Determination of Heat Transfer Coefficients in Air-Solid Fluidized Bed

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Abstract

The heat transfer of gas-solid fluidized bed has been investigated using pre-heating system. A Perspex column of 7cm inner diameter and 1.5 m height is used as a fluidizing column filling with three different sizes of sand with a height of 21 cm as packing.

The experimental work is carried out under superficial air velocities ranged of (0.029-0.365) m/s. The effect of opening area air distributors, particle diameter of sand packing, and different heat fluxes as powers have been investigated. The results show that the minimum fluidizing velocity increases with increasing the particle diameter and opening area of distributor. The gas-particle heat transfer coefficient is found to be increased by increasing superficial velocity, heat flux and opening area of distributor and decreased by particle diameter.

A mathematical correlation of dimensionless groups based on experimental data has been suggested. A comparison between results of present correlation and the speculations of previous ones has been made.

Keywords: fluidized bed, gas-solid system, heat transfer coefficient, Minimum Fluidization Velocity.
1. Introduction

The fluidized bed is commonly considered one of the best contacting methods used in an extensive range of applications in a variety of processing industry, for example in oil refinery plants. There are many dependable operations that make use of this technology, involving (cracking, ore roasting, coal carbonization and gasification, reforming of hydrocarbons, Fischer-Tropsch synthesis, coking, and coating concoction) (1).

Despite numerous studies on heat transfer in fluidized bed have been done, but still there is not enough researches in case of preheating fluidized bed reactors. Number of correlations have been suggested for the over-all and maximum heat transfer coefficient where some of them are listed in Table (1). Kettenring (2) was the first to outlined gas-solid transfer coefficients in case of silica gel and alumina irregular shape particles with a diameter of (1000, 713, 502 and 355 µm), heated and fluidized by air. Walton and coworkers (3) measured the gas to particle heat transfer coefficients in beds of crushed coal (711, 502 and 355 µm) fluidized with air.

While, the gas-to-particle heat transfer in fluidized columns filled with three different solid particles (glass beads (137-454) µm, washed alumina (228-914) µm and Dowex-50 (457-762) µm) under unsteady state heating condition using air and co₂ as a fluidizing gas has been investigated by Wamsley and Johanson (4). Ferron (5) examined gas-to-particle heat transfer in fluidized beds with silica-alumina catalyst (60) µm, using air and water as a fluidizing gas. The above investigators have correlated their experimental data using equation containing only the Reynolds number and Nusselt number. Al-Hattb (6) studied the influence of the particle size and Reynolds number for different values of heat flux on the heat transfer coefficient in gas-solid fluidized bed. The suggested correlation predicts the value of Nusselt number as a function of both Reynolds number and Froud number. All the dimensionless groups are determined based on particle diameter. Recently, Hamzehei and some researchers (7) investigated the heat
transfer and hydrodynamic in gas-solid fluidized column of (300 µm) sand particles experimentally and computationally. A pre-heating system has been used to increase the inlet gas temperature from ambient temperature to 473K. They have showed that the computational model could predict the heat transfer and the hydrodynamic behavior of gas-solid fluidized bed flows with reasonable accuracy.

The present work concerns in experimentally measurement of the overall heat transfer coefficient in gas-solid fluidized bed of preheating system under different working variables of particles diameter of sand, opening areas of distributor, heat fluxes and flow rate. Moreover, semi-empirical correlation of dimensionless groups has been established to predict the heat transfer coefficient of the system and compared with other correlations of well-known investigators.

**Table (1)** The results of some researchers with their correlations and related heat transfer coefficients.

| Researcher                         | Conditions                  | Correlations                                      | h range |
|------------------------------------|-----------------------------|----------------------------------------------------|---------|
| Kettenring et al. (1950)           | Steady state gas fluidized  | \( \text{Nu} = 0.0135 \text{Re}^{1.35} \)         | 3-10    |
| Walton et. al. (1952)              | Steady state gas fluidized  | \( \text{Nu} = 0.0028 \text{Re}^{1.7} (D_c/d_p)^{0.2} \) | 5-35    |
| Wamsley et. al. (1954)             | Unsteady state gas fluidized| \( h = 1270d_p^{1.27} \)                          | 0.07-9  |
| Fritz (1955)                       | Unsteady state gas fluidized| \( h = (G/(116+52.4L_B)) + (0.48/(274+447L+146L_B)^2) \) | 0.0005-0.012 |
| Ferron (1958)                      | Unsteady state gas fluidized| \( h = 0.000038 \text{Re}^{0.5} D_c^{-0.7} (L/D_c)^{-0.7} \) | 0.09-2.8 |
| Heertjes et. al. (1956)            | Steady state gas fluidized  | \( h = 1.31 \text{Re}^{0.76} \)                   | 6-28    |
Frantz (1961) | Steady state gas fluidized | \( \text{Nu} = 0.018 \text{Re}^{1.2} \) | 0.8-2
Sunkoori et. al (1960) | Unsteady state liquid fluidized | \( \text{Nu} = 0.00391 \text{Re}^{2.1} \) | 110-620
Glicksman et. al (1997) | Steady state gas fluidized | \( \text{Nu} = 6.0 + 0.05(\text{RePr}) \) | --
Al-Hattb (2009) | Steady state gas fluidized | \( \text{Nu} = \frac{0.073 \text{ Re}^{0.57} \text{Fr}^{0.48}}{5.23 + 0.0042 \text{Re}} \) | 5-23
Hamzehei et. al (2012) | Unsteady state gas fluidized | simulation | --

