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Abstract

The present work aims to study the fluidization and heat transfer characteristics around a single horizontal heated tube, in-line tubes arrangement and staggered tubes arrangement immersed in a gas fluidized bed. The experiments were carried out on a square test section column (16 cm x 16 cm x 95 cm) from a steel sheet of 3 mm thickness. Each arrangement of the tubes is installed in a separate test section. The plane tube bank for both in-line and staggered tubes arrangement consists of 5 rows and four tubes per row in equal vertical and horizontal pitch arrangement. The experimental setup has a transition channel with the same square section and 30 cm height. The other side of the transition channel has one hole of 5 mm inside diameter and 30 mm long to measure the pressure before the distributor plate. The bed material used is sand particles of different sizes (1400 μm, 1600 μm, and 1850 μm). For the tested arrangements, the results show that the average heat transfer coefficient increases with the increase of the fluidizing velocity and with the decrease in particle diameter. Also, the comparison between the in-line and staggered tube arrangement showed that, the average heat transfer coefficient in case of staggered is higher than that of the in-line tube arrangement. Moreover, empirical correlations for the average Nusselt number of the tested arrangements using the experimental data are presented.

KEYWORDS: heat transfer, fluidized bed, bundle of tubes, vertical channel

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1. INTRODUCTION

Fluidized beds are widely used in industry for mixing solid particles with gases or liquids in most industrial applications. The technique of fluidization is related to operations, and it is used in many industrial applications such as mining, metallurgical engineering, liquid settling, sedimentation and density classification. The major successful application of gas fluidized bed techniques is shown in the catalytic cracking process. Fluidized beds are used in industry applications, especially the heat exchangers that use this technique. The chemical and oil industries have concentrated largely on applications of gas-fluidized systems which exploit the advantage of the technique for handling solids in solid / fluid contacting operations. Although, the cracking process, the heat transfer properties of the fluidized systems have often been essential to successful operation, find application in fields as adverse as the heat treatment of metals and power station boilers [1&2].

Vreedenburg [3] studied experimentally the heat transfer between a horizontal stainless steel tube and air fluidized bed of 0.565 m diameter. The tube was mounted at 0.85 m above the orifice type distributor plate and the static head height was 1.2 m. The effects of bed temperature, air mass velocity, particle shape (round and sharp), mean particle diameter, particle density, and the tube diameter on the heat transfer coefficient were studied. The results showed that, the average heat transfer coefficient at the tube surface increases with increasing of the air mass fluidizing velocity up to certain value, and decreasing with tube diameter. The experimental results of Nu_T for fine and coarser sand particles were correlated. Andean and Glicksman [4] predicted the values of Nu_T within ±25% for the results of Vreedenburg [3] for fine silicon sand (d_p =167 μm). The results showed that the heat transfer coefficient is decreased with the increase in distance from the distributor plate. For small particles in the bed at low temperature, the contributions of gas convection and radiation are negligible with respect to particle convection component [5]. Grewal and Saxena [6] studied experimentally the average heat transfer coefficient between an electrically heated horizontal tube and air-fluidized beds of glass beads, dolomite, sand, silicon carbide, and alumina particles. The effects of particle size (d_p =178 to 504 μm), particle density (ρ_s=2660 to 4450 kg/m^3), specific heat (c_{ps} =0.44 to 0.929 J/kg K), mass fluidizing velocity (G = 0.1 to 0.7 kg/m^2 s), tube diameter (D_T=0.0127 to 0.0286 m), tube material (bronze and copper), electrical power input (Q=113.5 to 31505 W) and distributor plate were investigated. The experiments were carried out in (305×305 mm) square bed and with three different distributor plates, two plates of perforated type with different open areas on bubble cap type. They compared the experimental results, which covered wide range of operating conditions with the published correlations. Gelperin et al. [7], Gentti et al. [8], Saxena et al. [9], and

Borodulya et al. [10] reported that the average heat transfer coefficient was independent of static bed height. Ainshtein [11] proposed a study on the fluidized bed with different distances of distributor plate from the horizontal tube. The results indicated that the heat transfer coefficient is decreased with the increase in the height of the tube above the bed.

