mean are assumed to result in collisions between particles being swept along together by the mean flow.

Numerous studies have shown the capability of the kinetic theory approach for modeling bubbling fluidized beds (e.g. Pain et al., 2001; Sinclair and Jackson, 1989; Hrenya and Sinclair, 1997; Ding and Gidaspow, 1990; Gelderbloom et al., 2003).

The kinetic theory model clearly explains why fluidized beds are such good heat transfer devices. Their thermal conductivity, k can be expressed in terms of mean free path, l , the average of fluctuating velocity, v which is essentially the square root of the granular temperature and the density and the specific heat of particles as follows.

$$k = l\, v\, \mathrm{r} Cp$$

where r=density and *Cp* =heat capacity of particles.

The mean free path, *l* is essentially the particle diameter divided by the solids volume fraction. For dense systems it is of the order of particle diameter. For dilute conditions it will approach the size of equipment. The kinetic theory model has the potential to explain why fluidized beds are such good contacting devices.

Jenkins and Savage (1983) utilized the coefficient of restitution (between 1 and 0, for fully elastic collisions and for fully inelastic collisions respectively) to account for the loss of energy due to collision of particles, which is not considered in the classical kinetic theory. The energy dissipated as a result of collisions of granular inelastic particles has been used by Lun et al. (1984) to obtain the ratio of the velocity fluctuations to the mean flow as a function of the coefficient of restitution. A decrease in the coefficient of restitution results in less elastic collisions generating more fluctuating kinetic energy (Goldschmidt et al., 2001). In dense two-phase flows, the particle interaction time may be much larger than the particle mean free flight time. Thus, the assumption that a pair of particles completes its interaction before interacting with another particle may be invalid as the solids concentration increases (Zhang and Rauenzahn, 2000).

Despite the modeling challenges, application of CFD to model fluidized bed hydrodynamics continues to develop, as it has many advantages including design, optimization and scale-up of such systems. Some of the correlations used in the models, however, remain to be empirical or semi empirical. As a result, the model and its parameters must be validated against experimental measurements obtained at similar scale and configurations. Some of the challenges with respect to CFD model validation for gas–solid systems have been reviewed by Grace and Taghipour (2004).

Computational Fluid Dynamics (CFD) modeling

CFD applications to a number of unit operations and processes in the chemical process industries, oil and gas industry are increased in drastic rate in the past decade [H.S. Pordal et al, 2001, T.J. Fry et al 2001 and C. J. Matice et al 1997, 2001].

The governing equations for the present study can be summarized as follows:

Mass conservation equations of gas (g) *and solid* (s) *phases*:

$$\frac{\partial}{\partial t}.(\alpha_g.\rho_g) + \nabla.(\alpha_g.\rho_g.\vec{v}_g) = 0 \qquad \text{-------------} \qquad (1)$$

$$\frac{\partial}{\partial t}.(\alpha_s.\rho_s) + \nabla.(\alpha_s.\rho_s.\vec{v}_s) = 0 \qquad \text{--------------} \qquad (2)$$

Momentum conservation equations of gas and solid phases are as follows:

$$\frac{\partial(\alpha_g\rho_g\vec{v}_g)}{\partial t} + \nabla.(\alpha_g\rho_g\vec{v}_g{}^2) = -\alpha_g\vec{\nabla}P + \nabla.(\alpha_g\bar{\bar{\tau}}_g) + (\alpha_g\rho_g\vec{g}) - K_{gs}(\vec{v}_g - \vec{v}_s) \qquad \text{-------} \quad (3\text{-a})$$

$$\frac{\partial(\alpha_s\rho_s\vec{v}_s)}{\partial t} + \nabla.(\alpha_s\rho_s\vec{v}_s{}^2) = -\alpha_s\vec{\nabla}P + \nabla.(\alpha_s\bar{\bar{\tau}}_s) + (\alpha_s\rho_s\vec{g}) - K_{gs}(\vec{v}_g - \vec{v}_s) + G(\alpha_g)\vec{\nabla}\alpha_s \qquad \text{------} \quad (3\text{-b})$$

Fluctuation energy conservation of solid particles:

$$\frac{3}{2}\cdot\left[\frac{\partial}{\partial t}\cdot(\rho_s\cdot\alpha_s\cdot\Theta_s) + \nabla\cdot(\rho_s\cdot\alpha_s\cdot\vec{v}_s\cdot\Theta_s)\right] = \left(-p_s\bar{\bar{I}} + \bar{\bar{\tau}}_s\right) : \nabla\cdot\vec{v}_s + \nabla\cdot(k_{\Theta s}\cdot\nabla\cdot\Theta_s) - \gamma_{\Theta s}. \qquad \text{----} \qquad (3\text{-c})$$

Momentum Exchange Co-efficient by Syamlal–O'Brien drags function:

$$K_{gs} = \frac{3}{4}\frac{\alpha_s\alpha_g\rho_g}{v^2_{r,s}d_s}C_D\left(\frac{\text{Re}_s}{v_{r,s}}\right)\left|\vec{v}_s - \vec{v}_g\right| \qquad \text{---------} \qquad (4)$$

Where

$$C_D = \left(0.63 + \frac{4.8}{\sqrt{\text{Re}_s/v_{r,s}}}\right) \qquad \text{----------} \qquad (5)$$

And

$$v_{r,s} = 0.5\left(A - 0.06\,\text{Re}_s + \sqrt{(0.06\cdot\text{Re}_s)^2 + 0.12\,\text{Re}_s(2B - A) + A^2}\right) \qquad \text{----} \; (6)$$

With

$$A = \alpha_g^{\ 4.14}, \ \ B = 0.8\alpha_g^{\ 1.28} \quad \text{For } \alpha_g \le 0.85$$

or

$$A = \alpha_g^{\ 4.14}, \ \ B = 0.8\alpha_g^{\ 2.65} \quad \text{for } \alpha_g > 0.85$$

In this study the Syamlal and O' Brien (1989)'s drag function been applied to determine the momentum exchange coefficients. These correlations are provided by the above equations. The coefficient of restitution quantifies the elasticity of particle collisions between 1 and 0 for fully elastic and fully inelastic collisions respectively. It was utilized by Jenkins and Savage (1983) to account for the loss of energy due to collision of particles, which is not considered in the classical kinetic theory. The energy dissipated as a result of collisions of granular inelastic particles has been calculated to obtain the ratio of the velocity fluctuations to the mean flow as a function of the coefficient of restitution (Lun et al., 1984).

