Experimental Study of the Influence of Baffles on Hydrodynamics of Gas- Solids Fluidized Bed System

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ABSTRACT

The objective of the present study was to determine the effects of the internals baffles on the gas-solids fluidized bed hydrodynamics, using a circular fluidized bed of Geldert group B sand particles.

Bed expansion data were obtained for un baffled and two different types of baffled (rectangular and circular blades) gas-solids fluidized beds at varying operating conditions, namely air velocity ,particle size (755,424 , 205µm) and initial static bed heights.

Experimental work was carried out in (0.1m) diameter and (1m) height circular fluidized bed column.

The results of the study showed that the particle size affects the measured hydrodynamic behavior especially in the large particle size.

The insertion of baffles into a fluidized bed system improves the contacting efficiency of gas and particle phases.

The effect of rectangular baffles is more significant than the effect of circular baffles on the pressure drop in fluidizing bed.

Keywords: Fluidized Bed, Baffles on Hydrodynamics

دراسة عمليه لتأثير الحواجز على هيدروداينايميكية أعمدة التسييل غاز – صلب

الخلاصة

ان الموضوع الرئيسئ لهذة الدراسة هو لايجاد تاثير الاجزاء الداخلية (الحواجز) المستخدمة في اعمدة التسييل غاز- صلب على الخواص الهيدروديناميكيه لهذه الاعمدة باستخدام عمود زجاجي دائري وحشوة من الرمل حسب(تصنيفGeldert نوع B).

تم الحصول على معلومات تمدد الحشوة في حالة عدم وجود الحواجز ووجود الحواجز حيث استخدم خلال هذه الدراسة نوعين من الحواجز (الشكل الرباعي , الشكل الدائري) بالاعتماد على الظروف التشغيلية التالية :-

سرعة الهواء ,حجم الدقائق (٢٠٥ و٤٢٤ و٧٧٥) مايكروميتر, ارتفاع الحشوة (٠،١) م وارتفاع العمود المستخدم كان بقطر (١)م.

نتائج البحث بينت ان حجم الدقائق توثر على السلوك الهيدروداناميكي للعملية وخصوصا في الدقائق ذات الحجم الاكبر .

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ادخال الحواجز في اعمدة التسييل يحسن كفاءة التماس بين الغاز ودقائق المواد الصلبة المكونة للحشوة . لوحظ من خلال الدراسة بان تاثير الحاجز ذو الشكل الرباعي اكبر من تاثير الحاجز ذو الشكل الدائري على فرق الضغط في أعمدة التسييل .

INTRODUCTION

Gas- Solids Fluidized Beds are used in many industrial applications to drive chemical or combustion /gasification processes. Improvement of the quality of gas- solids fluidized beds can be well achieved by introduction of a suitable baffle to the conventional bed.

Many investigations [1, 4] have already carried out studies concerning the effects of turbulence promoters (baffles) of different shape, size and configuration on various aspects of bed dynamics for gas - solid fluidized bed. Dijk et al. [5] used x-ray technique to observe the effect of a baffle on bubbles. The bubbles where seen to lose (part of) their wake at the baffle. The effect of the baffles was thus to decrease the circulation rate. The influence of internal baffles on the behavior of fluidized beds was reviewed by Harrison and Grace [6] Sitani and Witehead [7] for reactor design purpose. In uniform solids it was observed that internal baffles restrict large-scale solids movement.

The particle dynamics in a gas- solid contactor with inclined baffle plates forming a zig-zag path were numerically simulated by. Nagata et al. [8] using a two fluid model, for the hydrodynamics of gas and solids in the moving fluidized bed.

Hydrodynamics of gas - solids fluidized bed were investigated extensively, however, bubble properties such as bubble diameter, play an important role in the hydrodynamics of gas–solids flow. In the literature, various correlations [9, 11] were proposed to calculate the bubble properties such as bubble diameter and bubble velocity.

Hilal [12] studied the effect of distributor type and baffles insert in fluidized bed column on the expansion of particles. Experimental data were correlated. The effect of vertical internal baffles on the particle mixing and grain drying characteristics in fluidized bed column was investigated by Lim et al. [13].

In this study an attempt was made to study the effect of two different types of baffles on the hydrodynamics of gas - solids fluidized bed.

EXPERIMENTS AND PROCEDURES

The fluidized bed column was constructed of (0.1 m) diameter glass column with a height of (1m). The distributor was a perforated plate made of aluminum sheet of 0.003m thickness with about 1.2% fraction free area [14]. Air from a laboratory compressor was used as the fluidizing gas. Fluidizing air was regulated with two rotameters before the perforated plate distributor. Three types of sand particle diameter were used in the fluidizing bed, 205,424 and 775 µm (using the geometric mean particle diameter, narrow cut solids) at group B of Geldart classification. The physical properties are listed in Table (1).

Two quantities (8.5 kg and 9.5kg) of sand particles were charged into the bed to give a static bed height of about (30 cm) and (40 cm) respectively. Air was introduced from the air bottom to the column flows upward. The bed was first fluidized and allowed to settle. Later for various flow rate of air, the height of fluidized bed and pressure drop were measured across the bed by means of a manometer. The flow rate of air was measured using a rotameter.

The un baffled condition was also investigated, where before the baffle modules was introduced into the bed and the system was allowed to reach a steady state. The schematic diagram of experimental apparatus employed in this study is illustrated in Figure (1).

RESULTS AND DISCUSSION

The minimum fluidizing velocities of the three types particle bed (205,424,775µm) were measured experimentally and calculated theoretically [15] and there average values are listed in Table (2) (i.e. average values of u_{mf} for different experiments, un baffled, rectangular baffle and circular baffle).

Plots of the pressure drop and superficial air velocity on a log- log scale are shown in Figures (2-7).The apparent terminal velocity u_t was calculated from two different correlations (Eq. (1) and (2)), Schiller and Naumann [16] and Khan and Richardson [17] and the average numerical values are presented in Table (2).

$$Ar = 18Re_t\,(1+0.15Re_t^{0.687}) \qquad \ldots (1)$$

$$Ar = (2.07Re_t^{0.27} + 0.33Re_t^{0.64})^{3.45} \qquad \ldots (2)$$

Bed expansion is an important macroscopic characteristic of fluidized beds, which can be easily measured experimentally; this property is usually expressed in terms of the bed expansion ratio(R).The mass velocity ratio (Gr.) has been plotted against experimental values of the reciprocal of bed expansion ratio (1/R) as shown in Figures (8-13) for different particles diameter and static head fluidized beds, when these figures are examined, it can be seen that there is a constant expansion ratio at low velocities for different particle diameters and for un baffled and baffled conditions, but there is a clear difference at high velocities. Fluidized beds undergo a number of state transitions as the superficial gas flow rate increases. Thus a bubbly bed exists at gas velocities somewhat in excess of the minimum fluidization flow rate but at higher, bubbles become unstable and the bed passes into the turbulent region.

