# Experimental Study of heat Transfer between the Shallow Fluidized bed and a Tube Bundle Immersed in it

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## Abstract:

This paper presents an experimental study of heat transfer between a shallow fluidized bed and the surface of a single horizontal tube and a tube bundle, which is immersed in it.

Carbon, which is prepared from the Date stones, is used as a solid to be fluidized and a compressed air as an external fluid.

The results showed that the overall heat transfer coefficient of tubes bundle, which immersed in fluidized bed is lower than single tube.

The modified correlations and analysis are offered by experimental work, which shows the relations between heat transfer coefficient and superficial mass velocity.

**Key word:** heat transfer - heat exchangers - fluidized bed heat exchangers

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**Received 11 April 2005** **Accepted 23 January 2006**

## Nomenclature

| | | |
|---|---|---|
| A | Area of heating surface | $m^2$,$mm^2$ |
| $d_p$ | Average particle diameter | mm |
| $d_{pi}$ | Average diameter of the successive screens | mm |
| $D_t$ | bed diameter | m , mm |
| G | Superficial mass velocity | kg / $m^2$.sec |
| $G_{mf}$ | Minimum mass velocity at fluidization | kg / $m^2$.sec |
| G | Acceleration due to gravity | m / $sec^2$ |
| h | Overall heat transfer coefficient | W/ $m^2$.C |
| Q | Heat supplied | W |
| $T_s$ | Average surface temperature | K |
| $T_b$ | Average bed temperature | K |
| $W_i$ | Weight fraction of particles in a specified size rang | % |
| W | Weight of the bed | Kg |
| $\mu$ | Viscosity | kg / m.sec |
| $\rho_f$ | Fluid density | kg/$m^3$ |
| $\rho_s$ | Density of solid particle | kg/$m^3$ |
| $\rho_b$ | Bulk density | kg/$m^3$ |
| $\Phi_s$ | Shape factor of particles | |
| $\varepsilon$ | Bulk bed porosity | |
| Nu | Nusselt Number | |
| Pr | Prandtl Number | |

## Introduction:

Fluidized beds are widely used in industry for mixing solid particles with gases or liquid. In most industrial applications, a fluidized bed consists of a vertically oriented column filled with granular material, and a fluid (gas or liquid) is pumped up ward through a distributor at the bottom of the bed. When the drag force of flowing fluid exceeds gravity, the particles are lifted and fluidization occurs [1].

At very low speed flows the air will pass between the particles without disturbing their position. The bed is then called static bed and acts like a filter. By increasing the airflow to a higher level, some particles in the bed will start to move depending on the uneven air distribution in the bed.

Then when bed's depth more than 50cm, it is called deep bed [2], but if the depth is beds less than 50 cm, it is called a shallow bed.

At higher air flows the particles will be separated from each other and the bed volume increases. This is called incipient fluidization. Higher contact areas between air and the particles will be the result, which is considered an advantage during combustion [3]. Even at higher airflow the particles will be mixed with each other and the bed is more like a boiling fluid. Full fluidization. This is the domain where most advantages with fluidization can be found.

Continuous increase of the air flow means bigger size of the air bubbles and at the end these will occupy the whole width of the bed [3].

A uniform fluidization, which is the most desirable regime of operation of industrial fluidized bed, is prone to instabilities. As the fluid flow increases, bubbles of clear fluid are formed at the bottom of the bed and these bubbles travel to the surface.

Several measurements have been reported for heat transfer to or from a tube immersed horizontally in a fluidized bed [ 2-8 ]. The effects of fluidizing velocity, particle size, shape and density, tube diameter and the bed temperature on the heat transfer coefficient were investigated. It was observed that the value of the total heat transfer coefficient increase with increase in the bed temperature, decrease in the tube diameter, and increase in the air mass velocity.

Because of the heat transfer between the bed and immersed surface is a complicated process, despite the efforts of study for more than 20 years, there is no satisfactory model for the prediction of bed to surface heat transfer coefficient applicable even that the experimental data is empirically for practical design use.