Solids stress accounts for the interaction within solid phase. It is derived from the granular kinetic theory. It is represented in the following form.

$$\bar{\tau}_s = -P_s\bar{I} + 2\alpha_s\mu_s\bar{S} + \alpha_s\left(\lambda_s - \frac{2}{3}\mu_s\right)\nabla.\vec{u}_s\bar{I} \qquad \text{--------} \qquad (7)$$

where,

$\bar{S} = \frac{1}{2}\left(\nabla\vec{u}_s + (\nabla\vec{u}_s)^T\right)$	Strain rate
P_s	Solids pressure
g_o	Radial distribution function
λ_s, μ_s	Solids bulk and shear viscosity
α_s	Volume fraction of solids

In the present study, both the experimental and simulation investigations are applied in a 2D fluidized bed column to enable meaningful comparison of the hydrodynamic results. CFD simulation is based on commercial CFD software, Fluent applying Syamlal and O' Brien (1989) drag function to calculate the momentum exchange coefficients.

EXPERIMENTAL SECTION

A schematic diagram of the experimental set up is shown in the Fig.-1. The experiment has been conducted in a Semi-cylindrical Perspex column of 16cm internal diameter, 90cm in height and 1cm thick. The pressure drop across the fluidized bed is measured by the manometer using carbon tetrachloride as the manometric liquid. A calming section is provided below the distributor plate for the uniform distribution of the air. The experiment has been carried out to study the effect of different system parameters on the bed hydrodynamics. Scope of the experiment has been presented in Table-1. The bed materials were fluidized by controlling the air flow rate through the rotameter. The bed dynamics viz. bed expansion ratio, bed fluctuation ratio and bed pressure drops were measured at different superficial velocities (calculated through the flow rate of air) for different operating conditions i.e. by varying different system parameters.

The simulation of fluidized bed was performed by solving the governing equations of mass, momentum and energy conservation [Anderson, j. D., 1995] using Fluent 6.1.22, CFD software (Fluent, 2003). A multifluid Eulerian model, which considers the conservation of mass and momentum for the gas and fluid phases, was applied. The kinetic theory of granular flow which considers the conservation of solid fluctuation energy was used for closure of the solids stress terms. The restitution coefficient is considered as 0.9 here from the literature. Considered simulation parameters for glass beads are shown in Table-2.

RESULTS AND DISCUSSION

The governing equations were solved using the finite volume approach. The 2D computational domain was discretized by 13200 quadrilateral cells. A time step of 0.001s with 20 iterations per time step was chosen. This iteration was adequate to achieve convergence for the majority of time steps. First-order discretization schemes for the convection terms are used. The relative error between two successive iterations was specified by using a convergence criterion of 10-3 for each scaled residual component. Sensitivity analysis was performed through a case study for investigating the effect of time step, discretization schemes and convergence criterion on the final modeling results. The phase-coupled SIMPLE (PC-SIMPLE) algorithm (Vasquez and Ivanov, 2000), which is an extension of the SIMPLE algorithm to multiphase flows, is applied for the pressure-velocity coupling. In this algorithm, the coupling terms are treated implicitly and form part of the solution matrix. The pressure-velocity coupling is based on total volume continuity and the effects of the interfacial coupling terms are fully incorporated into the pressure correction equation.

Singh, R. (1997) has explained the advantages of non-cylindrical conduits by carrying out thorough experimentation on hydrodynamic studies of the fluidized beds.

The correlations have been developed on the basis of dimensional analysis for the bed expansion ratio and fluctuation ratio [Kunii and Levenspiel,1991] by correlating different system parameters with the experimentally observed expansion and fluctuation ratios respectively. These correlation plots have been shown in Fig-2 (A) and (B).

Bed expansion ratio, R:

$$R = 1.164\left[\left(\frac{H_S}{D_C}\right)^{-0.0623}\left(\frac{dp}{D_C}\right)^{-0.325}\left(\frac{\rho_s}{\rho_f}\right)^{-0.161}\left(\frac{U}{U_{mf}}\right)^{0.856}\right] \quad ------- \quad (8)$$

Bed fluctuation ratio, r:

$$r = 7.233\left[\left(\frac{H_S}{D_C}\right)^{0.168}\left(\frac{dp}{D_C}\right)^{-0.0517}\left(\frac{\rho_s}{\rho_f}\right)^{-0.268}\left(\frac{U}{U_{mf}}\right)^{0.579}\right] \quad --------- \quad (9)$$

Experimental analyses were performed to identify the steady state pressure drop, ΔP, bed expansion ratio, $Hmax/H_S$ and bed fluctuation ratio, $Havg/H_S$ at different superficial velocities. The voidage, ε, i.e., the gas phase volume fraction in a fluidized bed, can be determined using pressure drop across the bed (Y. Tsuji et al, 1993). The overall pressure drop, the bed expansion ratio, bed fluctuation ratio and the voidage were 1955 Pa, 1.18, 1.6 and 0.6, respectively at minimum fluidization velocity (U_{mf}=1.382 m/s, Hs = 14 cms) for glass beads. These values were compared with those predicted by CFD simulation under different superficial gas velocities. The CFD simulations were performed using the transient Eulerian-Granular model in Fluent software. Various superficial gas velocities, some multiples of Umf, were examined in order to cover a range of velocities under and above Umf. Fig-3 shows a contour plot of solids fraction of a typical result using the

Syamlal-O'Brien drag model. At time 0.0 s, the bed was fluidized at U=1.382 m/s, a superficial gas velocity of ~1.5 Umf. Initially, the bed height increased with bubble formation until it settled to a steady-state bed height. In Fig-3, a comparison of the time-averaged bed pressure drop, using Syamlal-O'Brien drag function against superficial gas velocity is plotted. Good agreement has been observed between the simulation and experimental results at velocities higher than U_{mf}.