Pressure drop in terms of Euler Number under fluidization conditions has been plotted against the fluidizing velocity ratio (u/u_{mf}) as shown in Figures (14-19). For particle diameters 424 and 775 µm, it is observed that the bed pressure drop (Eu No.) is higher for baffled conditions in comparison with un baffled. This may be attributed to the breakup of channels and slugs. The wakes of rising bubbles lose their wake particles due to the presence of baffles. While for particle diameter 205 there is no clear effect of baffles on the values of Eu No. with fluidizing velocity ratio. The use of two types of baffles is quite effective in damping the bed fluctuation and thereby reduces the expanded bed heights when compared with an un baffled gas- solid fluidized bed with identical operating parameters. This behavior was observed by Kumar and Roy (4).The fluidization index (F.I.) represents the ratio of pressure drop to the weight flux in fluidized bed ($\frac{\Delta w}{\frac{w}{Ac}}$).

In this work there are two values of weight flux $(\frac{w}{Ac})$ in case of Hs=0.3 m and Hs=0.4 m. Plotting the fluidization index with fluidizing velocity ratio (u/u_{mf}) as shown in Figures (20-25) show increasing the value of F.1. with increasing fluidizing velocity and constant values can be observed at high velocity ratio because the pressure drop reaches to constant value at high fluidizing velocity. Hofmann (18) has shown that the fluidization index (F.I.) there by the pressure drop is more for the baffled bed than the bed without baffles. The same thing has also been observed here for particle diameter (424 µm) and (775 µm) but for particle diameter (205 µm) this is not significant.

CONCLUSIONS

It has been found that the effect of baffles is small for particle diameter (205 µm) whereas there is a marked decrease in bed expansion ratio for other particles diameters (424 µm and 775µm). The system parameters like, bed height, particle size and fluidizing velocity as well as the presence of baffles significantly influence the Euler Number .The effect of rectangular baffle is clear more than the effect of circular baffles. It could be concluded that the insertion of baffles into a fluidized bed system improves the contacting efficiency of gas and particle phases through breaking up large bubbles into smaller ones and the resulting more circulation of fluidizing gas and solid. This in turn enhances the heat and mass transfer rates of the system. The present results clearly showed that the particle size affects the measured hydrodynamic behavior especially for the large size.

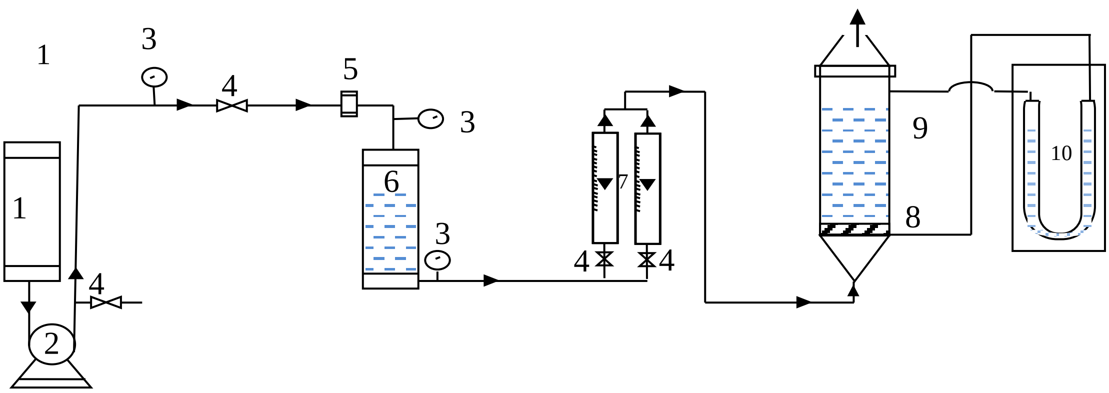


Figure (1) Schematic diagram of experimental apparatus.

1-Filter. 2-Compressor. 3-Pressure gauge. 4-Air directing valve.
5-Pressure regulator. 6-Silica gel dryer. 7-Flowing measuring.
8-Perforated plat distributer 9-Glass- Fluidized bed column.
10-Manometer.-

Table (1) Sand particles distribution and density.

Distribution µm	Particle diameter µm	ρ_p kg/m^3
140-300	205	2388
300-600	424	2430
600-1000	775	2499

Table (2) Measured and calculated minimum fluidizing velocity and terminal velocity.

Particle Diameter d_p µm	u_{mf} m/s		u_t m/s
	Calculated	Experimental	
205	0.033	0.037	1.4320
424	0.1378	0.1450	3.1907
775	0.3875	0.4126	5.6036

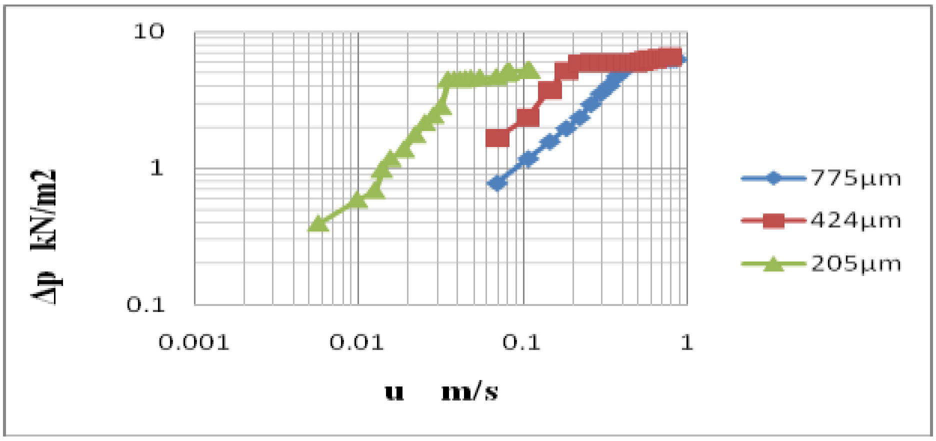


Figure (2) the relationship between the bed pressure drop and the superficial air velocity for different particles diameter, without baffles column, Hs=0.3m.

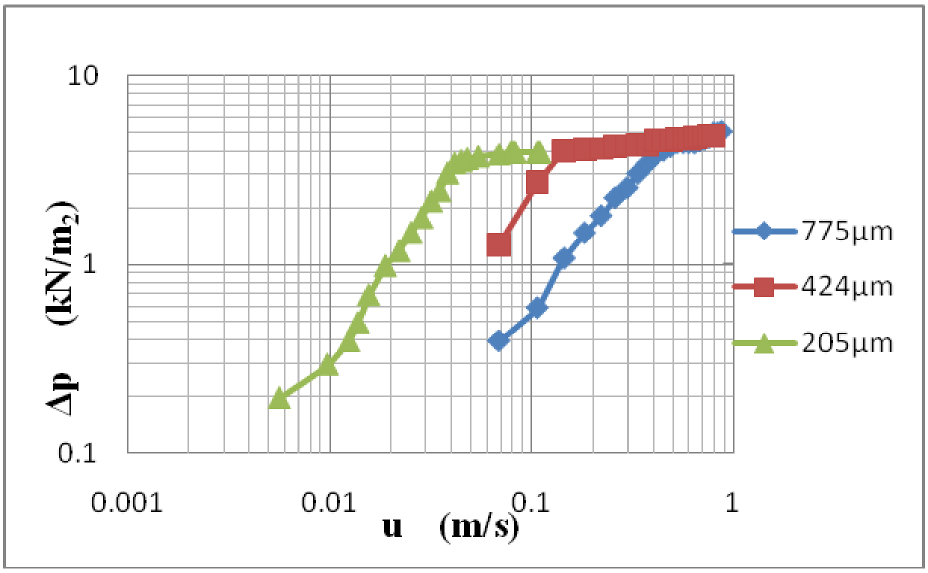


Figure (3) The relationship between the bed pressure drop and the superficial air velocity for different particles diameter, without baffles column, Hs=0.4m.