## Experimental apparatus and procedure:

The experimental apparatus used in this study has been described and shown as Fig. 1 . The experimental column is of rectangular cross-section (300*175mm), the bottom plate is perforated distributor. For the single tube test, the tube is arranged horizontally of 100mm height from the distributor, tube diameter is 25mm; for the tube bundle test, the tubes arranged in three layers and the lowest layer of three tubes is put in the height 85mm from the distributor. The middle layer is of two tubes are in the height of 130mm from the distributor and the top layer of three tubes is in the height of 185mm from the distributor. That is means the distance between the different layers at in vertical direction is 50mm. In the horizontal direction, the distance between the lateral tubes is 50mm as shown in Fig.2. Therefore the vertical and horizontal relative pitches are 2, the ratio of horizontal pitch to tube diameter is called the relative horizontal pitch, and similarly the relative vertical pitch is also defined.

The air is supplied by the blower (2890r.p.m/ 3 phase / 7.5 kW )for atmospheric fluidization and the orifice is used for measuring the flow air. The manometer is used for measuring the pressure drop in fluidized bed. The heater chamber consists of a spiral ceramic core of 10mm diameter, with winding electrical resistance wire ( 1 kW / 0.5 mm diameter) for heating the tube by A.C. In the temperature measurement system, there are 12 pairs of thermocouples fitted on the surface of each tube, 8 pairs on the tube center and 2 pairs for ends. In the positions of 50 mm higher and lower than the centerline of heating tube, there are thermocouples installed for measuring the bed temperature.

The fluidized particles used are of carbons, which were pre prepared from the date stones, were ground to irregular shapes.

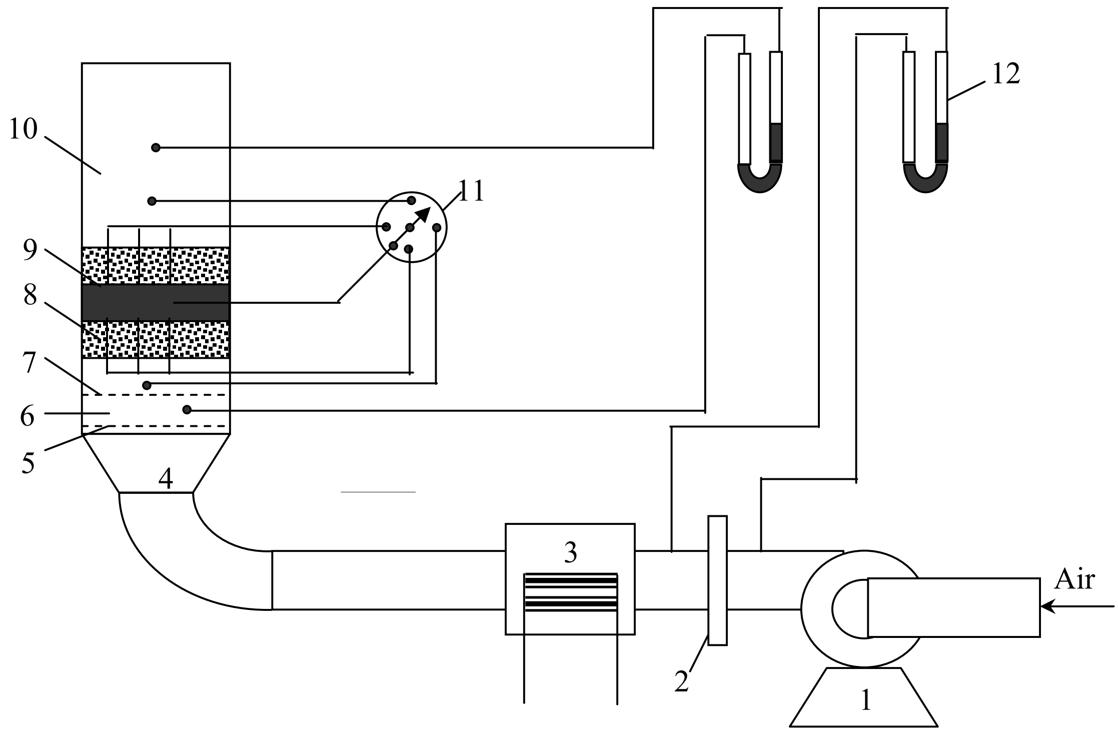


Fig (1) Scheme of Facilities

| | |
|---|---|
| Blower | 7. Distributor |
| Orifice | 8. Bed material |
| Heater chamber | 9. Immersed tubes |
| Air box | 10. Proper section |
| Air grate | 11. Potentiometer |
| Air baffler | 12. Manometer |