For fluidized bed systems, the particle size is the most influential parameter in heat transfer coefficient. Mclaren and Williams [12] studied experimentally the effect of particle size in fluidized bed heat transfer. They indicated that a considerable effect of d_p on h_w. Cherrington et al. [13] investigated experimentally the heat transfer coefficient for an immersed tube and found that it varied as $d_p^{0.36}$. Golan et al. [14] Presented the experiments with limestone of different sizes distribution concluded that d_p is important only for a bed of narrow size distributor. In the mixture of wide blend, smaller particles control the heat transfer and the d_p does not significantly affect the heat transfer. Zakkay et al. [15] found in the study of coal fired tests that the total heat transfer coefficient, which the sum of the individual components including the bubble fraction effect was decreased with the increasing in particle size. Sastri and Rao [16] had reported that the presence of 7% over size had no effect on the value of h_w. Andersson [17] studied experimentally the heat transfer coefficient for a circulating fluidized bed (CFB) boiler by using three particle diameters of 220, 340, and 440 μm silica sand. It was found that the change from 440 to 220 μm at constant fluidization velocity led to a considerable increase in heat transfer. Garic et al. [18] studied experimentally the wall-to-bed heat transfer in vertical hydraulic transport and in particulate fluidized beds. The experiments were performed by spherical glass particles of 800-2980 μm in diameter with water in a 25.4 mm inner diameter copper tube equipped with a steam jacket. They found that the heat transfer coefficient is generally higher while the flow is in fluidized bed. Masoumifard et al. [19] studied experimentally the heat transfer between a horizontal tube and gas-solid fluidized bed in order to verify the influence of the axial position and particle diameter on the heat transfer coefficient from a small horizontal tube (D_T =8 mm) immersed in the fluidized bed. The solid sand particles used were 280, 490, and 750 μm in diameter, fluidized by air. The heat transfer coefficient was found to increase by decreasing the particle size.

The effect of tube arrangement in fluidized bed systems on heat transfer were investigated by many investigators. From the experimental work of Gelperin et al. [20], Gelperin et al. [21], and Kofman et al. [22], the following conclusions were obtained: The average heat transfer coefficient of vertical tubes was about 5 to 15% higher than that of horizontal tubes. For vertical tubes inside the bed, the gas distribution within the bed becomes more uniform with increase in the number of tubes in the bundle. Also, for more closely packed bundles, the average heat transfer coefficient can be reduced by about 35 to 50%. Bartel and Genetti

[23] studied experimentally the rate of heat transfer from a bundle of electrically heated horizontal tubes in staggered array in an air fluidized bed. It was found that the total heat transfer coefficient for the bundle increased when the increase in tube spacing from 11 mm to 38 mm. With further increase to 119 mm, there was no additional increase in the total heat transfer coefficient for all particle sizes. Newby et al. [24] achieved the following conclusions: The average heat transfer coefficient for single, central subject tube locations considered changes very little for 15.9 mm diameter tube, but caused a significant increase by (10 to 20%) in the heat transfer coefficients for 31.8 mm diameter heaters. The totally fluidized bed pressure drop and bed expansion are not significantly influenced by the pressure of the compact of the tube bundle. Gelperin and Einstein [25] and Nayak and Rajarao [26] studied the heat transfer from bundle of horizontal tubes in-line and staggered arrays. Bed particle size, fluidizing velocity, and relative horizontal and vertical pitches were varied to study their effects on heat transfer coefficient. For in-line and staggered arrangements, it was observed that the total heat transfer coefficient was reduced with the decrease in horizontal pitch. Gelpern et al. [27] reported the experimental work that conducted by Korotyanskaya and Markevich [28], and Gelperin and Einstein [25] on two horizontal rows of tubes arranged one above the other with various angles with the center line of the rows. It was found that the heat transfer coefficient was increased from 3 to 5% when the very closely arranged tube rows were changed gradually from an in-line to crossed arrangement.

The above literature review shows that, the heat transfer between fluidized bed and a bundle of tubes is not extensively studied. So, the present work aims to study the fluidization and heat transfer characteristics around a single horizontal heated tube, in-line tubes arrangement and staggered tubes arrangement immersed in a gas fluidized bed.