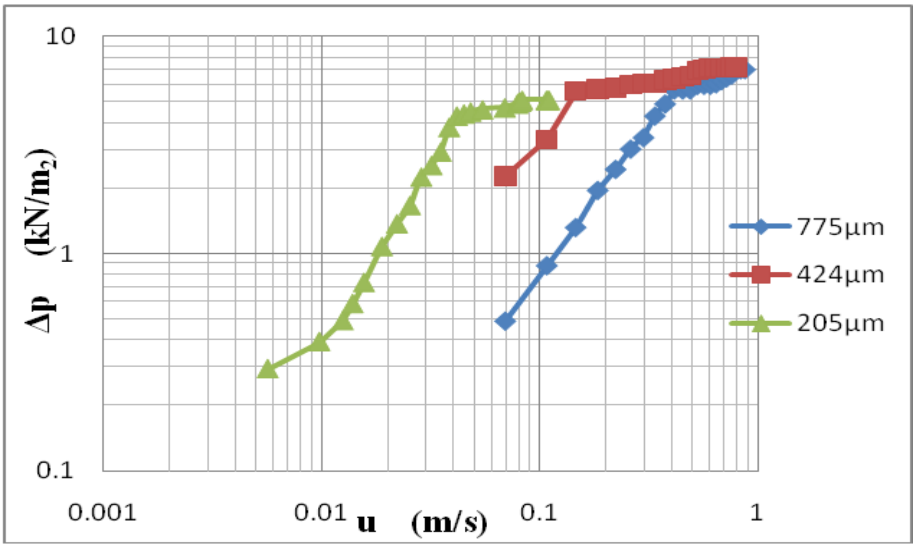


Figure (4) The relationship between the bed Pressure drop and the superficial air velocity for different particles diameter, with rectangular baffles column, Hs=0.3m.

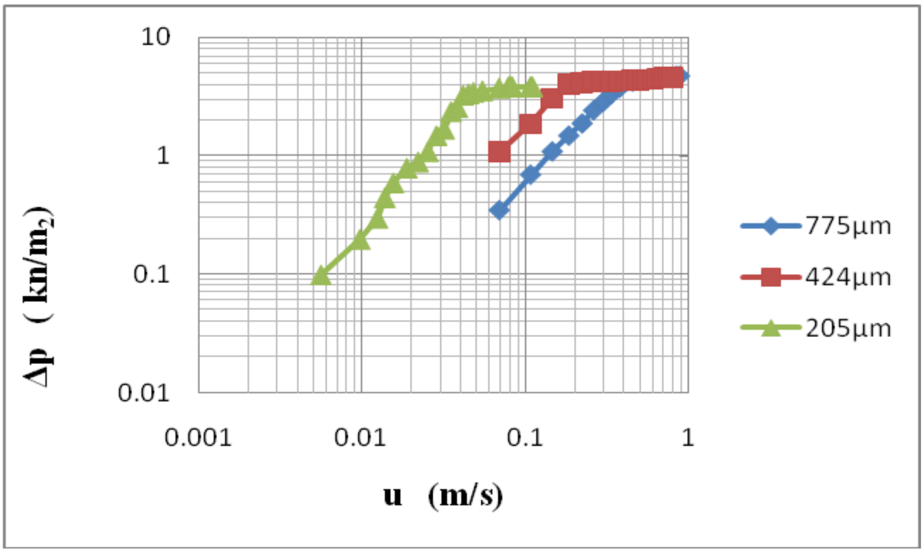


Figure (5) The relationship between the bed pressure drop and the superficial air velocity for different particles diameter, with rectangular baffles column, Hs=0.4m.

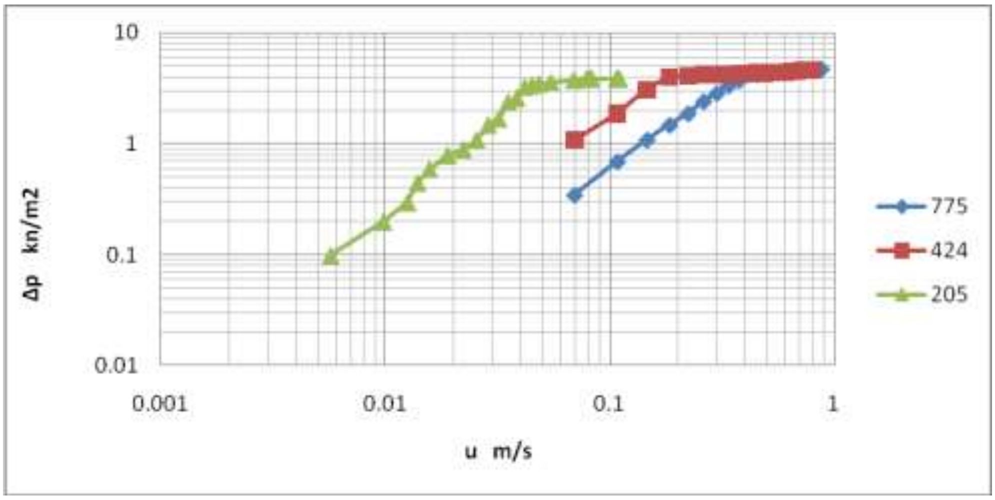


Figure (6) The relationship between the bed pressure drop and the superficial air velocity for different particles diameter, with circular baffles column, Hs=0.3m.

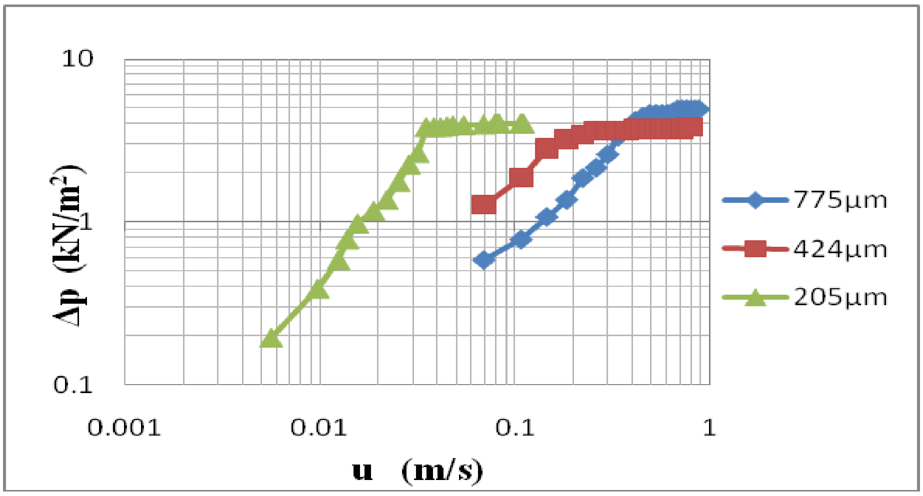


Figure (7) The relationship between the bed pressure drop and the superficial air velocity for different particles diameter, with circular baffles column, Hs=0.4m.

