

THE INFLUENCE OF AIR DISTRIBUTORS ON HEAT TRANSFER COEFFICIENT IN FLUIDIZED BED OF HEAT PIPE HEAT EXCHANGER⁺

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Abstract

Heat exchanger made of heat pipes elements immersed in a fluidized bed represents one of the most economic devices used in recovery of waste heat energy. In the present work ten copper heat pipes (each 22.2mm diameter and 550 mm length), fixed in staggered arrangement with (175mm) static bed height, and (176,353,707 μ m sand mean particle diameter) were used . Two types of air distributors (bubble cap and perforated plate) were used with an open area of (21.6%, 6.36%, 3.97%). The study covers the air mass velocity from (0.31 to 0.57 Kg/m² Sec), with hot bed temperature of 160C^o. The results show that the overall heat transfer coefficient for the heat exchanger using bubble cap distributor is slightly higher than that for perforated plate, while the power consumption was greater by 30% for bubble cap distributor. Bubble cap distributors pressure drop was 90% greater than perforated plate distributor for the same operating conditions. At the same time , it shown that heat transfer coefficient is increased with open area increase, while opposite response was found for power consumption .

Key word: Fluidization, Heat exchanger Heat distribution, Power consumption

المستخلص

تمثل المبادلات الحرارية ذات الانابيب الحرارية المغمورة في وسط مميح إحدى الوسائل الفعالة لأسترداد الطاقة

الصناعية. تناول البحث دراسة انتقال الحرارة في مبادل حراري مصنوع من عشرة أنابيب حرارية نحاسية بطول

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٥٥٠ ملم وبقطر ٢٢,٢ ملم مغمورا في طبقة ممبعة بالغاز ومرتبطة بشكل متخالف . مستخدمين ثلاثة اقطار لحبيبات

الرمال (١٧٦,٣٥٣,٧٠٧ مايكرومتر) بمعدل ارتفاع للطبقة الممبعة ١٧٥ ملم.

تم حساب معامل انتقال الحرارة مع تغير النسبة المئوية للمساحة المفتوحة للهواء

(٢١,٦%, ٣٦,٣%, ٣٩,٧%) لنوعين من الموزعات الهواء (مثقب وفقاعي) . أجريت التجارب عند درجة حرارة

١٦٠ م° للطبقة الساخنة ومعدل تدفق هواء (٠,٥٧ - ٠,٣١ كغم/م^٢ ثا) .

أظهرت النتائج أن معامل انتقال الحرارة للمبادل الذي يستخدم موزع الهواء الفقاعي كانت أكثر من موزع

الهواء المثقب، فيما كان معدل استهلاك القدرة أكثر وبنسبة ٣٠% لموزع الهواء الفقاعي . كما أن هبوط الضغط

عبر الموزع الفقاعي يصل إلى ٩٠% أكثر من موزع الهواء المثقب تحت نفسها الظروف التشغيلية . واخيراً يمكن

القول أن معامل انتقال الحرارة يزداد بزيادة النسبة المئوية للمساحة المفتوحة لموزع الهواء ،بينما يقل معدل

استهلاك القدرة .

الكلمات المفتاحية : التميع ، مبادل حراري ، موزع الهواء، استهلاك القدرة.

Notation:

Sub Script

A

Bed cross-section area(m²).

b

Bed

A_{ex}	Total external surface area of heat pipe at evaporator and condenser section (m^2).	c	Cold
C_p	Specific heat capacity of fluidized air ($J/Kg^{\circ}C$).	con	Condenser
d_p	Mean particle diameter (μm).	d	Distributor
d_o	Distributor orifice diameter (m).	e	Exit
G	Air mass velocity (Kg/m^2s).	ev	Evaporator
g	Acceleration due to gravity ($9.81 m/s^2$).	ex	Heat exchanger
H	Bed height (m).	f	Fluidized
h	Bed – surface heat transfer coefficient ($W/m^2^{\circ}C$).	g	Gas
M	Fluidized particle mass (Kg).	$H.P$	Heat pipe
m°	Air mass flow rate (Kg/s).	h	Hot
N	Number of orifice per unit area of distributor (m^{-2}).	i	Inlet
P	Pitch (m).	o	Outlet
ΔP_d	Pressure drop across distributor (N/m^2).	p	Particle
ΔP_b	Pressure drop across bed (N/m^2).	s	Solid
ΔP	Total pressure drops (N/m^2).		
Re_p	$= \frac{d_p u_s r_g}{m}$ Particle Reynolds number		
T	Temperature ($^{\circ}C$).		
u_B	Bubble velocity (m/s).		
u_s	Superficial gas velocity (m/s).		

u_{mf} Minimum fluidizing velocity (m/s).

GREEK SYMBOLS:

ρ Density (Kg/m³)

μ Viscosity (Kg/m s)

ε Void friction

Introduction

Heat exchangers made of heat pipes have now become one of the most effective and economic devices for the recovery of waste heat energy. The studies of overall performance of heat pipe exchangers using a bank of heat pipes as heat elements, were still limited.

Huang and Tsuei [1] present an analytical method for the prediction of the thermal performance of such heat exchangers based on the conductance model. Hsieh [2] report the thermal performance of a bank of heat pipes applied to a waste heat recovery.

Design procedure to predict the performance of air –to-water heat pipe heat exchanger is presented by Azad And Gibbs [3] based on an NTU-effectiveness approach to deduce heat transfer characteristics. Hsieh and Huang [4] present a numerical calculation of thermal performance and pressure drop for the heat pipe heat exchangers with aligned tube rows. It was found that counter-flow exhibits a better heat transfer rate.