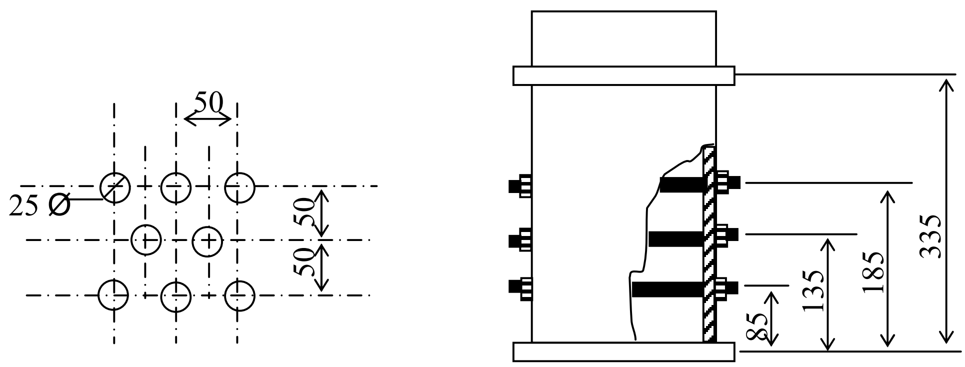


Fig (2) The Distance between the Different Layers

The average diameter is calculated as follows;

$$1/d_p = \sum_{i=1}^{n} ( w_i/d_{pi}) \qquad (1)$$

The shape factor of particle calculated by [9]:

$$\frac{1}{\Phi_s^2 \, \varepsilon_{mf}^3} \approx 14 \qquad (2)$$

And volume concentration for packed bed as [10] :

$$( 1- \varepsilon_0 ) = \rho_b/\rho_s \qquad (3)$$

Where the Bulk density calculated:

$$\rho_b = m / V \qquad (4)$$

Table ( 1 ) carbon specification, which is used in experimental work

| | |
|---|---|
| Average diameter | 1.402mm |
| Real density | 1495 Kg/m$^3$ |
| Bulk density | 788 Kg/m$^3$ |
| Shape factor | 0.670 |
| Volume concentration | 0.527 |

Table ( 2 ) The incipient mass velocity

| | | | |
|---|---|---|---|
| Fixed bed heights | 110 | to | 165 mm |
| Single tube | 0.12 | to | 0.15 kg / m$^2$ sec |
| Tube bundle | 0.15 | to | 0.20 kg / m$^2$ sec |
| Temperature of heating surface | 300 | to | 350 $^o$ K |
| Temperature of fluidized bed | 300 | to | 400 $^o$ K |

The overall heat transfer coefficient is calculated for steady condition according to the following formula:

$$h = \frac{Q}{A ( T_b - T_s )} \qquad (5)$$

## Results and analysis:

The over all heat transfer coefficient is obtained and is variation its shown with superficial mass velocity in figures (3 ,4 ).

Figure (3) shows that the heat transfer coefficient varies with the velocity approaching to an optimum for single tube, and fig ( 4 ) for the tube bundle. The magnitude from 60 to 380 W/m$^2$.C for single tube, and 30 to 280 W/m$^2$.C for tube bundle.