The model predictions, however, do not show close agreement with experiments at velocities under U_{mf}. This may be attributed to the solids not being fluidized, thus being dominated by inter particle frictional forces, which could not be predicted by the approach taken by the multi-fluid model for simulating gas-solid phases. The correlations developed by the dimensional analysis for expansion and fluctuation ratios for the fluidized beds are in well agreement with experimental results. Standard and mean deviations between the experimental (R_{exp} / r_{exp}) and calculated (R_{cal} / r_{cal}) values have been found to be 11.5 & 0.06% and 13.67 & 0.96% for bed expansion and fluctuation ratios respectively. A sample plot of the simulated solids volume fractions of the 2D bed has been shown in Fig.-3 for glass beads of size 4.3mm with a static bed height of 14cms at different times. Fig. - 4 (A) and (B) show the comparison of the experimental expansion ratio with CFD simulated and correlated expansion ratio respectively. Pressure drop of the bed has been studied by considering different superficial velocities, both above and under the minimum fluidization velocity. Fig- 5(A) to (D) show the comparisons between the experimental and simulated pressure drops plotted against superficial velocities for different initial static bed heights (viz.14, 18, 22 and 26 cms) for a particular size (4.3mm) of glass beads . Fig- 6 (A), (B), and (C) show the comparisons between the experimental and simulated pressure drops for the glass beads of a given static bed height (14cms) for different particle sizes. Fig- 7 (A), (B) and (C) show the comparisons between the experimental and simulated pressure drops for materials of different densities.

From the developed correlations, based on dimensional analysis it is found that the bed expansion ratio increases with increase in velocity ratio (U/U_{mf}), but decreases with the increase in density ratio(ρ_s/ρ_f), size ratio (dp/Dc) and bed aspect ratios. Whereas the bed fluctuation ratio was found to increase with the increase in velocity ratio and bed aspect ratio and the same was observed to decrease with the increase in density ratio(ρ_s/ρ_f) and size ratio (dp/Dc). Therefore it is better to maintain lower values of velocity ratio, density ratio and size ratio for optimum bed dynamics.

CONCLUSIONS

Experimental work has been carried out to study the hydrodynamics of a semi-cylindrical gas-solid fluidized bed. The effects of various system parameters (viz., static bed height, flow rate of fluid, particle size, and the density of the particles) on the bed dynamics (such as minimum fluidization velocity, bed pressure drop, bed expansion ratio and the bed fluctuation ratio) were studied. Experimental results were correlated by dimensional analysis approach and compared by the computational approach.

It has been observed that the minimum fluidization velocity increases with increase in particle size and particle density. It was also observed that it is not a function of the static bed height. The calculated values of the bed expansion/fluctuation ratios obtained from the developed correlations show good agreement with the experimental values over wide ranges of parameters.

Flow behavior is studied computationally using a commercial CFD package FLUENT 6.1. In the experimental setup a gas is fed through a small diameter inlet to the bottom of the calming section, where as in the CFD simulation a flat velocity profile is defined at the bottom of the calming section. But it has been proved by some authors that the error in considering this assumption is negligible (F. Taghipour et al., 2005). A multifluid Eulerian model integrating the kinetic theory for solid particles using Fluent CFD software was capable of predicting the gas–solid behavior of a fluidized bed.

Comparison of the model predictions, using the Syamlal–O'Brien drag function and experimental measurements on the time-average bed pressure drop, velocity profile, bed expansion, and qualitative gas–solid flow pattern indicated reasonable agreement for most operating conditions thereby ensuring the wide scope of the semi-cylindrical fluidized bed in various industries over wide range of parameters.

NOMENCLATURE

C_D	:	Drag coefficient, dimensionless
D_C	:	Column diameter of fluidized bed, m
dp	:	Diameter of particle, m
e	:	Restitution coefficient, dimensionless
g	:	Aacceleration due to gravity, m/s^2
g_0	:	radial distribution coefficient, dimensionless
Hs	:	Static bed height, m
Hmax	:	Maximum (expanded) bed height, m
Havg	:	Avarge expanded bed height, m
I	:	Stress tensor, dimensionless
$K_{\Theta S}$	:	Diffusion coefficient for granular energy, Dimensionless
Kgs	:	Gas/solid momentum exchange coefficient, Dimensionless
P	:	Pressure, Pa
r	:	Bed fluctuation ratio ($Havg/H_S$), Dimensionless
R	:	Bed expansion ratio ($Hmax/H_S$) , Dimensionless
Re	:	Reynolds number, Dimensionless
t	:	time, s
U	:	Superficial velocity of fluid, m/s
U_{mf}	:	Minimum fluidization velocity, m/s

Greek Letters:

α	:	volume fraction
η	:	dynamic viscosity, Pa s
$\vec{v}$	:	velocity, m/s
$\vec{\nabla}P$	:	pressure drop, Pa
Θ	:	granular temperature, m^2/s^2
$\gamma_{\Theta s}$	:	collision dissipation of energy, $kg/s^3\ m$
γ	:	rate of angular energy dissipation
λ	:	bulk viscosity ,Pa.s
μ	:	shear viscosity, Pa s
β	:	interphase momentum transfer coefficient,
$\overline{\overline{I}}$	:	stress tensor, Dimensionless

$\bar{\bar{\tau}}$: stress tensor, Pa

Subscripts:

g : *for gas*
s : *f or solid*
avg. : *for average*
max. : *for maximum*
exp : *for experimental values*
sim : *for CFD simulated values*

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Figure Caption:

Fig-1 : Experimental set-up

Fig-2 (A) : Correlation plot for the bed expansion ratio against the system parameters

Fig-2 (B) : Correlation plot for the bed fluctuation ratio against the system parameters

Fig-3 : Simulated solids volume fraction profile of 2D bed (U = 2.073m/s, i.e. ~1.5Umf, drag function: Syamlal–O'Brien, e_{ss} = 0.9).