There are many limitations such as fouling of the heat pipes due to dirty dust laden gases and the low gas –to surface heat transfer coefficient. These limitations result in the requirement for a large extended surface area. To overcome the above limitations, the heat pipes are located in fluidized bed.

Fluidized beds have been used extensively in heat exchanger, in both physical operations and chemical processes, because of their unique ability to rapidly heat transport and maintain of a uniform temperature. Although gas – fluidized beds are characterized by their good heat transfer properties, there are many instances when heat transfer remains a limiting factor in the design of a given system.

The heat transfer coefficient between the immerse tubes and a fluidized bed is very complicated because of the many fluidized bed variables : particle size and distribution, gas and particle thermal properties, fluidizing velocity, distributor design, bed design, operating temperature and pressure, and heat pipe design.

The performance of the gas distributor often determines success or failure of a fluidized bed. Particle and gas properties play a key role in successful design together with the critical pressure drop ratio ($\Delta P_d/\Delta P_b$), hole size, geometry and spacing; these influence jet penetration, dead zone, particle sifting, attrition and mixing.

Geldart [5] reports that distributors having holes smaller than 1mm diameter are expensive .On the other hand if holes are larger than $(5dp)$ the bed will drain into the wind box when de-fluidized. To overcome this problem bubble-cup air distributor was commonly

used. Also, if the number of orifice per unit area (N) is too small (large P and d_o) the problem of jets penetration and inadequate particle movement may occur. Whether or not there are jets or rapidly coalescing bubbles, it is certain that if surfaces such as heat – exchanger tube are positioned too close to the distributor, erosion by the particles can occur. If the bed is shallow, break through of jets to the surface may occur and this is usually undesirable. The particle movement induced by the gas issuing from a hole depends on the flow properties of the solids (practically characterized by u_{mf}) and the gas flow rate per hole. At low gas flow rates there is little movement, aerating all the solids at the plate. However, the small bubbles produced at plate do not have sufficient energy to cause the rigorous movement required to remix larger particles which may have segregated or de-fluidized. On the other hand, if drilled plates having large holes are used, the large distance between them (for a given pitch /diameter ratio) allows solids to settle out so, there is a minimum gas velocity to avoid dead areas between holes. It is worth saying that orifices that are too small are liable to be clogged whereas those that are too large may cause an uneven distribution of gas [6]. A plate having percent of open area corresponding to the smaller and larger values will in general be inappropriate for efficient and uniform operation. At smaller values of percentage open area, the distributor plate will have greater pressure drop than necessary for a smooth fluidization and will require greater energy consumption, at no benefit to the quality of fluidization [7] . The distributor pressure drop was found to increase with fluidizing velocity, and decrease with percentage open area of the distributor plate, and is independent of the bed weight or height for a given distributor design.

The simplest fact about a fluidized system is that the drag exerted by the gas flow supports the weight of the bed, and the pressure drop across the bed is independent of the gas velocity and this has been confirmed experimentally at moderate gas flow rate for uniformly

fluidized system [8]. Sathiyamoorthy et. al.[9] conclude that the choice of the distributor to bed pressure drop ratio ($\Delta P_d/\Delta P_b$) for stable operation of a fluidized bed depends on the (u_s/u_{mf}) ratio. In the range of ($0.1 < \Delta P_d/\Delta P_b < 0.3$) the system may not be expected to be in stable operating condition due either to only a limited number of operating orifices, or to non-uniformity in fluidization.

It was found that the pressure ratio ($\Delta P_d/\Delta P_b$) increases rapidly with increase in fluidizing velocity [7].

The behaviour of solids fluidized by gases falls in to four clearly recognizable groups , characterized by density difference ($\rho_s - \rho_g$) and mean particle diameter [10]. Group B [with ρ_s between $400-4000 \text{ Kg/m}^3$ and d_p in the range $20-40\mu\text{m}$] exhibits much less stable bed expansion , free bubbling commences at or a little above the minimum fluidizing velocity . While Group D [With $d_p > 600\mu\text{m}$] is of large size and /or density and spouts readily with relatively poor solid mixing . The power consumption is an important factor in any process using fluidized beds and occasionally it can be so high that it cancels the advantages of this type of operation .High pressure drops may significantly increase the power consumption of the blowers often a major cost factor . High-pressure drop can also hinder the circulation of solids [6].

Experimental rig.

The schematic diagram of the present test rig is explained in Fig.1. The fluidized bed consists of two adjacent fluidized beds separated by an insulation partition wall. Each one

contains fluidized bed, distributor and wind box flanged together. Each bed is 30x15cm in cross section with an overall height of 63cm. A sight glass of 5x55cm is located on the front side to show the regimes of fluidization. The bed contains the bed material and part of heat pipes

A staggered arrangement ($X/D=2$, $S/D=2$) of ten heat pipes were emerged between the hot and the cold beds as shown in Fig.1. It is well established that bed – surface heat transfer coefficient increases using staggered heat pipe arrangement, and with increasing bed temperature [12][13][14]. For these reasons, it was decided to do the experiments with staggered arrangement rather than square array with hot bed temperature of 160C. The specification of each heat pipe is listed in Table 1.

Two types of air distributors were used as shown in Fig.1., Three were perforated plates with different open area , and one distributor with bubble caps were investigated. Weir mesh was fixed under the perforated plate to avoid sand particle dropping into the wind box and to produce a uniform flow. The specifications of these tested plates are shown in Table 2.

The graded sand was used as a fluidized bed material, the static bed height of each beds was 175mm. Three sand particle types with mean diameter of (176, 353, 707) were investigated. The size distributing and the mean particle diameters of each were determined by standard sieving techniques.