2. EXPERIMENTAL SETUP

The experimental investigation is carried out in an open circuit specially built and designed for the present research work. The experimental setup is shown in Fig. (1). This figure illustrates a general view of the individual components of the testing installation creating a bed column of square size (16 x 16 cm). The bed is supplied with air from a medium pressure of 110 psi reciprocating compressor. The air entering the base of the transition channel size (16 x 16 x 30 cm) that placed before the test section through a valve which controlled the air flow rate. The transition channel makes the flow uniform at the inlet of the distributor plate. Figure (1), shows schematically the experimental setup with its component which basically consists from a compressor, transition channel, distributor plate, test section, and tested tube bank.

The transition channel is made from 3 mm mild steel plate having one flange on the lower end. The square section of the transition channel is (16 x 16 cm) and the height is 30 cm. This section is provided with 3 holes, each at a height of 40 mm from the distributor plate. The distance between each two adjacent holes is 55 mm. The opening holes are made for inserting probes for measuring the velocity before the distributor plate. The other side of the transition channel has one hole of 5 mm inside diameter and 30 mm long to measure the pressure before the distributor plate. The distributor plate used in the experiments is made from stainless steel, mesh 40 i.e. (40 holes in the square inch), as shown in Fig. (2). It plays a key role in maintaining a high degree of quality of fluidization necessary to make the system effective.

The square test section column (16 x 16 x 95 cm) is designed and fabricated to facilitate the performance of experiments to basic dynamic and it is tall enough to keep deep bed. The section is constructed from steel sheet of 3 mm thickness for three sides while the fourth one is made of 6 mm sheet from heat glass. Figure (3) shows the schematic diagram of the test section. The glass sheet in the test section allows visual observation of the bed when fluidized.