2. The Problem of the Study

Comprehension of heat transfer rates between fluid and solid particles is fundamental in such physical processes as particles drying or quenching of hot fluids likewise in chemical reactions involving large heat effects that occur on and within thefluidized catalyst particles.

In spite of numerous studies on heat transfer in fluidized bed have been done, but still there is not enough researches in case of gas-particles heat transfer coefficient measurements in preheating fluidized bed reactors. This because the perplexity of the contacting pattern between particles and gas, the tactic is to detect average heat transfer coefficients for the whole bed rather than local coefficients. This can be done by gross measurements of temperature of the entering and leaving gas and the bed temperature. In order to get an averaged heat transfer coefficient over the whole bed, one has to suppose a flow model for the gas and the particles (8).

On a synoptic scale the temperature of the solid particles in the bed is roughly regular, forasmuch the gas temperature varies very rapidly in a very short section of the bed close to the distribution grid and very slightly beyond it. Temperature measurements taken in beds with thermocouples were interpreted very differently. Some investigators interpret a bare thermocouple reading as the gas temperature while others as that of the solids. Some use thermocouples protected by a mesh or a cloth and assume the
measurement to give the gas temperature, others use suction thermocouple measurements to obtain the gas temperature. Direct measurements of solid particle temperature are even more difficult and unreliable (8).

Out of the three distinct mechanisms of heat transfer in fluidized beds namely (i) fluid-to-particle, (ii) particle-to-particle and (iii) surface-to-bed, only the second and the third mechanisms have been widely studied and reported earlier. No such studies has been made for the first mechanism, So an effort is made here to give a concise report of the gas-solids heat transfer measurements in preheating fluidized bed reactor. The report covers hydrodynamics and heat transfer coefficient calculations, besides the effect of various parameters on the heat transfer coefficient and minimum fluidizing velocity.

3. Experimental Setup

A bench scale experimental setup for studying heat transfer in a gas-solid fluidized bed was designed and fabricated as shown schematically and photographically in Figures (1) and (2) respectively. The setup consists of a Perspex column with a height of 150 cm and a diameter of 7 cm. The air was injected through a perforated plate with three different open area percentages (1, 1.64 and 2.5 %) and an orifice diameter of 2 mm. Sand particles with a diameters of (269, 379 and 502) µm and a density of 2620 kg/m$^3$ were fluidized with air at various thermal conditions. Typically, the static bed height was 21cm with a solid volume fraction of 0.6. A roots-type blower supplied the fluidizing gas. A pressure-reducing valve was installed to avoid pressure oscillations and achieve a steady gas flow. The airflow rate was measured using an air flow meter. Initial solid particle temperature was 300K. An electrical heater with a power of 1440 W located in the plenum was used to increase the inlet gas temperature. Moreover, pressure tapes with glycerin manometer are connected throughout the column for pressure drop measurements.

Four thermocouples (Type k) were used in the experimental setup. Two thermocouples were installed in the fluidized bed column at (10 and 20) cm above the distributor to measure the average of bed temperature. Another thermocouple was
installed at 40 cm above the distributor to measure the outlet gas temperature. Also, a one thermocouple was used in plenum to measure the inlet gas temperature. Figure 1 shows the locations of the pressure tapes and thermocouples. The range of experimental conditions illustrated in Table (2).

Figure (1): Schematic Diagram of the Experimental Apparatus. 1. Fluidization column, 2. Bed materials, 3. Distributor, 4. Plenum 5. Heater, 6. Variac transformer, 7. Air flowmeter, 8. Ball valve, 9. Blower, 10. Power supply, 11. Heat sensors, 12. Thermometer and 13. U-tube manometer.
Figure (2): The experimental rig photograph.