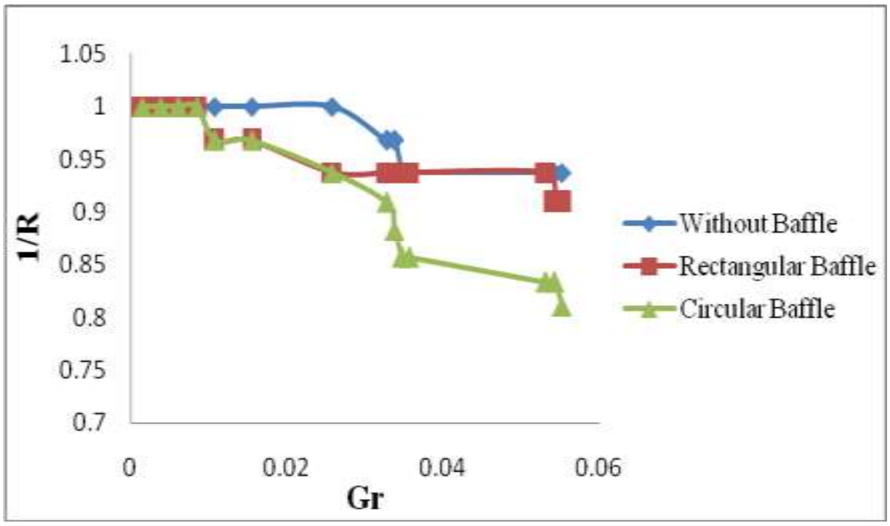
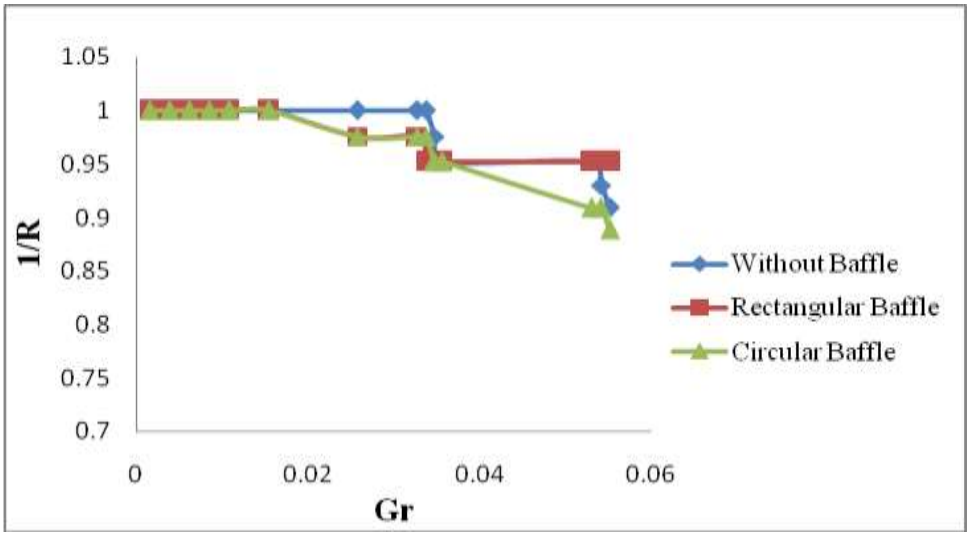


Figure (8) Variation of the reciprocal of bed expansion ratio with mass velocity ratio for baffled and unbaffled beds, dp=205μm, Hs=0.3m.



Figure(9) Variation of the reciprocal of bed expansion ratio with mass velocity ratio for baffled and unbaffled beds, dp=205μm, Hs=0.4m.

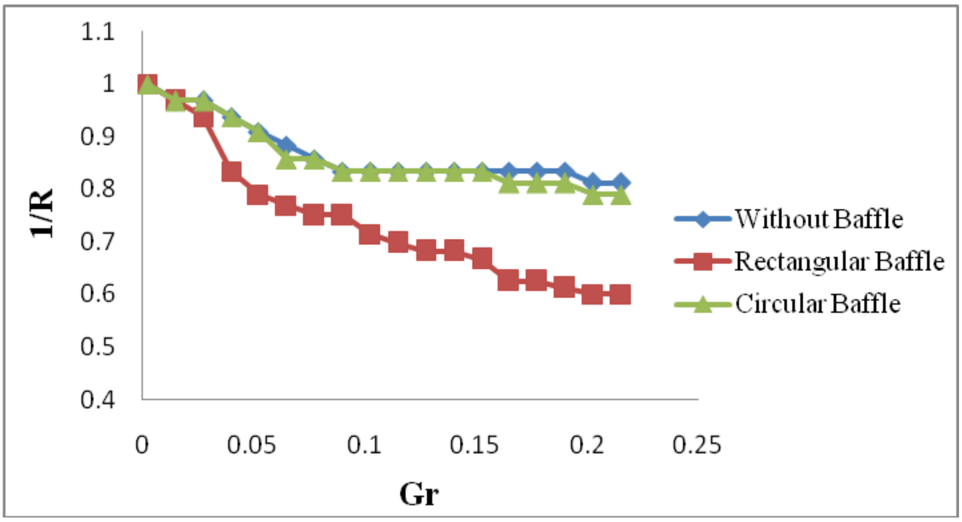


Figure (10) The variation of the reciprocal of bed expansion ratio with mass velocity ratio for baffled and unbaffled beds, dp=424μm, Hs=0.3m.

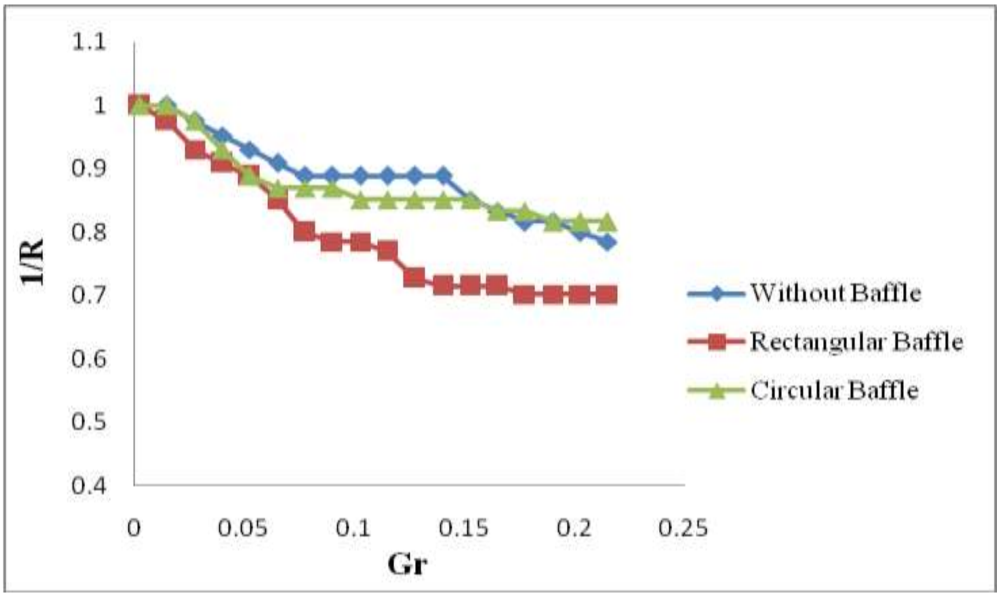


Figure (11) The variation of the reciprocal of bed expansion ratio with mass velocity ratio for baffled and unbaffled beds, dp=424μm, Hs=0.4m.