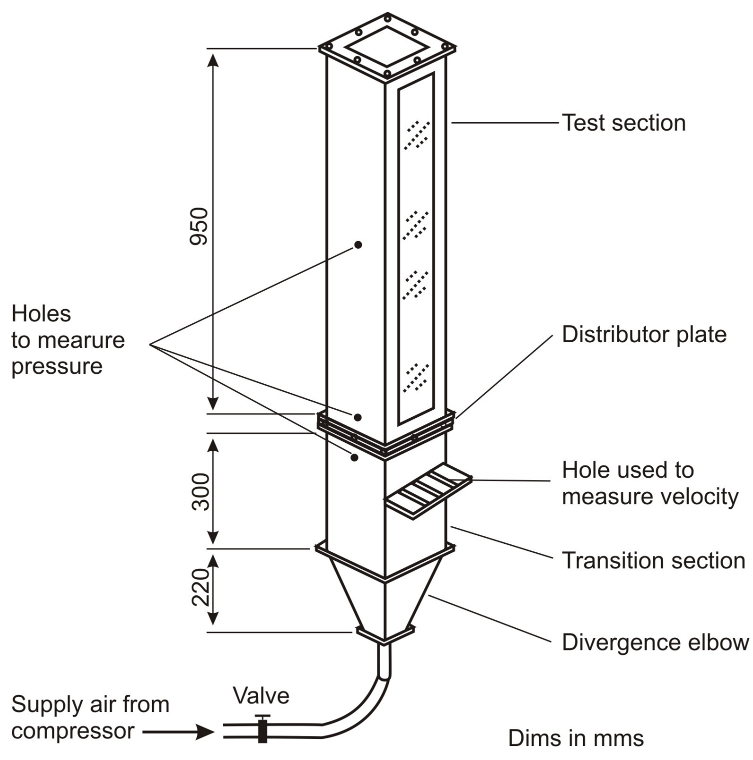

Fig. (1) Experimental set up

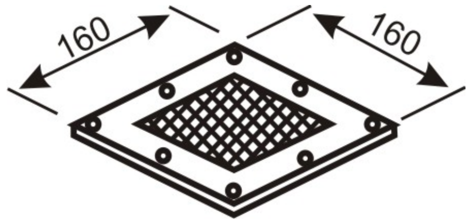

Fig. (2) The distributor plate

A cartridge heater is used to heat the stainless steel tested tube with inner diameter of 16 mm and thickness of 3 mm. Both sides of the heated tube are insulated to minimize the axial heat loss. The tested tube contains four thermocouples of K-type (Chromel-Alumel), as shown in Fig. (4). Each arrangement in a separator test section has the same dimensions and the position of the tubes in each form is horizontally in the same direction of the heater. Only one heater is used to transfer the heat to stainless steel tested tube of 22 mm outer diameter. The other tubes are dummy made of wood with 22 mm outer diameter, where the plan of tube bank for both arrangements consisting of 5 rows and 4 tubes per row in equal vertical and horizontal pitch arrangement, as shown in Fig. (5). Each of the longitudinal and transverse pitches are equal to 40 mm. The arrangement scheme of tubes were half tubes beside the walls which minimize the by pass flow near the walls of the bed.