Table (2): Experimental condition

|   |   |   |
|---|---|---|
| 1. | Distributor plate |   |
|   | Percent perforated area (%) | 1 | 1.64 | 2.5 |
|   | Hole diameter (mm) | 2 | 2 | 2 |
|   | Pitch (mm) | 17 | 13 | 11 |
|   | No. of holes | 13 | 21 | 30 |
| 2. | Sand particles size |   |
|   | Mean diameter, dp (µm) | 269 | 379 | 502 |
|   | Particle density $\rho_p$ (kg/m³) | 2620 | 2602 | 2587 |
| 3. | Bed height |   |
|   | Column inner diameter (cm) | $D_c = 7$ |
|   | Packed bed height (cm) | $H_o = 21$ |
| 4. | Fluidizing velocity |   |
|   | Hydrodynamic measurements range (m/s) | 0.029 - 0.365 |
4. Experiment procedure

i. A specific weight of sand has been poured into the test column from the top to the desired static bed height above the distributor plate which gave loosely fixed bed depth of (21cm).

ii. The blower is turned on.

iii. The ball valve must be adjusted to provide the desired fluidizing velocity.

iv. In order to get the required amount of heat flux. The variac (voltage regulator device) is adjusted to a desired value.

v. The recording of the readings of the bed temperatures was started just when the surface temperature of heater became constant, with time interval of (0.25 minute) between each reading.

vi. When steady state was demonstrated in heat transfer experiments when the average temperature variation of the bed (1-3) K / min.

vii. The globe valve is changed to get a new fluidizing velocity and then the measuring has to be started again. The range of fluidizing velocity is from a fixed bed to a fluidizing velocity equal to three times of the minimum fluidizing velocity.

viii. The heat transfer coefficient is calculated from the energy balance to the system (Heat lost by the gas equals to the heat gained by the bed or solids) as shown in equation (8);

\[ h_b = \frac{m_f c_p_f (T_f - T_a)}{a_s (T_b - T_{gb})} \]  

(1)

Where:

- \( m_f \) and \( c_p_f \) are the mass flow rate and the specific heat capacity of the air respectively;
- \( a_s \): surface area for single particle;
- \( T_b \): average bed temperature (i.e. the average of two readings);
- \( T_{gb} \): the average bulk gas temperature, (i.e. the average of out and in of air readings).

ix. The pressure drop has been recorded at each run.
x. The above procedure was repeated for each distributor type, particle type, and different heat flux.

Whereas, the physical properties of air were used at the average bulk temperature of air.

5. Results and discussion

Hydrodynamic and heat transfer results under several parameters have been demonstrated as below:

5.1 The Minimum Fluidizing Velocity

Minimum fluidizing velocity is calculated theoretically by the equation of Carman-Kozeny (9), as shown in eq. (2).

\[ u_{mf} = 0.0055 \frac{d_e^2(\rho_g-\rho_e)g}{\mu} \left( \frac{e^3}{1-e} \right) \]  

(2)

The experimental values of minimum fluidized velocity have been achieved from the chart illustrating the pressure drop against gas velocity. These values have been pointed as intersection between the line for the fixed bed (the increasing line before fluidization state) and the horizontal line which representing the constant line of pressure drop with further increasing of gas velocity after the fluidization state. Table (3) shows comparison between of experimental and calculated values of minimum fluidizing velocity.

Table (3): The values of experimental and calculated minimum fluidization velocity.

| Particle diameter \( \mu m \) | Minimum fluidization velocity | Opening area | Experimental (m/sec.) | Calculated (m/sec.) | % Error |
|-------------------------------|-------------------------------|--------------|----------------------|---------------------|---------|
|                               |                               |              |                      |                     |         |
|                               |                               | 269          | 1%                   | 0.058483            | 0.0597585| 2.10    |
|                               |                               |              | 1.64%                | 0.073099            | 0.0597585| 22.3    |
|                               |                               |              | 2.5%                 | 0.087719            | 0.0597585| 45.6    |
|                               |                               | 379          | 1%                   | 0.109851            | 0.117485 | 6.40    |
|                               |                               |              | 1.64%                | 0.124269            | 0.117485 | 5.70    |
|                               |                               |              | 2.5%                 | 0.131579            | 0.117485 | 11.9    |
|                               |                               | 502          | 1%                   | 0.197342            | 0.196444 | 0.45    |
|                               |                               |              | 1.64%                | 0.204678            | 0.196444 | 4.19    |
|                               |                               |              | 2.5%                 | 0.248538            | 0.196444 | 26.5    |
It is clearly known that these minimum velocities have been affected by several variables as follow:

i. **The effect of distributor opening area:** Figure (3) shows the pressure drop versus superficial air velocity for different opening areas of distributor and dp=379 µm. Moreover, it is clear that increasing the distributor opening area causes an appreciable decrease in pressure drop across the distributor plate. Figure (3) also indicates that for a higher opening area of distributor, the asymptotic value of \( u_{mf} \) is higher. Figure (3) reveals 20% increase in \( u_{mf} \) when the opening area increases from 1% to 2.5%. It is clear that increasing the distributor opening area causes an appreciable decrease in \( u_{mf} \). A possible explanation for this might be the large opening area has a low resistance against the air and subsequently low pressure drop across the plate. These results are in good agreement with literatures (10).

![Figure (3): The effect of the opening area of perforated plate on the minimum fluidization velocity for sand particle (dp=379µm)](image)

ii. **The effect of particle diameter:** Figure (4) shows the trend of pressure drop with superficial air velocity for different particles diameter. Figure 4 shows that for a higher particle diameter of sand, the value of \( u_{mf} \) is higher. It is clear that the \( u_{mf} \) increases distinguishably, as the particles diameter increases. Such a behavior occurs
because the larger and heavier particles require higher air flow rates to be fluidized, because the drag and gravitational forces acting on the bed of particles must be balanced. In the fixed bed region, the larger particles provide more permeable beds, favoring the flow of air through the bed and reducing the pressure drop. This finding is completely agreed with the results gained by several researches (7,11).

![Graph showing the effect of the particle diameter on the minimum fluidization velocity (φ=1%)](image)

**Figure (4):** The effect of the particles diameter on the minimum fluidization velocity (φ=1%)

### 5.2 The heat transfer coefficient

This section presents the heat transfer coefficient calculated via Equation (1) under effect of different operating conditions.

**i. Effect of gas velocity:** Figure (5) shows the heat transfer coefficient versus superficial air velocity for different opening areas of distributor at heat flux of 264 W and dp=502 µm. Figure 5 indicates that increasing the air velocity from 0.058 to 0.301 causes a significant rapid rise in \( h_p \), thus heat transfer coefficient increases rapidly over a comparatively narrow velocity range (\( u_o > u_{mf} \)). A continuous increase in the heat transfer coefficient is still observed when the air velocity continues to increase, but the rate of increment becomes gradually smaller, until reaches the maximum value. The reason behind that is the perpendicular motion of
the particles along the bed, the scrubbing action in the laminar film starts decreasing considerably thermal impedance (1). Moreover, another reason may be considered as the temperature distribution along the fluidization column increases (6). This is due to the increase in the solid particles enthalpy with increasing the fluidized bed velocity, because the quantity of air entering to the solid particles increases. The additional increase in gas velocity produces a lessening in particle occupants over the bed which reduces the heat transfer coefficient. In the same manner, this behavior has appeared for each heat flux. In order to understand the reduction in particles population, Figure 6 shows the parallel changes in bed expansion behavior that are noticed with increasing superficial gas velocity. The figure reveals that increasing air velocity causes a decrease in solid volume friction. A decreasing in solid volume friction causes an increase in $h_p$, until reaches a maximum value when solid volume friction equals 0.45. Then any additional increase in air velocity reduces $h_p$. The solid volume friction calculated via the following equation (12):

$$e = 1 - \frac{W_s}{Ah(\rho_s - \rho_g)}$$

(3)
Figure (5): Heat transfer coefficient as a function of superficial velocity for sand particle (269μm) at q=220 w and φ=2.5%

Figure (6): Variation of transfer coefficient and void fraction as a function of superficial velocity for sand particle (269μm) at q=220 w and φ=1%

ii. The effect of particles diameter: Figure (7) shows the heat transfer coefficient versus superficial air velocity for different particles diameter at heat flux of 220 W and φ=1%. Figure (7) reveals that the curve of heat transfer coefficient versus the
superficial air velocity $dp = 269 \, \mu m$ is higher than the curve achieved for $dp = 369$ and $502 \, \mu m$ at the same range of gas flow rate, heat flux and distributor type. These findings are in good agreement with other researchers found in the literature (13). In accord with Farbar and Depew (14) a raised thermal resistance corresponding with the particle surface area result in a reduction in the heat transfer coefficient with particle diameter. The reason of that is the temperature increases as the mean solid particles diameter decreases, where the finer particles can cause higher heat transfer coefficient due to the small conductive resistance of small particles. Moreover the smaller particles can rise the effect heat transfer area covered by itself (6).