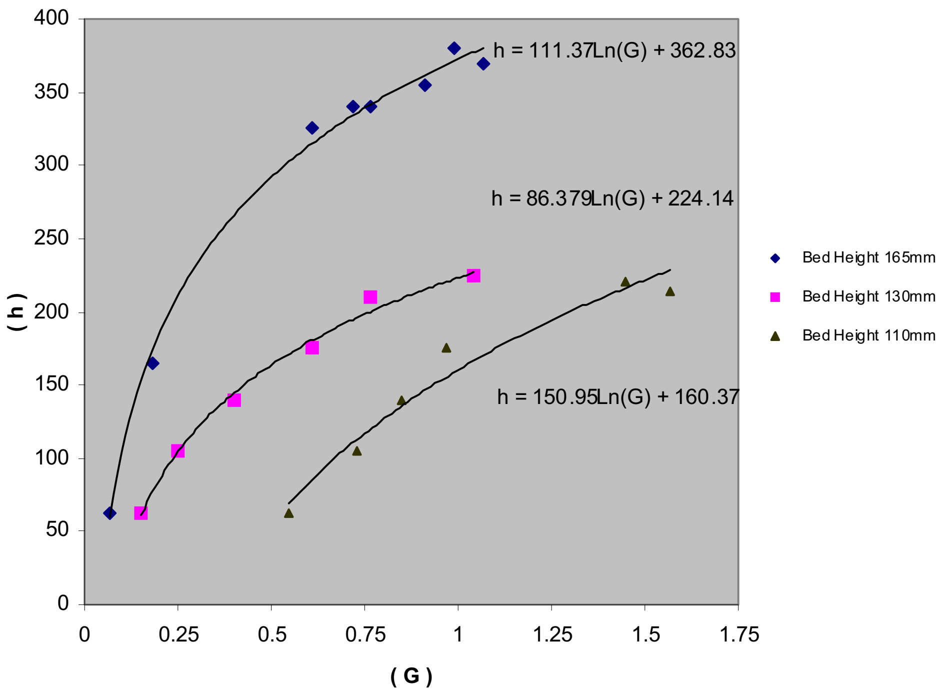


**Fig ( 3 ) Rrelations between Heat Transfer Coefficient and Superficial Mass Velocity for Single Tube**

The experimental correlation combines the heat transfer coefficient with superficial mass velocity presented in this work. For the single tube and packed bed height 163 mm is:

$$h = 111.37 Ln(G) + 362.83 \quad (6)$$

and the tube bundle under the same condition is :

$$h = 80.467 Ln(G) + 238.83 \quad (7)$$

This means that the heat transfer coefficient of single tube is higher than tube bundle under the same condition. The interstitial space is no good for the particles to play the role as the heat "carrier" and also the gas is more stagnant to heat transfer.

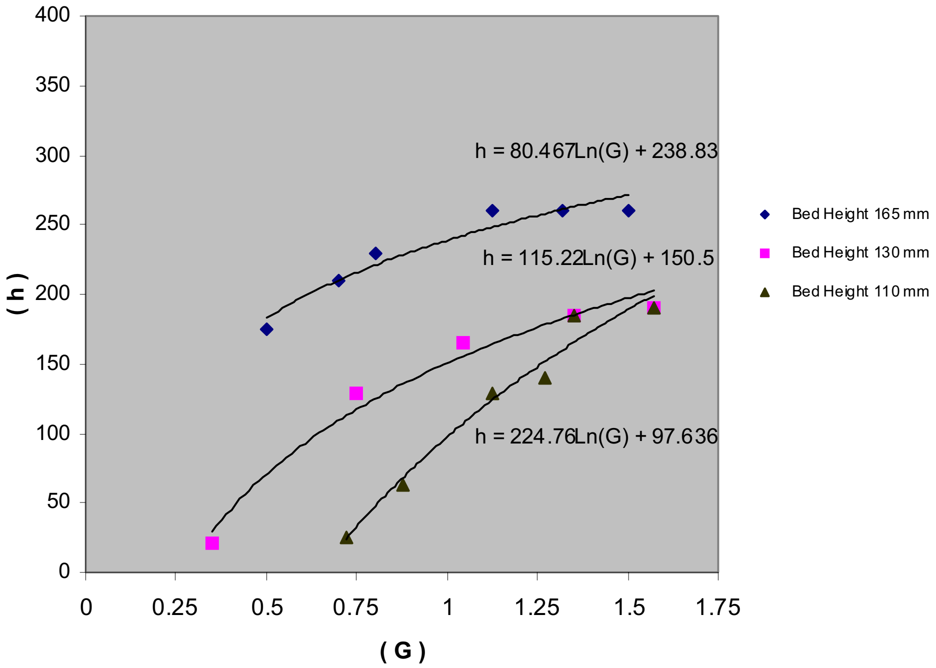


**Fig ( 4 ) Rrelations between Heat Transfer Coefficient and Superficial Mass Velocity for Tube Bundle**

Fig (5) show the overall heat transfer coefficient of different layers of the tubes position. The height of fixed bed is selected arbitrarily as 110mm. And the experimental correlation combines of the heat transfer coefficient with superficial mass velocity for the different layers of tube bundle presented in this fig.