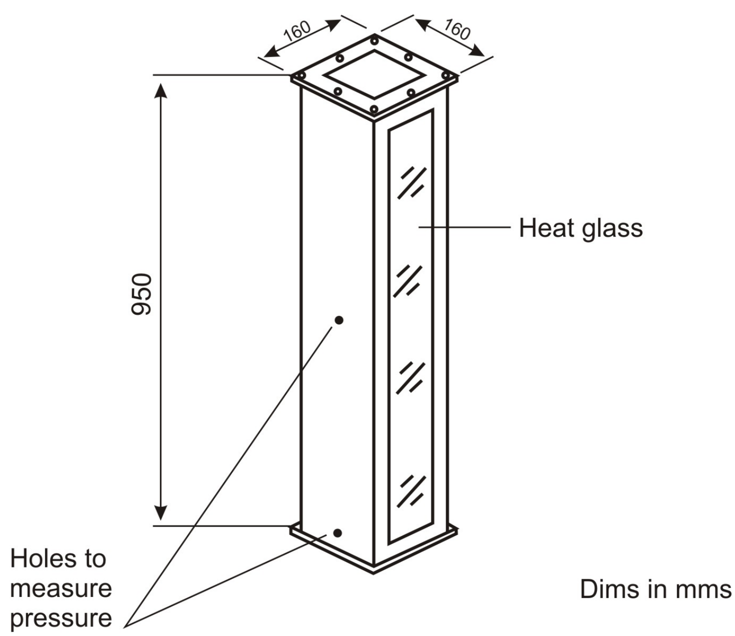

Fig. (3) Schematic diagram of the test section

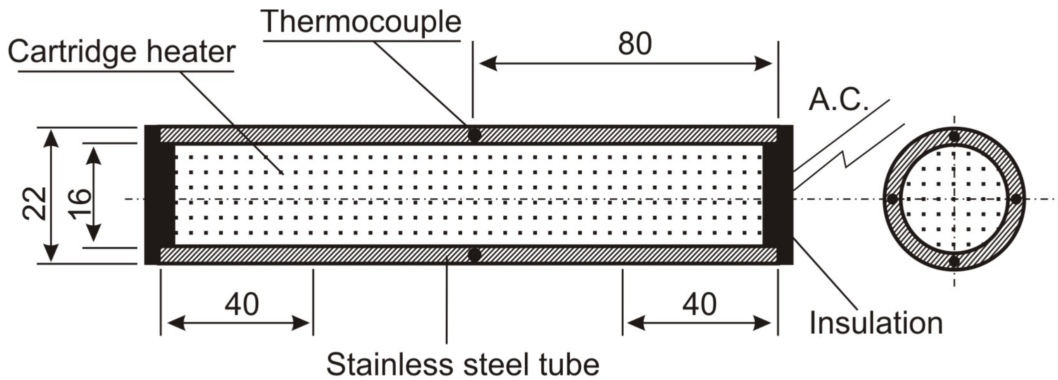

Fig. (4) Detailed of tested heat transfer tube

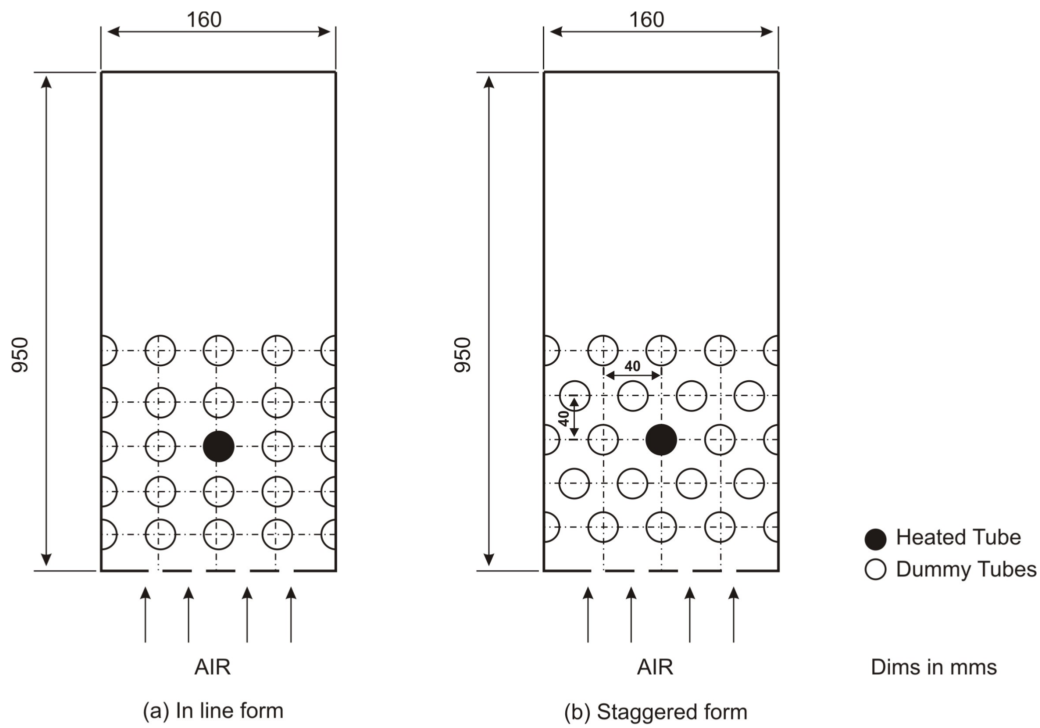

Fig. (5) Tubes arrangement

Pressure in the bed is measured by using manometer, through holes on the side wall of the test section. The tested tube wall temperature and the bed temperature are measured using K-type thermocouples. Four thermocouples are fixed on the surface of the stainless steel tube to measure the temperature of the heater, where each point from the remaining three points is located at 90 degree from the other, as shown in Fig. (4). Eight thermocouples are located and fixed on stainless steel wire inside the test section to measure the bed temperatures. A digital thermometer with a sensitivity of ± 0.1 °C is employed for all the temperature measurements. Air flow velocity is measured by using standard pitot tube type (P.T.) inserted inside the transition channel to measure the value of the dynamic pressure through three holes in the side of the transition channel at the same level from the distributor plate, and the static pressure is measured from a hole at the surface of the transition channel. Heat transfer tube is internally heated by cartridge heater of 150 Ω total resistance. The power consumed in the heater coil is measured with digital multimeter to measure the voltage and the resistance. The particles are found by sieve analysis using group of sieves to determine mean particle diameter. The particles are considered to be spherical in shape.

3. THE PROCEDURES AND METHOD OF CALCULATIONS

The system is started in the following manner, at known bed height and particle size: 1-Adjust the level of water inside U tubes manometer. 2- Open the tap of air flow gradually up, turn on the heater. 3- Increase the velocity of flow to minimum fluidizing velocity.