The fluidized air was supplied from two centrifugal blowers sending air through 101.6mm galvanized pipe. The supplied air for each fluidized bed was independently regulated by gate

valve located at the blower suction. In order to simulate a fluidized bed that was being heated by a hot flue gas stream, an electrical air heater was manufactured and located before the wind box. Four electrical heaters each of 5KW power were used. The power supply to the heater was connected to the temperature control which allowed easy control on hot bed temperature. The airflow rate in each of the two supply lines was measured by an orifice plate meter, which was designed and manufactured according to BS1042.

Thirty copper-constantan thermocouples connected to digital temperature recorded were used to measure the temperature around the system as shown in Fig.1. Static pressure tapings connected to water manometers were used to measure the pressure drop across the distributor and the fluidized bed at both hot and cold beds.

Calculations

The distributor to bed pressure drop ($\Delta P_d/\Delta P_b$) can be calculated experimentally or using the experimental correlation [9]:

$$\frac{\Delta P_d}{\Delta P_b} = 2.7 \left(\frac{u_{mf}}{u_s - u_{mf}} \right)^{2.32}$$

The pressure drop across the bed can be determined by [6]:

$$\Delta P_b = H_{mf} (1 - e_{mf}) (r_s - r_g)$$

$$\text{and } e_{mf} = 1 - \frac{M_s}{r_s (A H_{mf})}$$

The minimum fluidizing velocity can be observed or calculated using the following equation [6]:

For small particles:

$$u_{mf} = \frac{dp^2(r_s - r_g)g}{1650m} \quad \text{With } Re_p < 20$$

For large particle

$$u_{mf}^2 = \frac{dp(r_s - r_g)g}{24.5r_g} \quad \text{With } Re_p > 1000$$

Bed – Surface heat transfer coefficient for the tubes is calculated [11]:as follows

$$h_{ho} = \frac{Q_{H.P}}{A_{ho}(T_{hb} - T_{ev})}$$

$$h_{co} = \frac{Q_{H.P}}{A_{co}(T_{con} - T_{cb})}$$

Under steady state condition ,the heat transfer rate by the heat pipe , $Q_{H.P}$,

is calculated from the energy balance on the cold bed taking into consideration heat transfer to the surrounding and heat transfer through the partition:

$$Q_{H.P} = m'_c[(Cp_f T)_{ce} - (Cp_f T)_{ci}] - Q_{partition}$$

The performance evaluation criteria used to study the influence of air distributors configuration on the heat exchange surface based on the method of Kays [16] consists of plotting mean heat transfer coefficient h for the geometries under consideration against their pumping power requirements. The best system is that which produces the highest heat transfer coefficient for a given value of pumping power. The pumping power can be defined by the following equation.

$$E = \frac{m}{r_g}(\Delta P_b + \Delta P_d)$$

Results And Discussion

Mean particle diameter:

Fig (2) and Fig (3) show the variation in maximum bed – surface heat transfer coefficient with air mass velocity for different mean particle diameter in hot and cold beds respectively. It is clear that for the same open area, decreasing the mean particle diameter results in an increasing in bed surface –heat transfer coefficient due to an increase in average gas conduction paths between the first row of particle and the pipe surface which coincides with literature [5][11][15].

Fluidizing velocity:

Refers to Fig (2) and Fig (3), it is clear that increasing fluidizing velocity results in an increase in the bed – surface heat transfer coefficient due to decreasing the particle residence time on the pipe surface. The figures shown also, that the increment of heat transfer is small, since the mass velocity increases by 80% resulting in an average increasing in heat transfer by 15%.

Rong and Chang [15] state that when the superficial velocity is increased, the jetting velocity is also increased which causes the heat transfer coefficient to increase in the grid region. At the same time the particle packing density is lower for high velocity which causes a reduction in heat transfer coefficient. The opposing effect of these two parameter is that the heat transfer coefficients reach a maximum value, and then drop [8], [12], which is out of range in the present work.

Percentages of open area and distributor type:

The variations in bed - surface heat transfer coefficient with air mass velocity for different percentage open areas in hot and cold bed are shown in Fig (4) and Fig (5). It is clear that increasing the percentage of open area, increases the bed – surface heat transfer coefficient due to enhancing of the particle mixing. For the same open area the type of distributor plate has a little effect on the heat transfer coefficient, since each type almost supplies the same flow rate to the bed.

Distributor pressure drop (ΔP_d):

Fig (6) shows the variation in distributor pressure drop versus the fluidization rate for different open area, while Fig (7) shows the variation in distributor pressure drop versus the

percentage open area for variable air mass velocity in perforated plate. It is concluded that (ΔP_d) increases as the fluidization rate increases, and the percentage open area decreases.

Distributor pressure drop sharply increases with gas velocity, as it causes a rise in a frictional resistance due to local fluid flow. At smaller values of percentage of open area, the distributor plate will have greater pressure drop than necessary for smooth fluidization [7].

The satisfactory or stable bed operation implies that all holes are in continuous operation. Non-operational holes are surrounded by immobile solids and though they deliver gas to the bed, it is not sufficient to produce bubbles [5].

It is worth noting that for the same open area and fluidizing velocity, (ΔP_d) is higher for the bubble cap distributor within (90%) than perforated plate, This may be to high resistance in gas flow due to changing the air path before entering the bed.

Fig (8) represents the variation in distributor pressure drop with air mass velocity for different mean particle diameters. Particles with high mean particle diameter need higher pressure drop because to increasing the (U_{mf}) an (U_s) need good fluidization.