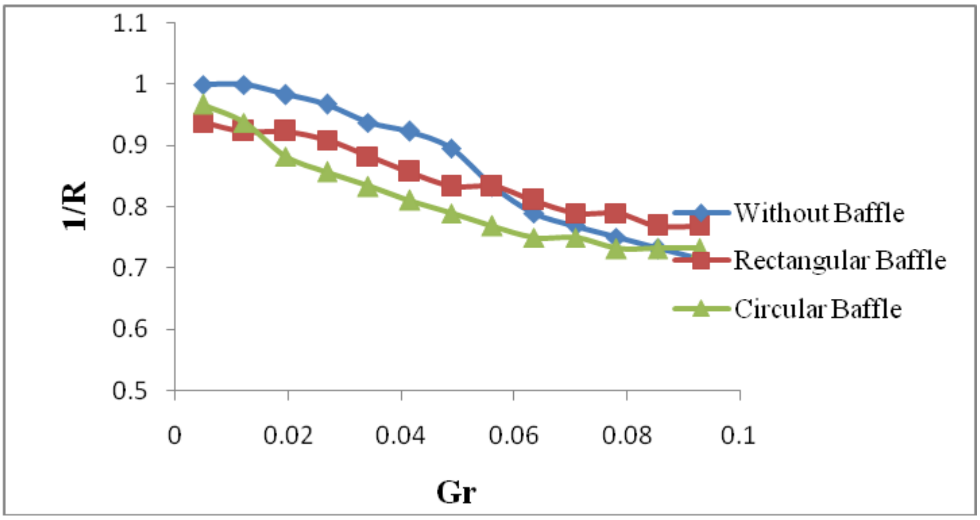


Figure (12) The variation of the reciprocal of bed expansion ratio with mass elocity ratio for baffled and unbaffled beds, dp=775μm, Hs=0.3m.

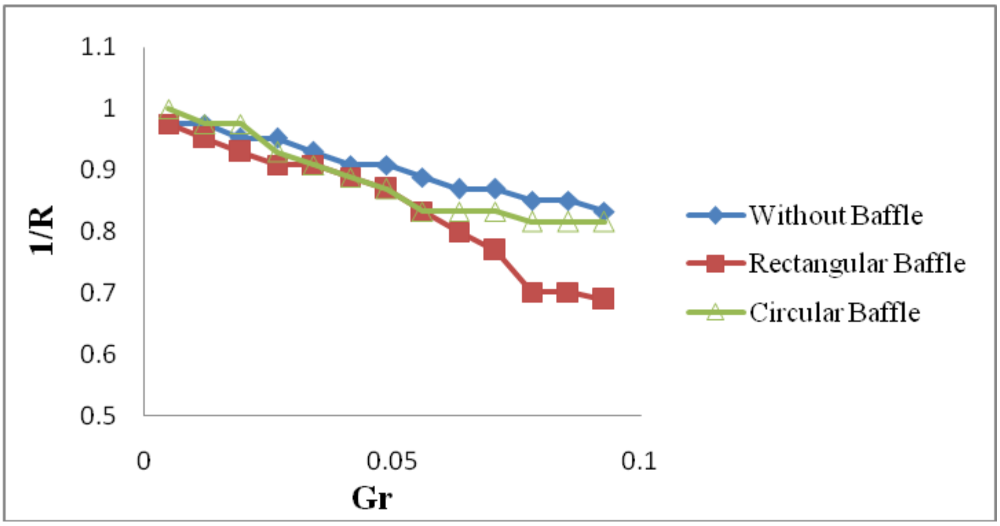


Figure (13) The variation of the reciprocal of bed expansion ratio with mass velocity ratio for baffled and unbaffled beds, dp=775μm, Hs=0.4m.

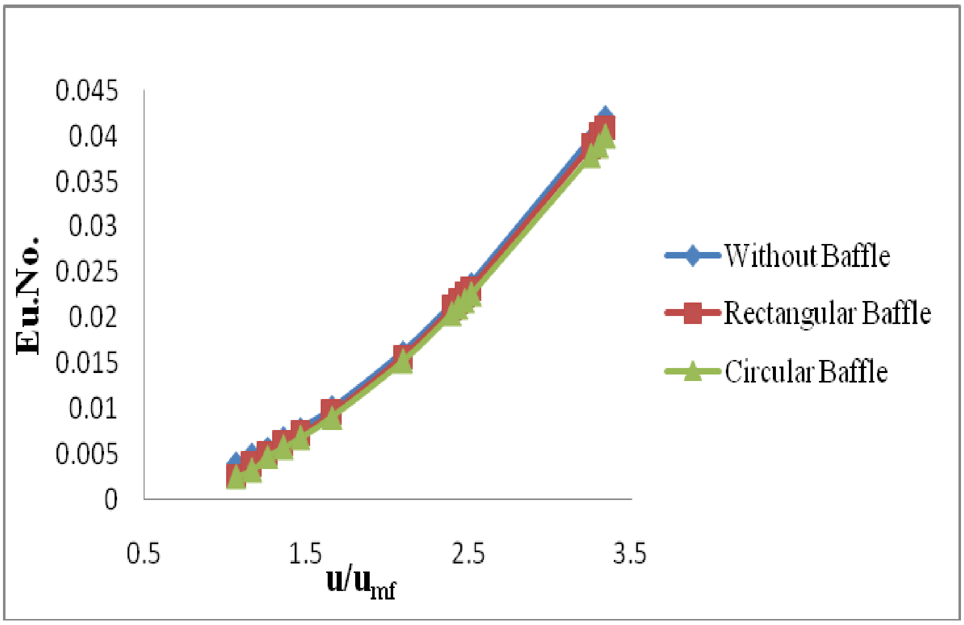


Figure (14) The relation between Eu. No. and velocity ratio u/u_{mf} for dp=205μm, Hs=0.3m fluidized bed.