Fig-4 : Comparison plot for the experimental, correlated and CFD simulated expansion ratios

Fig-5(A) : Comparison plot for the experimental and CFD simulated pressure drop for the system of glass beads (ρ_s = 2360kg/m^3), Hs=14cms, dp = 4.3mm

Fig-5(B) : Comparison plot for the experimental and CFD simulated pressure drop for the system of glass beads(ρ_s = 2360kg/m^3), Hs=18cms, dp = 4.3mm

Fig-5(C) : Comparison plot for the experimental and CFD simulated pressure drop for the system of glass beads(ρ_s = 2360kg/m^3), Hs=22cms, dp = 4.3mm

Fig-5(D) : Comparison plot for the experimental and CFD simulated pressure drop for the system of glass beads(ρ_s = 2360kg/m^3), Hs =26cms, dp = 4.3mm

Fig-6(A) : Comparison plot for the experimental and CFD simulated pressure drop for the system of glass beads, Hs =14cms, dp = 2.58mm

Fig-6(B) : Comparison plot for the experimental and CFD simulated pressure drop for the system of glass beads, Hs =14cms, dp= 2.18mm

Fig- 6(C): Comparison plot for the experimental and CFD simulated pressure drop for the system of glass beads, Hs=14cms, dp=1.29mm

Fig-7(A) : Comparison plot for the experimental and CFD simulated pressure drop for the system of dolomite (ρ_s = 2720kg/m^3), Hs=14cms, dp = 4.3mm

Fig-7(B) : Comparison plot for the experimental and CFD simulated pressure drop for the system of steel balls (ρ_s = 6720kg/m^3), Hs =14cms, dp = 4.3mm

Fig-7(C) : Comparison plot for the experimental and CFD simulated pressure drop for the system of sago(ρ_s = 1309kg/m^3), Hs=14cms, dp= 4.3mm

Table Caption:

Table-1: Scope of the experiment

Table-2: CFD Simulation parameters

Table-1: Scope of the experiment

Sl. No.	Bed Material	Particle size, $d_p x10^3$, m	Particles density, ρ_p, kg/m^3	Static bed height, $H_s x10^2$, m
1	Glass beads	4.3	2360	14
2	Glass beads	4.3	2360	18
3	Glass beads	4.3	2360	22
4	Glass beads	4.3	2360	26
5	Glass beads	2.58	2360	14
6	Glass beads	2.18	2360	14
7	Glass beads	1.29	2360	14
8	dolomite	4.3	2720	14
9	Steel balls	4.3	6720	14
10	Sago	4.3	1309	14

Table-2: CFD Simulation parameters

Description	Value	Comment
Particle density, ρ_p, kg/m^3	2360	Glass beads
Gas density, ρ_g, kg/m^3	1.225	Air
Mean particle diameter, d_p, mm	4.3, 2.58, 2.18 and1.29	Uniform distribution
Restitution coefficient, e	0.9	From literature
Initial solids packing, ε_{s0}	0.60	Fixed value
Superficial gas velocity, U, m/s	0.691,1.382,2.073,2.764 and 3.455	~0.5, ~1.0, ~1.5, ~2.0 and ~2.5 U_{mf}
Bed width, m	0.16	Fixed value
Bed height, m	0.92	Fixed value
Static bed height, m	0.14	Fixed value
Inlet Boundary Condition	VELOCITY	Superficial gas velocity
Outlet Boundary Condition	PRESSURE	Fully developed flow