Bed-pressure drop (ΔP_b):

The pressure drop versus velocity diagram is a useful indication of the quality of fluidization especially when visual observation is not possible. The observed pressure drop data may deviate slightly from the calculated, which is attributed to the energy loss by

collision and friction between particles as well as between particles and the surface container and heating elements [6].

The variations in bed pressure drop with air mass velocity and (u/u_{mf}) for different percentage open areas are shown in Fig (9) and Fig (10).

Bed – pressure drop increases with decreases in percentage of open area due to increasing the superficial gas velocity. Large pressure drop fluctuations suggest a slugging bed, while abnormally low pressure drop suggests incomplete contacting with the particles, only partly fluidized bed results in a channelling [6]. It is very clear that (ΔP_b) has a limited increase for bed with bubble caps over those using perforated plate.

Furthermore for larger percentage of open area, (ΔP_b) is slightly affected by increasing the air mass velocity. With gas velocity beyond the minimum fluidization, the bed expands and bubble are seen to rise with resulting non-homogeneity in the bed for superficial velocity in the range between (0.38 – 0.7 m/s). With lower percentage of open area, there is sharp increase in (ΔP_b) .

The measured values of (ΔP_b) were seen to increase with decreasing mean particle diameter. Bed – pressure drop for dp (176.7 and 353.5 μm) has almost an equal value since it represents the same group classification [Group B], while it is a large deviation for (707.7 μm dp) which represent [Group D] [10]. That is clearly seen in Fig (11).

High pressure drop may significantly increase the power consumption of the blower, often a major or cost factor. High pressure drop can hinder the circulation of solids [6].

Fig (12) shows the variation in bed – surface heat transfer coefficient against the total power consumption for different percentages of open area. It is concluded that the maximum bed-surface heat transfer coefficient with lower power consumption can be achieved using large percentage of open area of open area and air mass flow rate, bubble cap distributor higher within 30 % than bed with the similar perforated plate.

Conclusions

- 1- Bed-surface heat transfer coefficient increases with air mass velocity merease for different mean particle diametes and open area of bubble cap and perforated plate.
- 2- Incrasing the percentage of open area results in an increase in the bed–surface heat transfer coefficient .The type of distributor has a little effect.
- 3- Distributor pressure drops increase with air mass velocity, percentage open area, and mean particle diameter. Bubble cap distributor has higher (ΔP_d) than that of perforated plate with the same open area.
- 4- Bed pressure drop increases with decreasing percentage of open area ,and mean particle diameter . Large pressure drop results in poor fluidization .
- 5- The total power consumption decreases as the percentage of open area increases, and results in high bed-surface heat transfer coefficient.

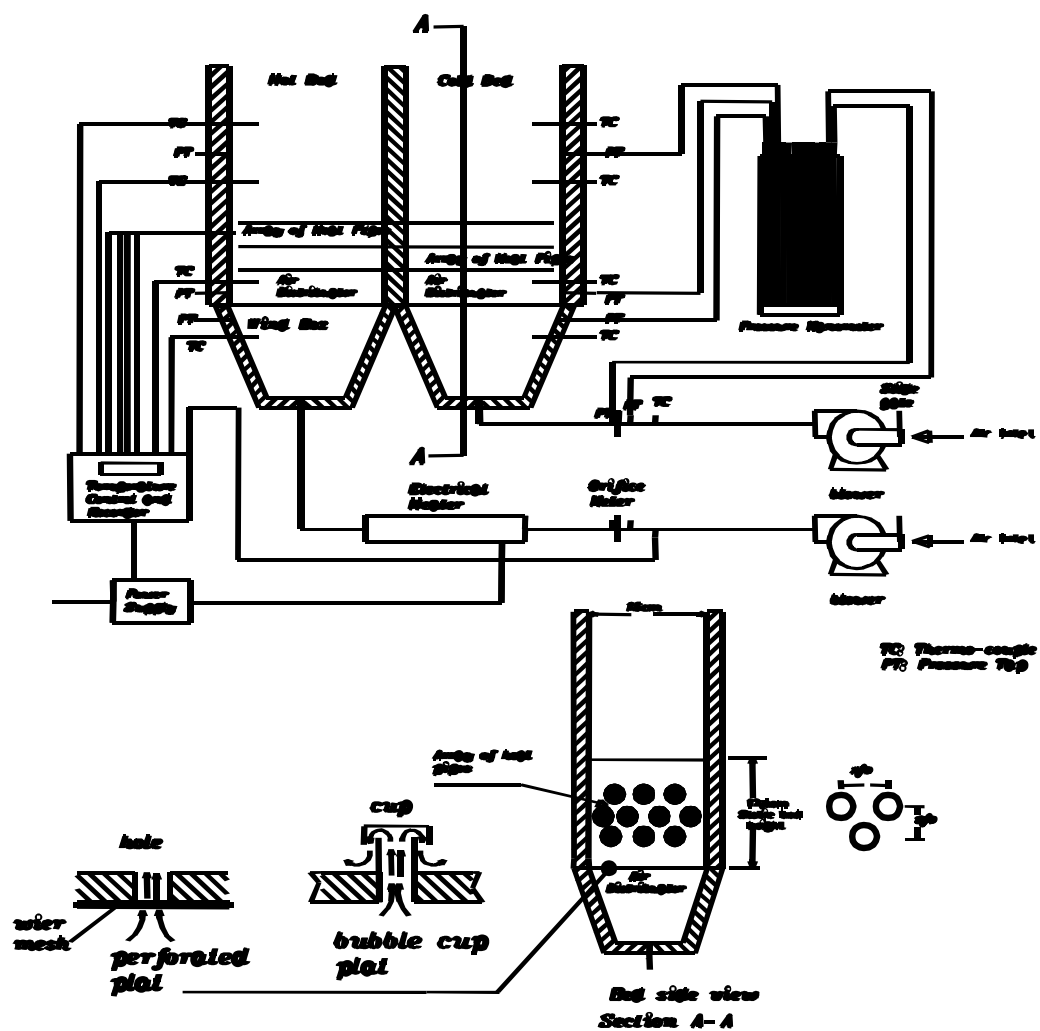
Table (1) the specifications of Heat Pipe.