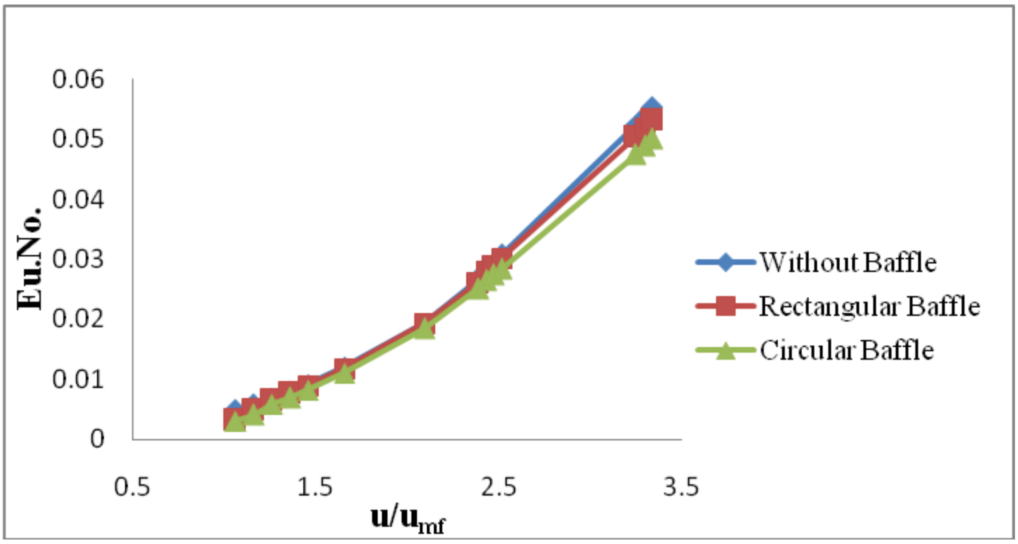


Figure (15) The relation between Eu. No. and velocity ratio u/u_{mf} for dp=205µm, Hs=0.4m fluidized bed.

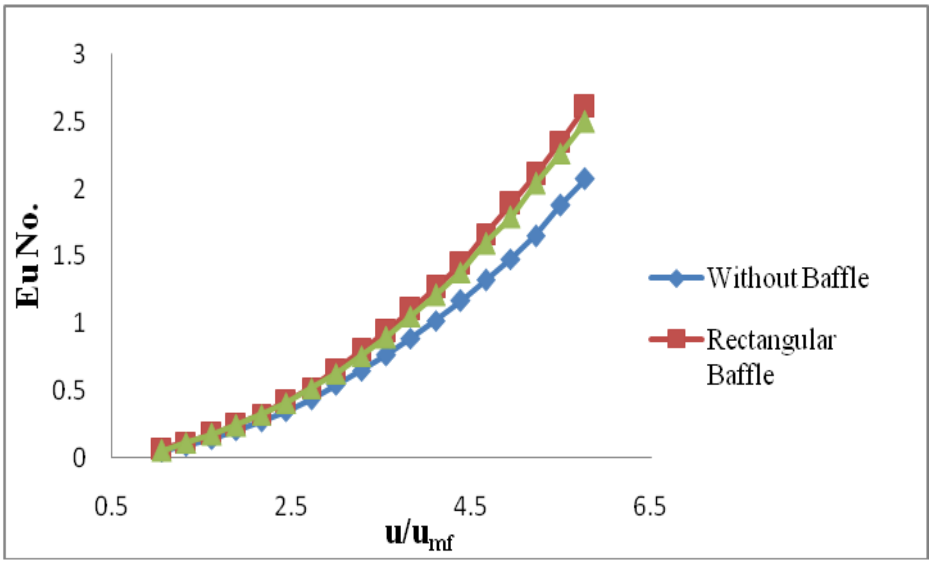


Figure (16) the relation between Eu.No. and velocity ratio u/u_{mf} for dp=424µm, Hs=0.3m fluidized bed.

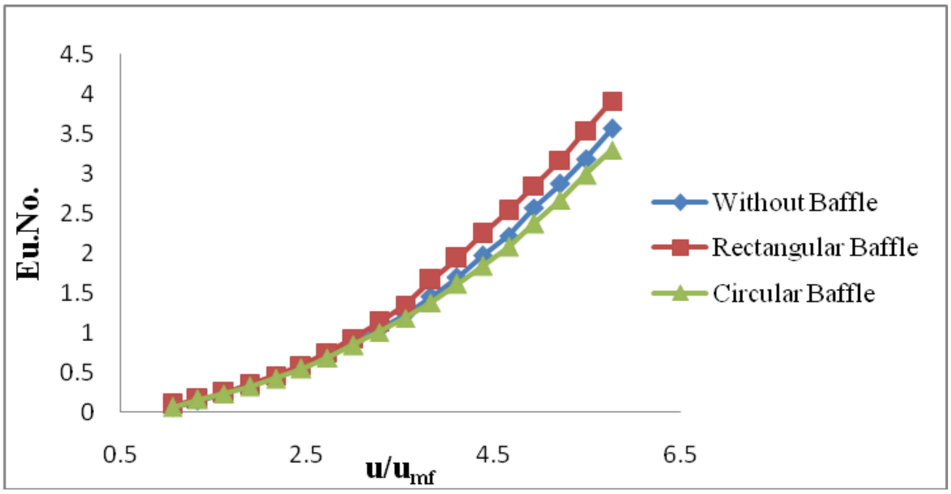


Figure (17) The relation between Eu.No. and velocity ratio u/u_{mf} for dp=424μm, Hs=0.4m fluidized bed.

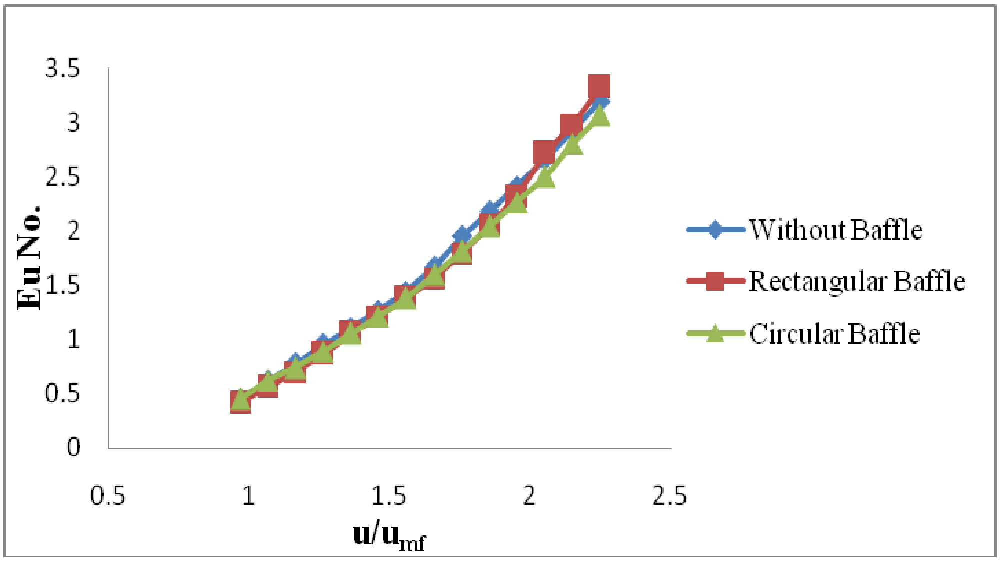


Figure (18) The relation between Eu.No. and velocity ratio u/u_{mf} for dp=775μm, Hs=0.3m fluidized bed.