For the bottom layer of tubes is:

$$h = 79.46Ln(G) + 323.5 \qquad (8)$$

Middle layer of tubes is:

$$h = 107.19Ln(G) + 246.52 \qquad (9)$$

Top layer of tubes is:

h = 84.099Ln(G) + 144.83 (10)

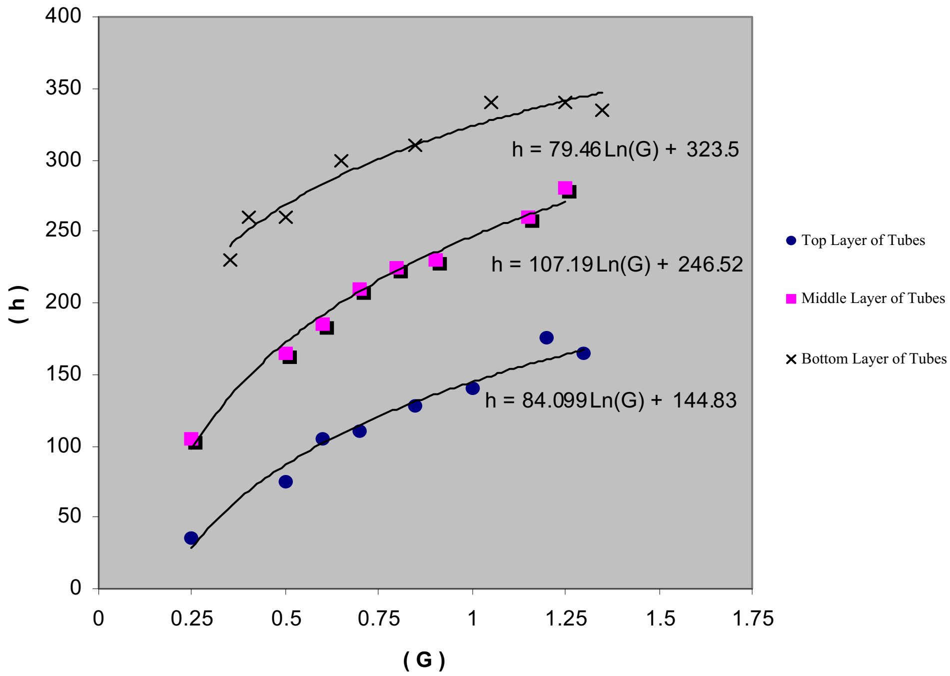


**Fig ( 5 ) Heat Transfer Coefficients for Different Layers of Tube - Bundle**

It shows that the heat transfer coefficient of the top layer has the smallest magnitude; those of the bottom layer has the largest magnitude and those of the middle layer is just in between.

The reason for these changes is probably due to the interstitial space effect of the movement of particle, and the temperature difference of different layers of tube bundle and its depth, for heat transfer is also joining the effect.

The experiment results are useful, as the tube bundle is beneficial to be used in preparing carbon for industrial applications, due to the direct touch between solid particles and fluid. The benefit of the increased heating surface would be compensated by the decrease of heat transfer and the drop of temperature difference between the bed and immersed heating surface. Therefore the optimum design for these two compensating factors is interesting and important.

**Proposed correlation:**

The Vreedenberge's correlation of heat transfer between the bed and immersed single tube is popularly known [11], afterwards it was modified by Andeen and Clicksman to consider the porosity and the volumetric heat capacity is added by Grewal and Saxena [12]. All of these had been verified by the experiment data. It is found that the agreement is considered to be good between these three correlations, the maximum discrepancy is less than ± 20 %.