Table (2) the specifications of Distributor Plate

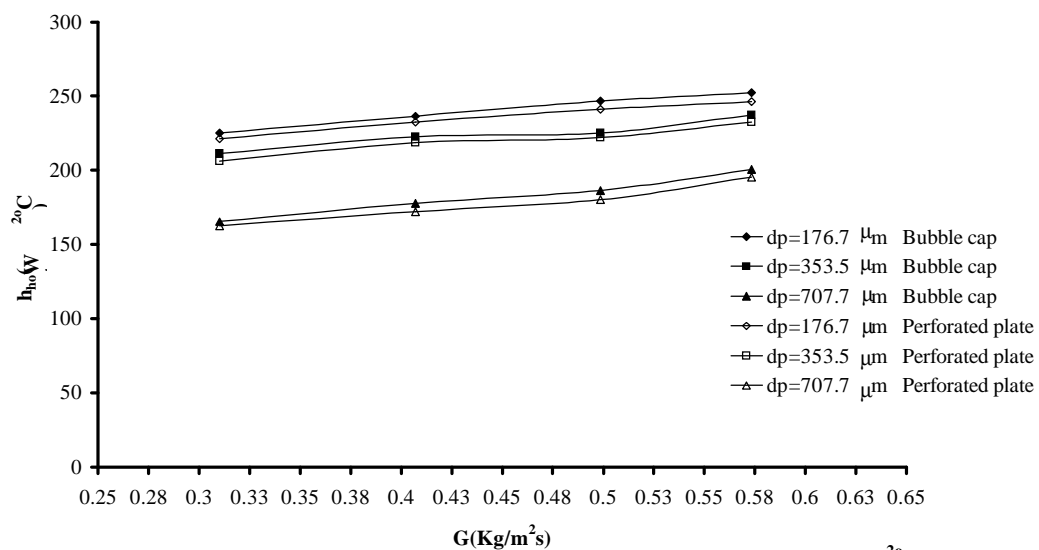
Tube	Copper
Outer diameter	Do=22.2mm
Inner diameter	Di=20.38 mm
Evaporator length	Le=275mm

Condenser length	Lc=275mm
Wick structure	Phosphor bronze 7 layer 120 inch ⁻¹
Work fluid	Distilled water =33.8 gram

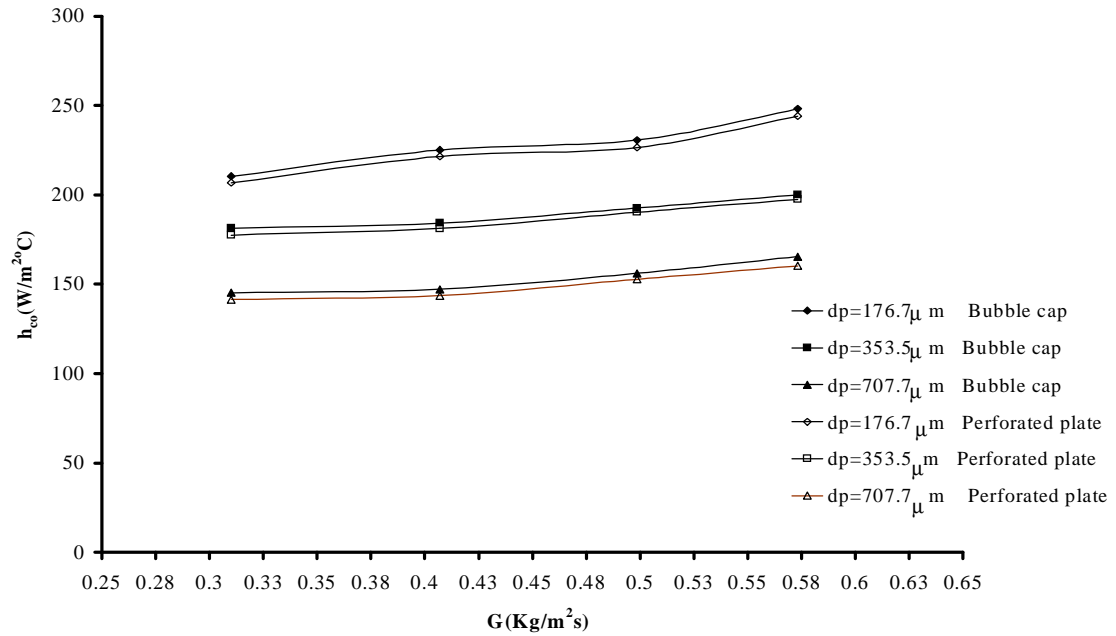
Distributors	A _{Perf.}	B _{Perf.}	C _{Perf.}	D _{Bubb.}
Percentage of open area	21.6	6.36	3.97	6.36
Pitch mm	10	20	25	20
No. Orifice	450	45	72	45
Orifice diameter mm	5	9	5	9



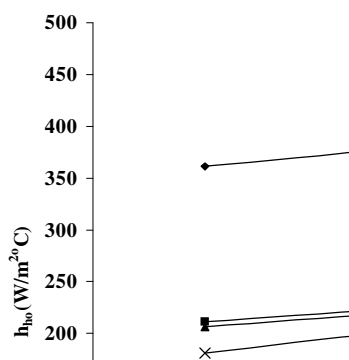
Fig(1) Schematic Diagram of the Test Rig.