![Figure (7)](image_url)

**Figure (7):** Effect of particles diameter on the gas-particles heat transfer coefficient at ($q=220 \, W$ and $\varphi=1\%$)

**iii. Effect of heat flux:** Figure (8) shows the variation of gas-to-particle heat transfer coefficient as a function of gas superficial velocity for different power supplies (220, 264 and 330) W. Figure 3 indicates that for a higher heat flux, the values of $h_p$ are higher, because increasing heater’s power causes an increase in gas temperature. The obvious changes are in the thermal conductivity of fluidizing
air which can be expected to affect the gas-to-particles heat transfer coefficient. The thermal conductivity of fluidizing air increases with any increase in temperature. This increase results in an increase in the coefficient due to a decrease in the chief resistance to heat flow (15).

Figure (8): Variation of heat transfer coefficient as a function of superficial velocity for different power supplied for (dp=269μm of sand and φ=1%)

iv. Effect of air distributors: Figure (9) shows the effect of opening area percentage of distributor on the trend of $h_p$ with superficial air velocity for dp = 269 μm of sand and heat flux of 264 W. The figure reveals that increasing $\phi$ causes an increase in $h_p$. The increase in $h_p$ with $\phi$ is ascribed to the enhancing of the particle mixing and the amount of heat that come up with the gas, also the good distribution of gas through the bed as increasing the percentage of opening area of the distributor. Which is approved by other workers (15,16).
5.3 Correlation of data

In this study on heat transfer between a gas and particles, it is well proven that the Reynold number and Prandtl Number are a good characteristic dimensionless group for (Nu); to summarize our experimental and using the method of Quasi – Newton on (STATISTICA v10.1 enterprise). The following correlation is proposed:

\[ \text{Nu}_p = 0.0268(\text{Re}_p)^{1.26}(\text{Pr})^{0.48} \left( \frac{k_s}{k_g} \right)^{-0.0375} \left( \frac{1 - e}{e} \right)^{-0.173} \]  

Figure 10 shows the comparison between the experimental data of Nu and the correlated values according to equation 4. The correlation shows a good relation between the data, where the correlation coefficient (R) = 0.927388. Figure 11 shows the comparison between the present correlation of Nu and the previous ones. The new correlation is close to the Walton correlation rather than the other correlations.
Figure (10): Correlated verses experimental data of Nusselt number (Nu)

Figure (11): comparison the present work correlation with others (Nu)

6. Conclusion

From the analysis of the current experimental hydrodynamic and heat transfer data in a bed of different particle types and for different heat fluxes and different opening area
of distributors, some general conclusions can be drawn: The value of minimum fluidizing velocity increases with an increase in the value of particles diameter and an increase in the opening area friction of distributor. The heat transfer coefficient increases with an increase in the thermal conductivity of air and, with a given bed and different particles diameter, the heat transfer coefficient increases with an increase in superficial gas velocity and attain a maximum value at velocities slightly above the minimum fluidization velocity. The gas to particle heat transfer coefficient increases with an increase in the heat flux. The value of gas-particles heat transfer coefficient increases with an increase in opening area friction of distributor.

**Nomenclature**

| Symbol | Definition                                      | Unit  |
|--------|-----------------------------------------------|-------|
| a_s    | Surface area of single particle               | m²    |
| d_c    | Column diameter                               | m     |
| d_b    | Bed diameter                                  | m     |
| d_p    | Particle diameter                             | m     |
| g      | Acceleration of gravity                       | m/s²  |
| h_o    | Packed bed height                            | m     |
| h_mf   | Bed height at minimum fluidization            | m     |
| h_p    | Gas to particle heat transfer coefficient     | W/m².K|
| T_B    | Average bulk of gas temperature              | K     |
| T_b    | Average bed temperature                       | K     |
| T_hs   | Heater surface temperature                    | K     |
| T_i    | Inlet gas temperature                         | K     |
| T_o    | Outlet gas temperature                        | K     |
| u_mf   | Minimum fluidized bed                         | m/s   |
| k_g    | Gas thermal conductivity                      | W/m.K |
| k_s    | Solid thermal conductivity                    | W/m.K |
| W_s    | Weight of sand                                | Kg    |
| e      | Void friction                                 | -     |
| pr     | Prandtl number                                | -     |
| Re     | Reynold number                                | -     |
| φ      | Opening area percentage                       | -     |
| \(u_{mf}\) | Minimum fluidized velocity | m/s |
| --- | --- | --- |
| \(\rho\) | density | m\(^3\)/kg |

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