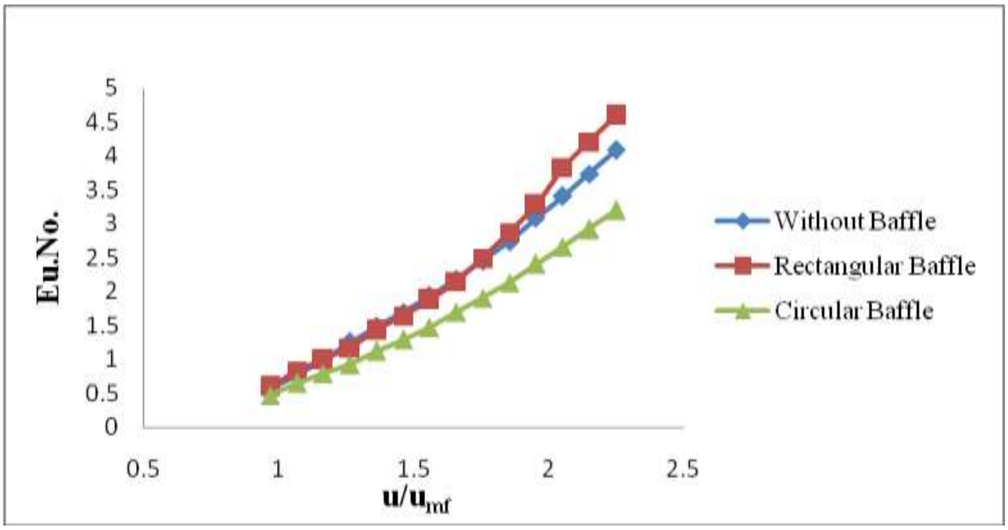


Figure (19) The relation between Eu.No. and velocity ratio u/u_{mf} for dp=775μm, Hs=0.4m fluidized bed.

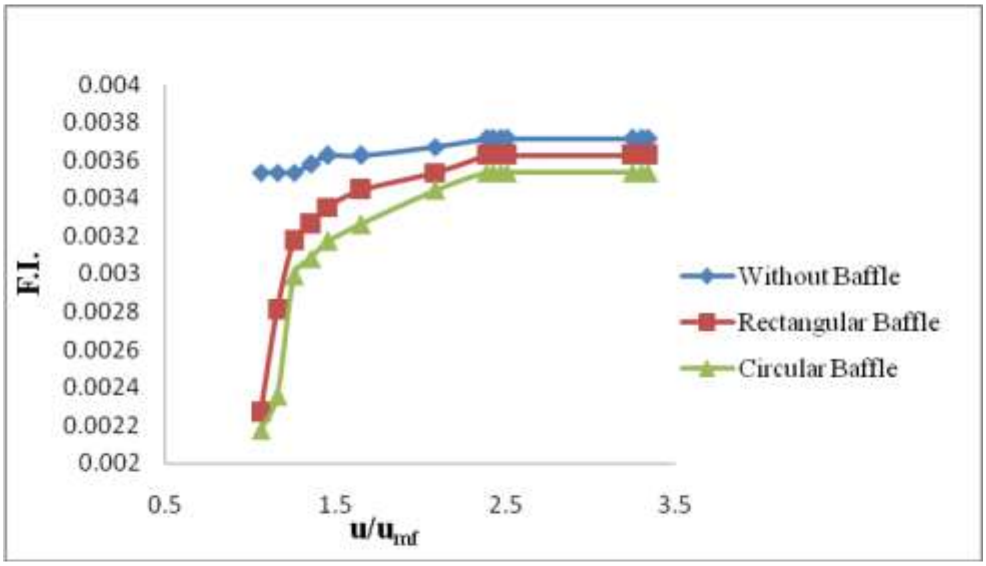


Figure (20) The variation of F.I. with velocity ratio u/u_{mf} , dp=205μm, Hs=0.3m.

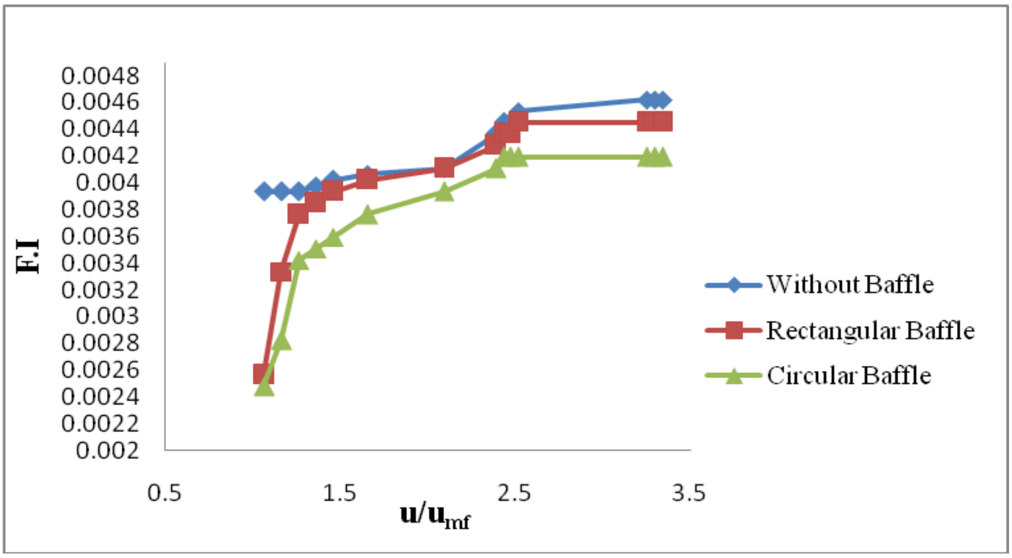


Figure (21) The variation of F.I. with velocity ratio u/u_{mf}, dp=205µm, Hs=0.4m.

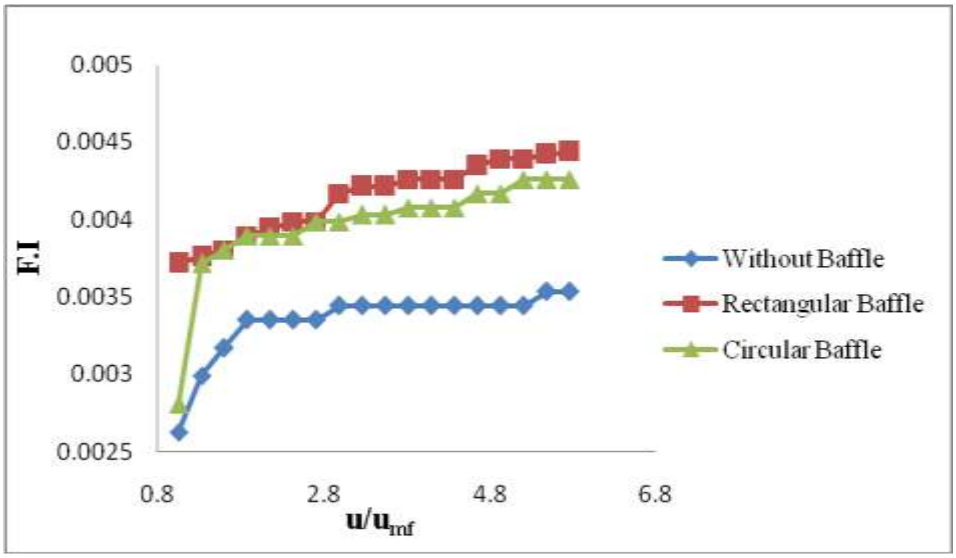


Figure (22) The variation of F.I. with velocity ratio u/u_{mf} , dp=424µm, Hs=0.3m.