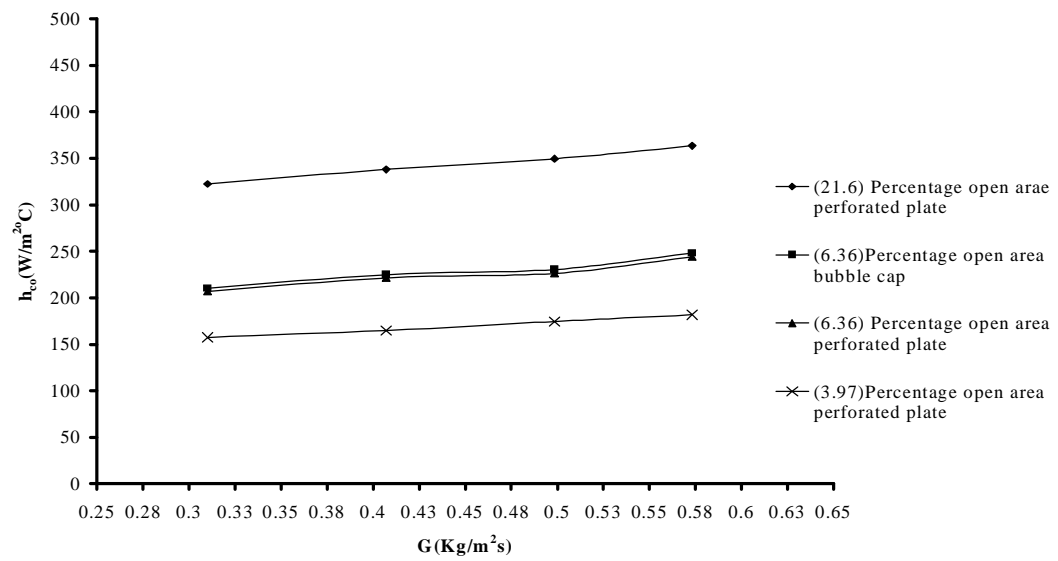


Fig(2) Variation of maximum bed surface heat transfer coefficient h_{ho} (W/m²°C) in hot bed with air mass velocity G (Kg/m²s) for 6.36% open area for perforated plate and bubble cap



Fig(3) Variation of maximum bed-surface heat transfer coefficient h_{co} (W/m²°C) in cold bed with air mass velocity G (Kg/m²s) for 6.36% open area for perforated plate and bubble cap





Fig(5) Variation of maximum heat transfer coefficient h_{co} (W/m²°C) in cold bed with air mass velocity G (Kg/m²s) for $d_p = 176.6$ mm