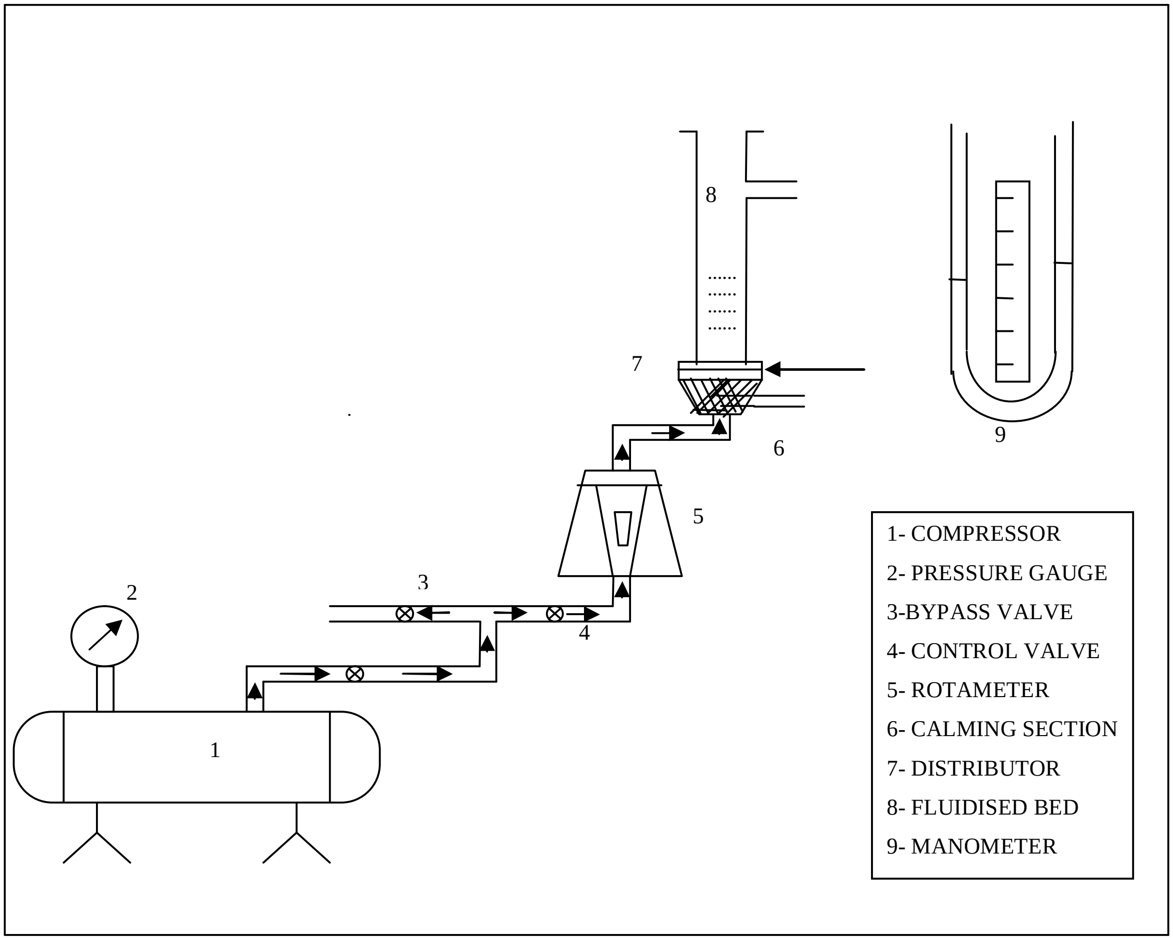

1
2
3
4
5
6
7
8
9
1- COMPRESSOR
2- PRESSURE GAUGE
3-BYPASS VALVE
4- CONTROL VALVE
5- ROTAMETER
6- CALMING SECTION
7- DISTRIBUTOR
8- FLUIDISED BED
9- MANOMETER

Fig.-1

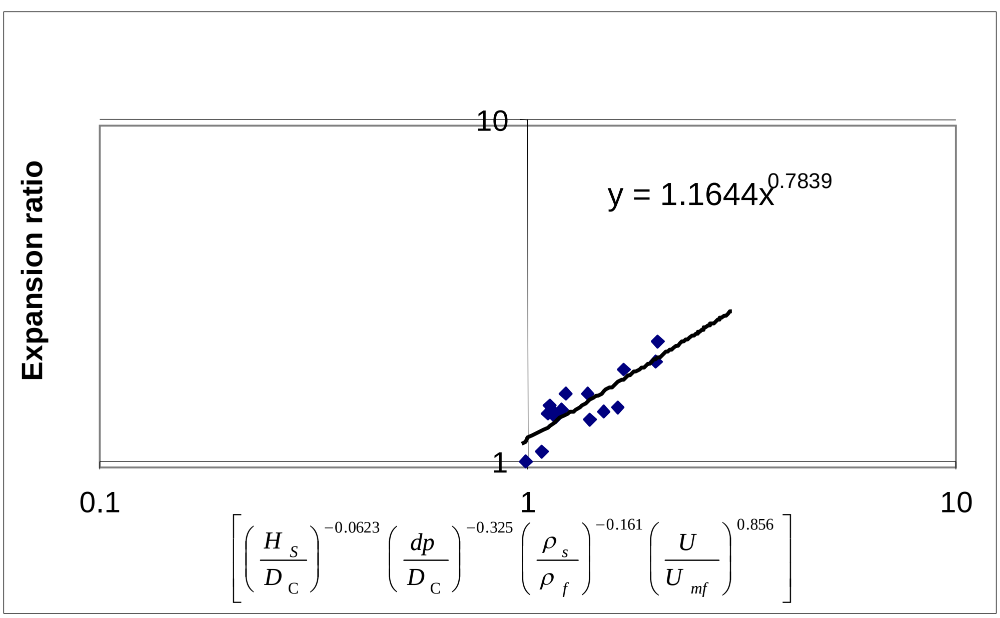
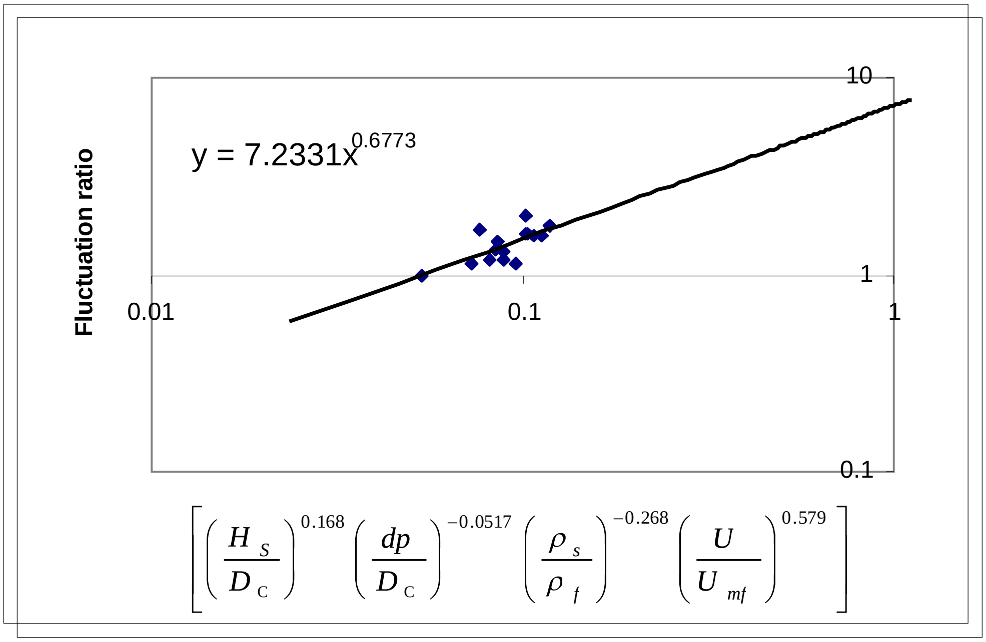
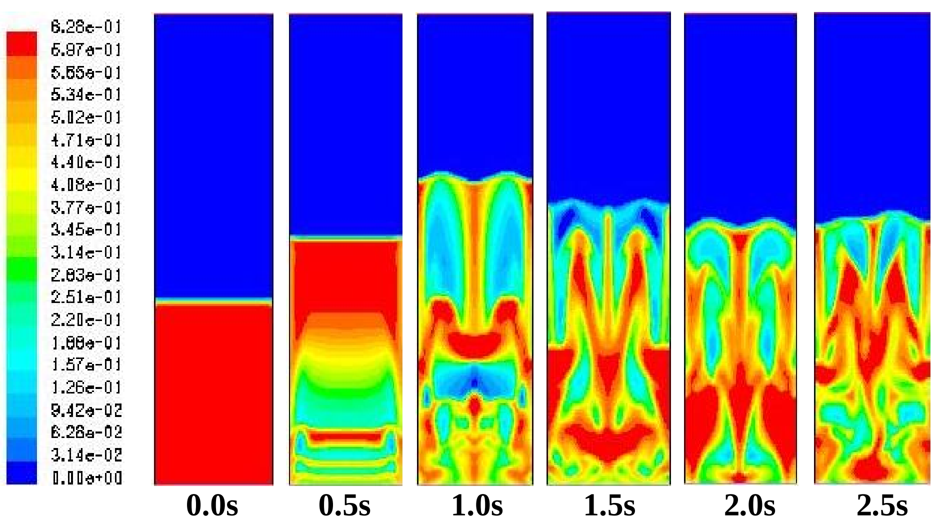