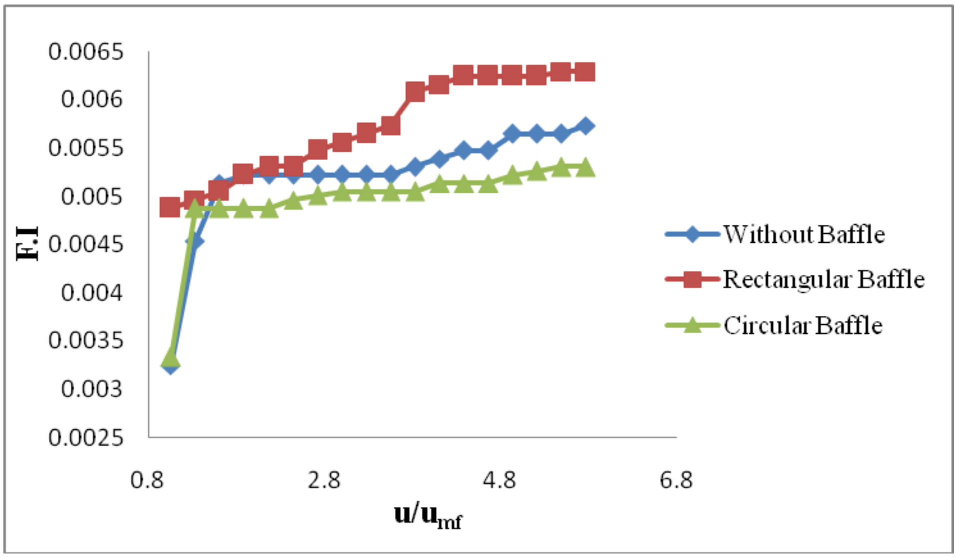


Figure (23) The variation of F.I. with velocity ratio u/u_{mf} , dp=424μm, Hs=0.4m.

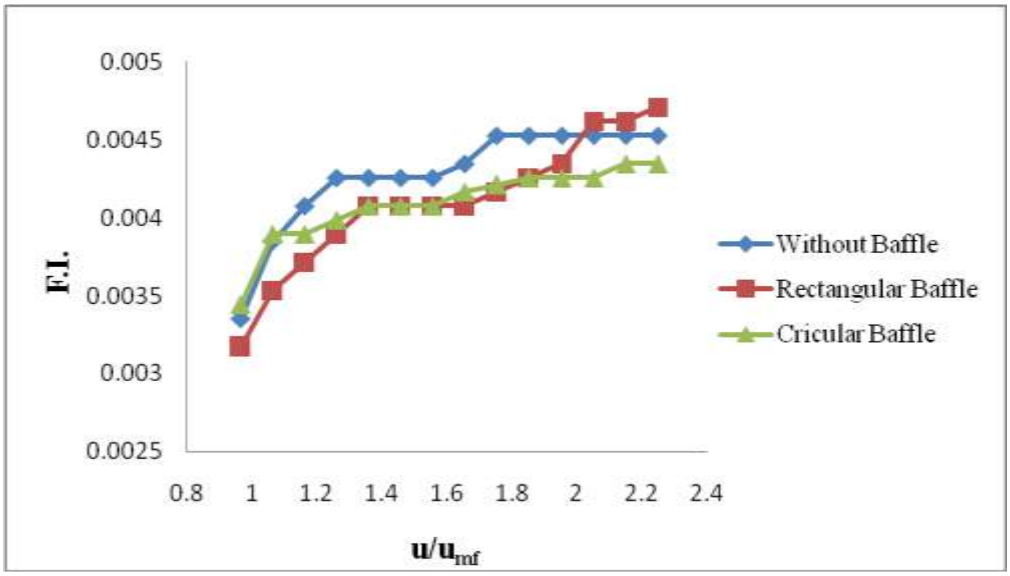


Figure (24) The variation of F.I. with velocity ratio u/u_{mf}, dp=775μm, Hs=0.3m.

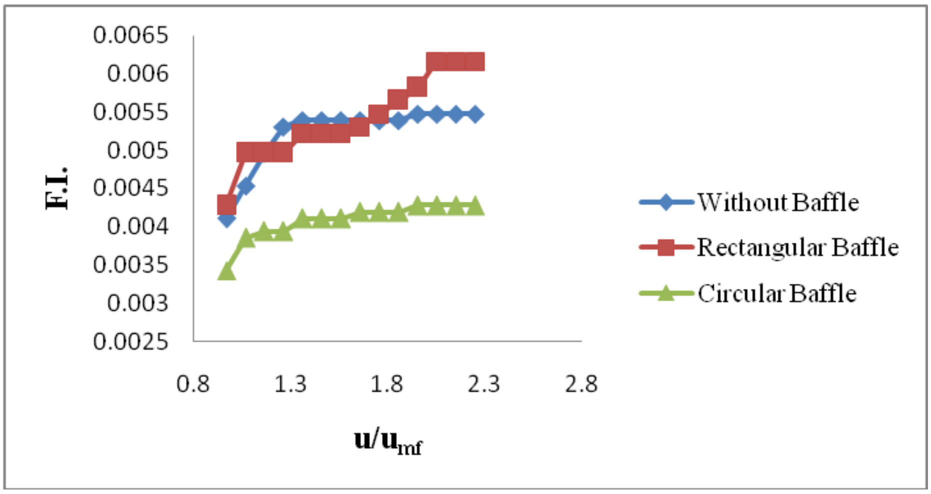


Figure (25) The variation of F.I. with velocity ratio u/u_{mf}, dp=775μm, Hs=0.4m.

NOTATION:

u	Superficial air velocity	m/s
u_{mf}	Minimum fluidization velocity	m/s
H	Bed height from distributer	m
Δp	Pressure drop	N/m^2
ρ_g	Gas density	kg/m^3
ρ_p	Solid particle density	kg/m^3
d_p	Particle diameter	m
Eu	Euler No. Dimensionless ($=\Delta P/\rho u^2$)	(-)
F.I.	Fluidization index ($=\Delta P/(w/A_c)$)	(-)
w	Weight of bed	kg
A_c	Column Area	m^2
D_c	Column diameter	m
u_t	Terminal velocity	m/s
R	Expanding bed ratio H/Hs	(-)
Re_t	Reynolds number ($=u_t d_p \rho/\mu$)	(-)
Ar	Archimedes number ($=d_p^3(\rho_p - \rho_g)\rho_g/\mu^2$)	(-)
G_r	Mass velocity ratio $(G-G_{mf})/G_t-G_{mf})$	(-)
G	Mass flow rate	$\frac{kg}{m^2.sce}$

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