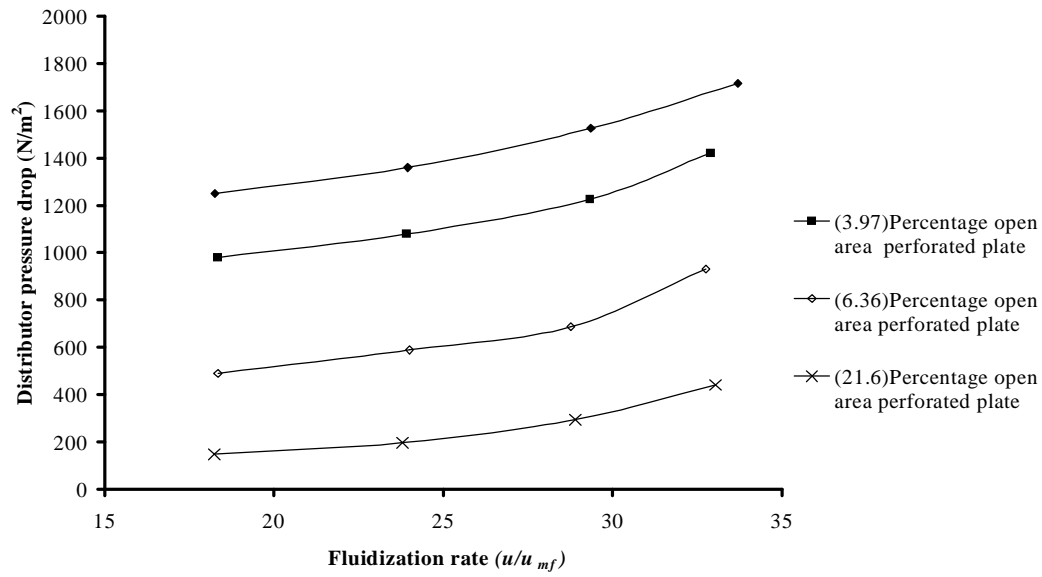
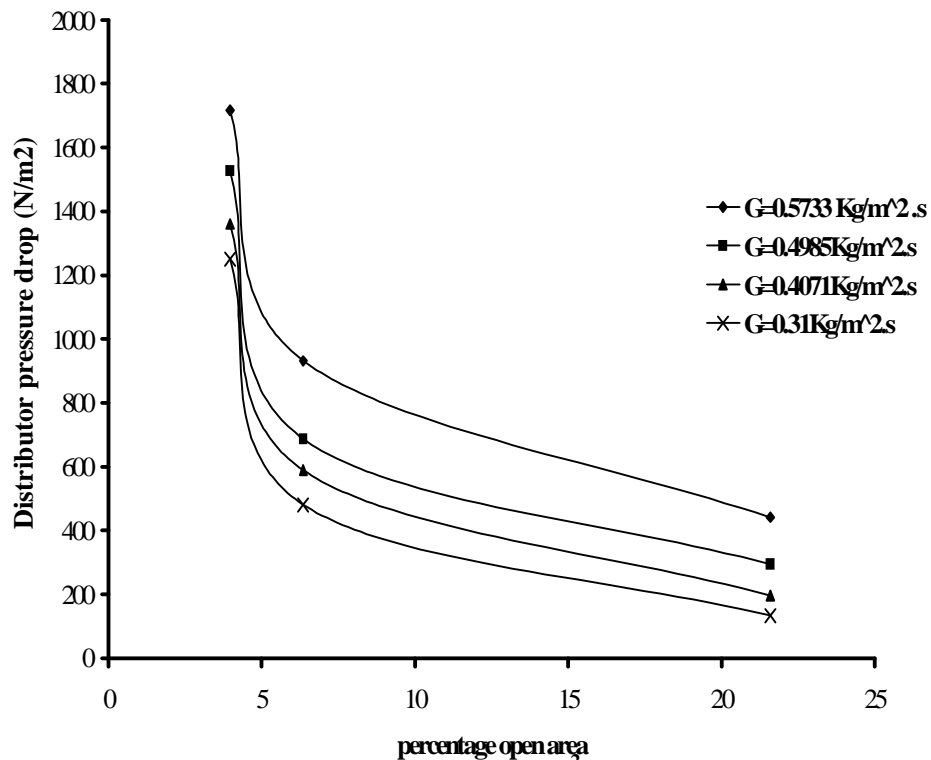
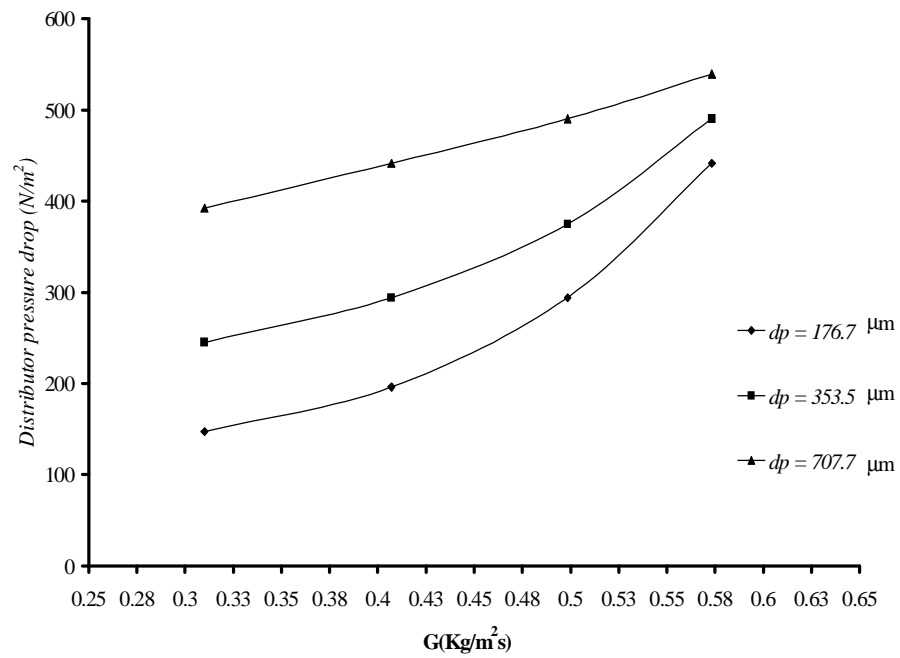


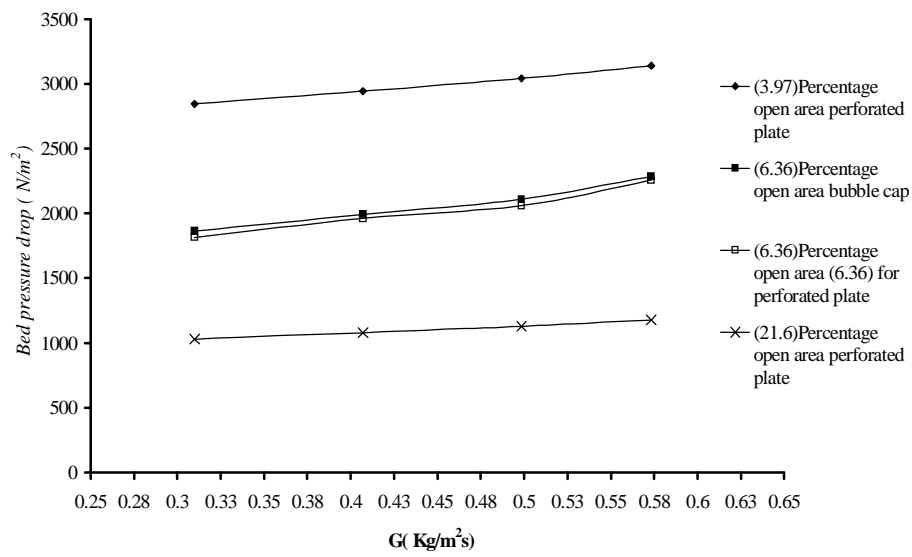
Fig (6) Variation of distributor pressure drop (N/m^2) with fluidization rate (u/u_{mf}) for $d_p=176.7$ mm