Expansion ratio
10
1
y = 1.1644x^0.7839
0.1
1
10
$\left[\left(\frac{H_S}{D_C}\right)^{-0.0623}\left(\frac{dp}{D_C}\right)^{-0.325}\left(\frac{\rho_s}{\rho_f}\right)^{-0.161}\left(\frac{U}{U_{mf}}\right)^{0.856}\right]$

Fig.-2 (A)



y = 7.2331x^0.6773
Fluctuation ratio
10
1
0.1
0.01
0.1
1
$$\left[\left(\frac{H_S}{D_C}\right)^{0.168}\left(\frac{dp}{D_C}\right)^{-0.0517}\left(\frac{\rho_s}{\rho_f}\right)^{-0.268}\left(\frac{U}{U_{mf}}\right)^{0.579}\right]$$

Fig.-2 (B)



6.28e-01
5.97e-01
5.65e-01
5.34e-01
5.02e-01
4.71e-01
4.40e-01
4.08e-01
3.77e-01
3.45e-01
3.14e-01
2.83e-01
2.51e-01
2.20e-01
1.88e-01
1.57e-01
1.26e-01
9.42e-02
6.28e-02
3.14e-02
0.00e+00
0.0s
0.5s
1.0s
1.5s
2.0s
2.5s

Fig. -3

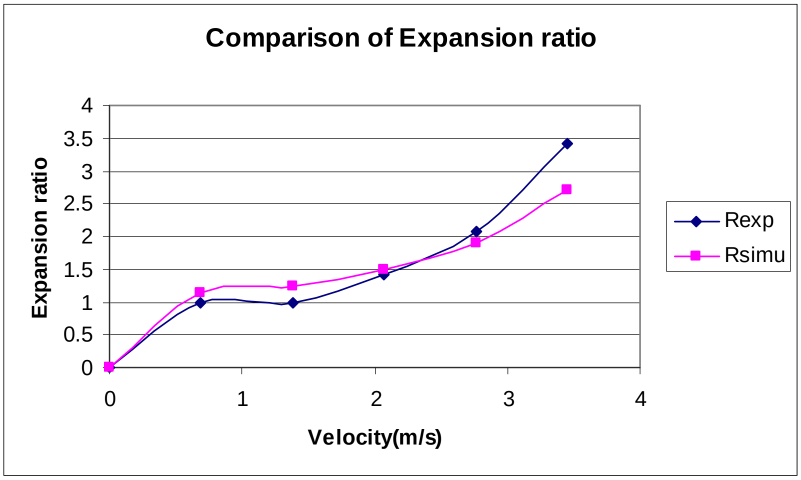


Fig.- 4(A)

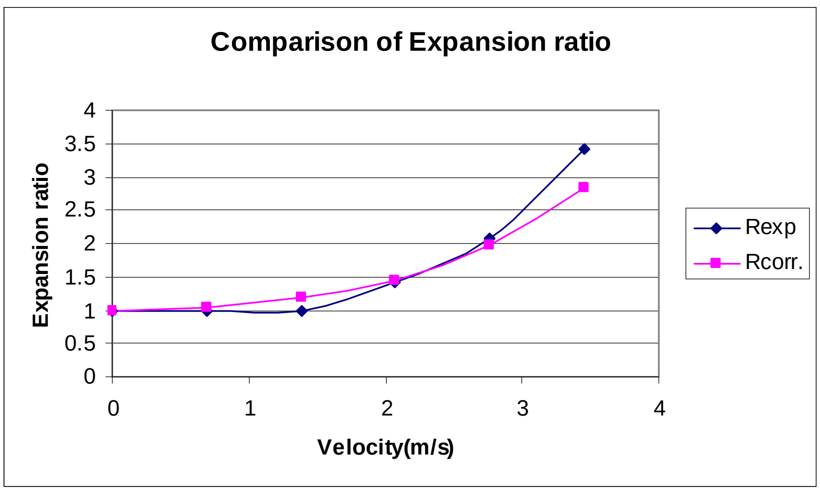


Fig.-4(B)