The experimental research results and the (Least Square Method) are used to obtain correlation, for single tube, which is:

$$Nu = 610\left(\frac{GD_t\rho_s}{\rho_f\mu}.\frac{\mu^2}{d_\rho^3\rho._s g}\right)^{0.22}\Pr^{0.3} \qquad (11)$$

Under the experiment condition, the standard error of estimate is ± 14 %, and the correlation factor is 0.9969 for single tube. In the case of tube bundle, the correlations of Andeen and Clicksman are chosen for verifying [12], it was found that the standard error of estimate is ± 25 %, and the correlation factor is 0.9680. To follow the correlations of Andeen and Clicksman, it would be expressed as:

$$Nu = 350(1-\varepsilon)\left(\frac{GD_t\rho_s}{\rho_f\mu}.\frac{\mu^2}{d_\rho^3\rho_s g}\right) \qquad (12)$$

## Conclusions:

1. The overall heat transfer coefficient of tube bundle is lower than the single tube under the same condition.
2. Bundle of tube has been used to consider the optimization for increasing heating surface.
3. The heat transfer coefficient of the top layer of tubes has lowest magnitude than the middle and bottom layers, and the middle layer is just in between.
4. For heat transfer coefficient calculations, available correlation for engineering applications proposed.
5. Acceptability of the heat transfer coefficient correlation term in equation (11) under the experiment condition doesn't increase up to the standard error of estimate is ± 14 %. In the case of single tube and for the tube bundle don't increase up to the standard error of estimate is ± 25%.
6. Based on the review and the efforts done, it can be expected that more ideal correlations combining the theory and practice are forthcoming.

## References:


1. Lev,S.T., Ramakrishna,R., Philip,S., “Dynamics of a Shallow Fluidized Bed”, The American Physical Society, vol.60,No. 6, pp.7126-7130,1999.
2. Al-Ali,B.,M., “Fin Spacing Interaction in Fluidized Bed Heat Exchangers”, Ph.D.Thesis, University Aston in Birmingham, 1976.
3. Jeevan,J.,Arturo,M., “Heat transfer Measurement in Fluidized Bed Combustion Reactor”, Laboratory Notes, Division of Heat and Power Technology , Royal Institute Of Technology , Stockholm,2002.
4. Mcgaw,D.R., “Heat Transfer in Shallow Cross Flow Fluidized bed Heat Exchanger –I.A Generalized Theory”, Int.J. Heat Mass Transfer, Vol.19, pp.657-663.Pergman Press, 1976.
5. Al-Busoul, M., Abu-Zaid, M., “Predication of Heat Transfer Coefficient Between Immersed Surface and Fluidized Beds”, Int.Comm. Heat Mass Transfer, Vol.27, No.4.pp.549-558, 2000.
6. Yasuo, K., Hiroshi, I., Kiyoshi, T., “Fluidization and Heat Transfer Characteristics Around a Horizontal Heated Circular Cylinder Immersed in Gas Fluidized Bed”, Int.J. Heat Mass Transfer, Vol.31, No.2, pp. 349-358,1988.
7. Saxena,S.C.,Gerwal,N.S.,Gabor,S.S.,Zabrodsky,Galershtein,D.M., “Heat Transfer Between a Gas Fluidized Bed and Immersed Tubes, Advances in Heat Transfer”, ( Edited by Irvine,T.F.,Hartentt,J.P.),Vol.14,pp.149-247,Academic Press, New York.,1979.
8. Al-Busoul,M., “Local Heat Transfer Coefficient Around a Horizontal Heat Tube Immersed in a Gas Fluidized Bed”, Int.J. Heat Mass Transfer, 2002.
9. Wen, C.V., Yu, Y.H., “A Generalized Method For Predicting The Minimum Fluidization Velocity”, AICHE, J., 12 (3), pp.610-612, 1966.
10. Al-Busoul, M., “Local Heat Transfer Coefficient in a Fluidized Bed”, Derasat Engineering Sciences, J., Vol.26, No.1, pp.147-154, 1999.
11. Vreedenberg, H.A., “Heat Transfer Between a Fluidized Bed and a Horizontal Tube”, Chem.Eng. Sciences, 9,pp.52-60, 1958.
12. Gerwal, N.S., Saxena, S.C., “Heat Transfer Between a Horizontal Tube and a Gas – Solid Fluidized Bed”, Int.J. Heat Mass Transfer, 23,1505,1980.