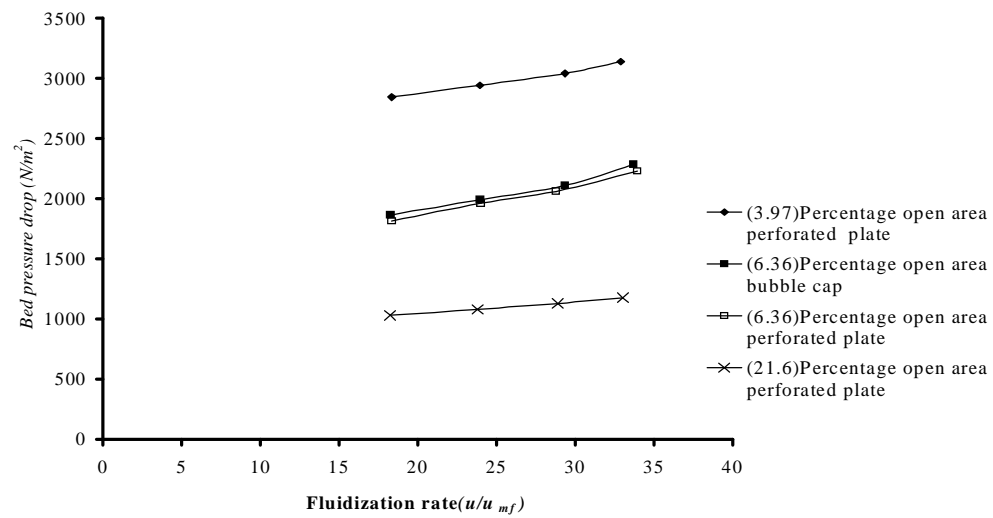
Fig(7) Variation of distributor pressure drop (N/m^2) with percentage open area perforated plate for $d_p=176.7$ mm



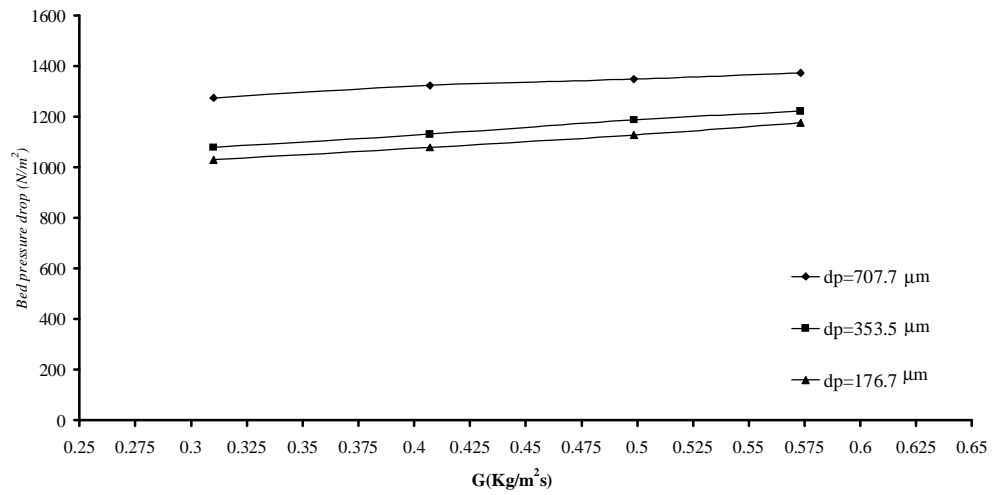
Fig(8) Variation of pressure drop (N/m^2) in distributor plate with mass flow rate per unit area G (Kg/m^2s) for (21.6) percentage open area perforated plate



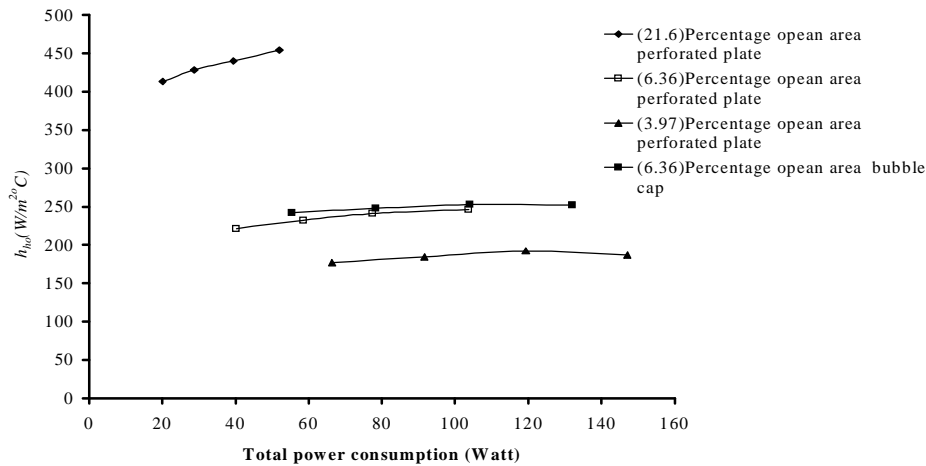
Fig(9) Variation of bed pressure drop (N/m^2) with air mass velocity $G(Kg/m^2 s)$ for $dp=176.7 \text{ mm}$



Fig(10) Variation of pressure drop (N/m^2) in hot bed with Fluidization rate (u/u_{mf}) for $dp=176.7 \text{ mm}$



Fig(11) Variation of pressure drop (N/m^2) in hot bed with air mass velocity G ($\text{Kg/m}^2 \text{s}$) for (21.6) percentage open area perforated plate



Fig(12) Variation of maximum heat transfer coefficient h_{bo} ($W/m^2°C$) with total power consumption (Watt) for $dp=176.7$ mm

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