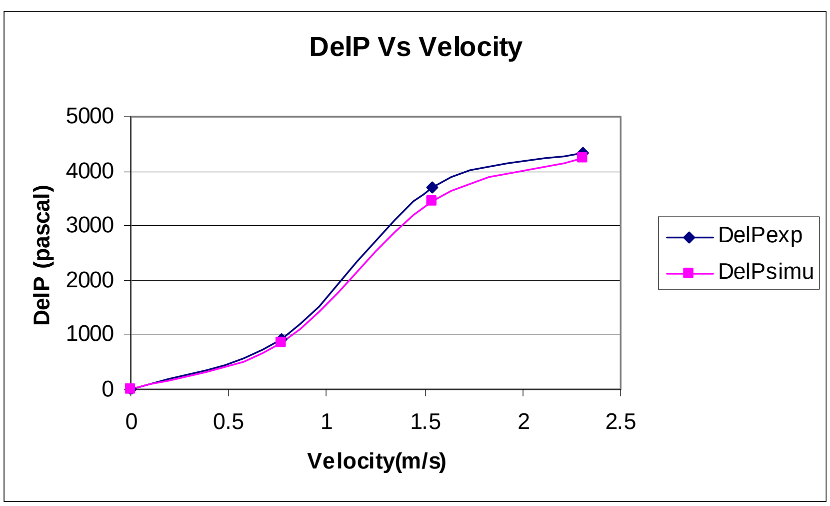


Fig.-5(A)

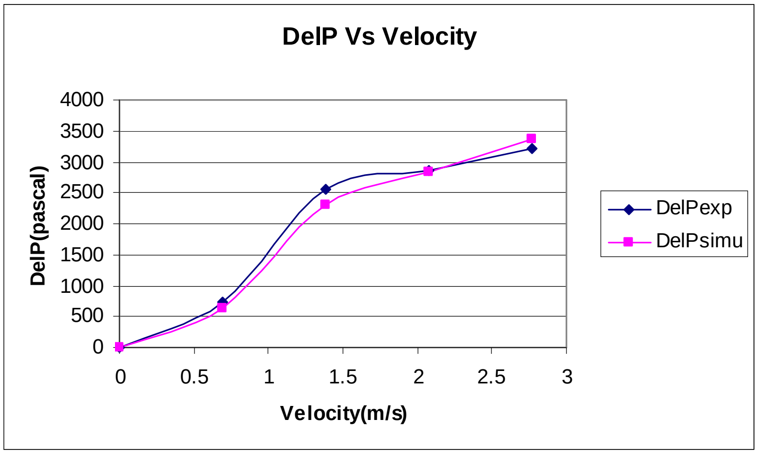


Fig.-5(B)

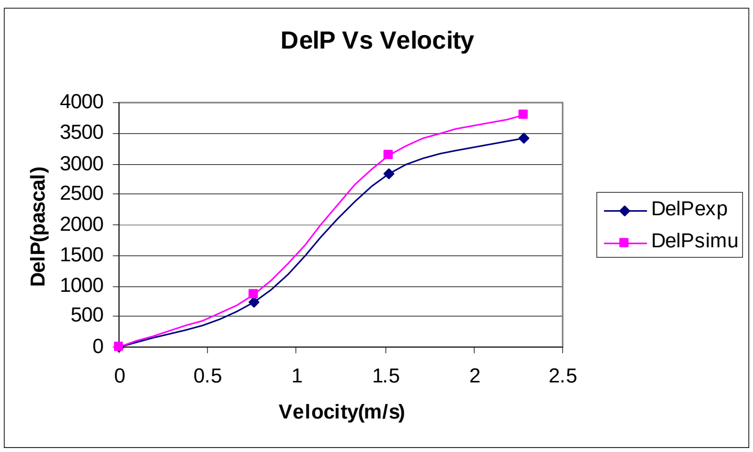


Fig.-5(C)

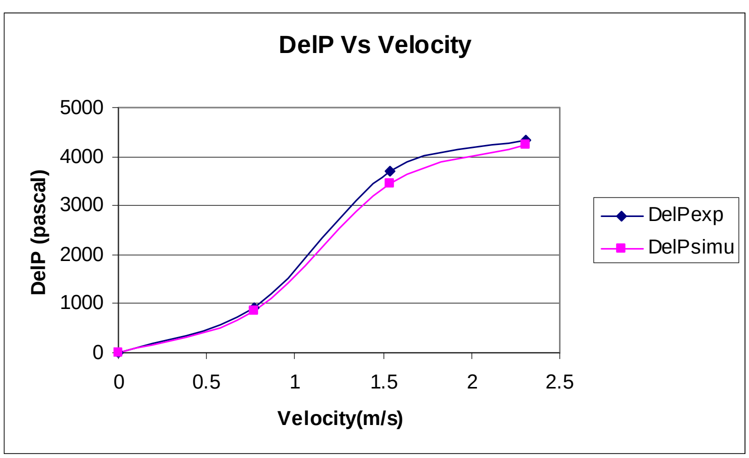


Fig.-5 (D)

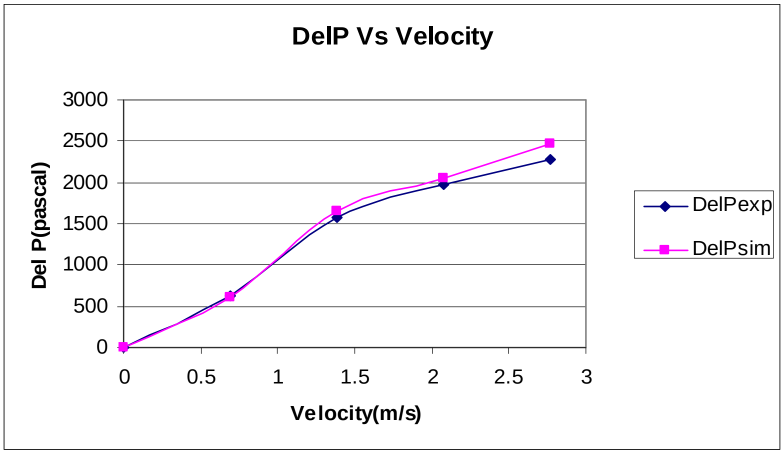


Fig.- 6(A)

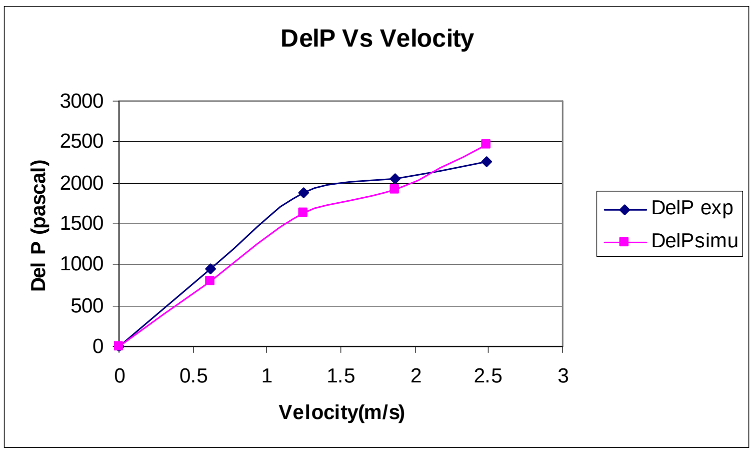


Fig.- 6(B)

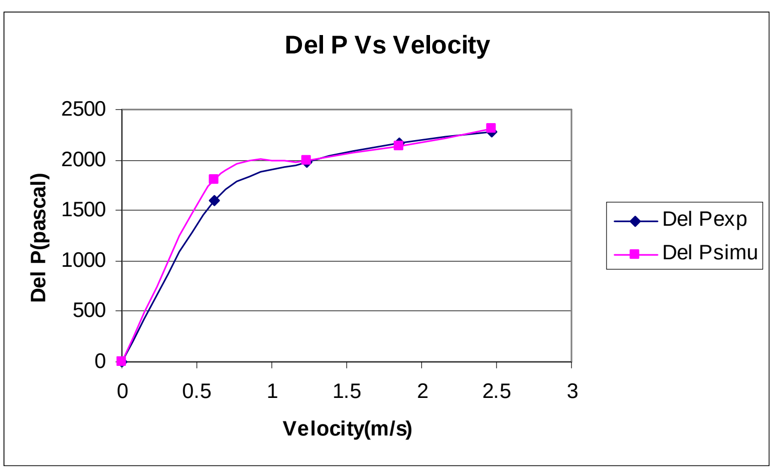


Fig.- 6(C)

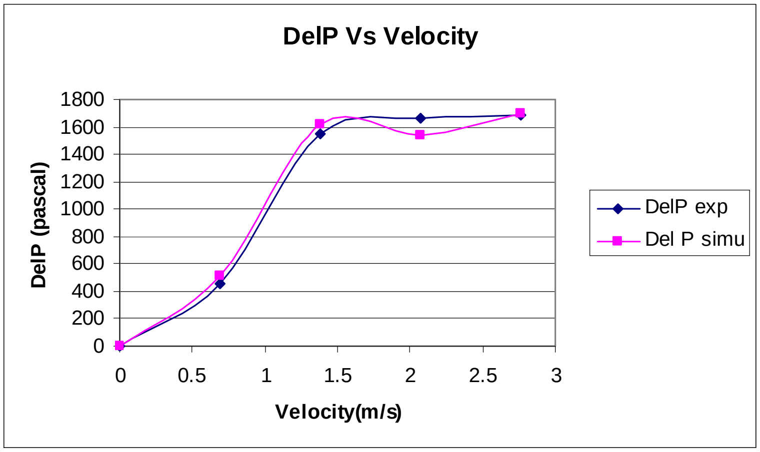


Fig.-7(A)

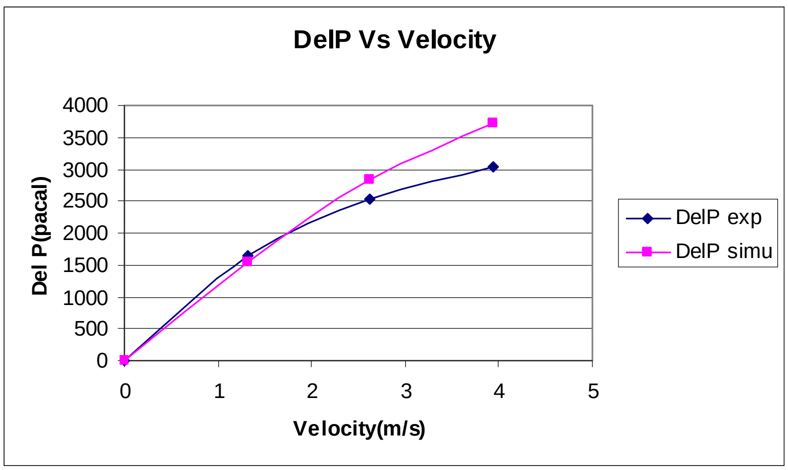


Fig.- 7(B)

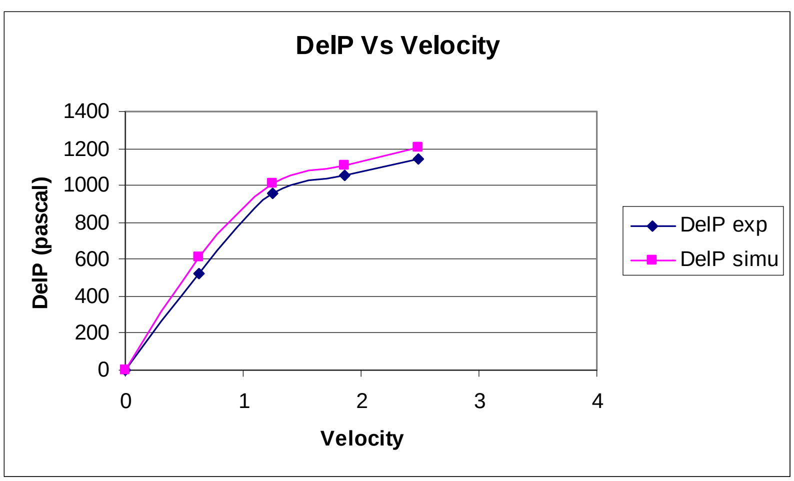


Fig. - 7(C)