The fluidizing velocity is determined by conventional method measuring the bed pressure drop as a function of the air velocity before the distributor plate (U). The tube wall temperature and bed temperatures are measured with the help of thermocouples and a digital thermometer (temperature recorder). The steady state condition is reached after 2.5 - 3 hours, and the wall temperature is nearly constant through the digital thermometer. The average heat transfer coefficient, h_w, for the tested heated tube is calculated as:

$$h_w = Q / (A(T_w - T_b)) \tag{1}$$

where Q : Electrical power supplied to heater, W

A : Surface area of heated tube, m^2

T_w : average surface temperature of heated tube, K

T_b : average bed temperature, K

The Reynolds number based on the tube diameter, Re_T, and based on the particle diameter, Re_p, are calculated respectively as:

$$Re_T = \rho_a U^* D_T / \mu_a \tag{2}$$

$$Re_p = \rho_a U^* d_p / \mu_a \tag{3}$$

where ρ_a: The density of air, kg / m^3

U^* : The average velocity of air inside the test section, m/s
D_T : The outside diameter of heated tube, m
d_p : Mean particle diameter, m
μ_a : Dynamic molecular viscosity of air, N. s / m^2

Also, the average Nusselt number based on the tube diameter, Nu_T, and based on the particle diameter, Nu_p, can be estimated respectively as:

$$Nu_T = h_w D_T / k_a \quad (4)$$

$$Nu_p = h_w d_p / k_a \quad (5)$$

The uncertainty of heat transfer coefficient or Nusselt number is estimated to be about ± 2%. Also, uncertainty of about ± 1% is found in measuring the flow velocity or reported Reynolds number.

4. RESULTS AND DISCUSSION

4.1. Minimum Fluidization Velocity

A series of experiments is carried out for fluidized bed having single tube. The other series of experiments is conducted to investigate the effect of in-line and staggered tube banks on the pressure drop across the fluidized bed. The bed height (H) is 17cm for all cases.

The variation of the pressure drop across the fluidized bed having single tube with air velocity for different sand particle diameters of d_p =1400 μm, 1600 μm, 1850 μm, is indicated in Figs. (6), (7), and (8), respectively. Also, the influence of air velocity on the pressure drop across the bed with sand particles of d_p = 1600 μm and bed height of 17cm for in-line and staggered tubes arrangement is shown in Figs (9) and (10), respectively. From all these figures, it is observed that the pressure drop across the bed increases with the increase of air velocity up to a certain value. This value depends on both the particle diameter and the tube arrangement. With further increase in air velocity, the pressure drop is still constant. However, the minimum fluidization velocity occurs at that particular velocity. It is found that the increase of bed particle diameter increases the minimum fluidized bed velocity at the same static head. Also, these figures show that the minimum fluidized bed velocity in staggered arrangement is greater than that in-line arrangement. Moreover, fluidization occurs as: by increasing the air flowing velocity gradually, some particles in the bed will start to move in an oscillating condition but within a very restricted zone around themselves to offer less resistance to the fluid flow. At higher air flowing velocity, particles will be separated from each other and there is a movement of particles. Then the bed volume increases and this is the point of incipient fluidization.

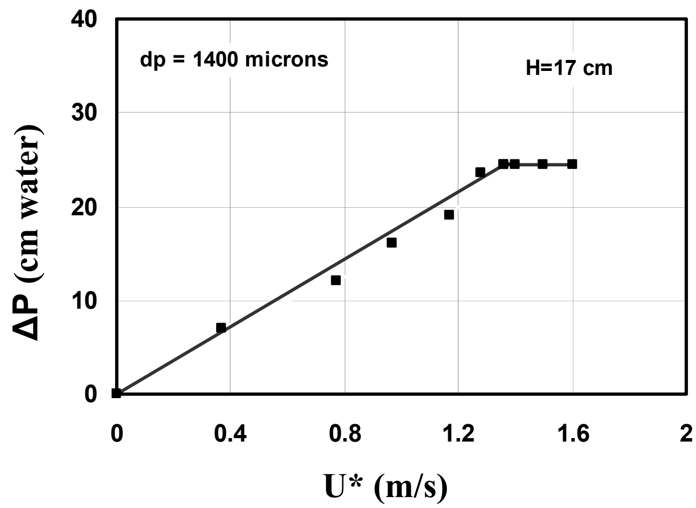

Fig. (6) Variation of pressure drop with air velocity for single tube immersed in a sand bed of d_p = 1400 microns and H = 17 cm

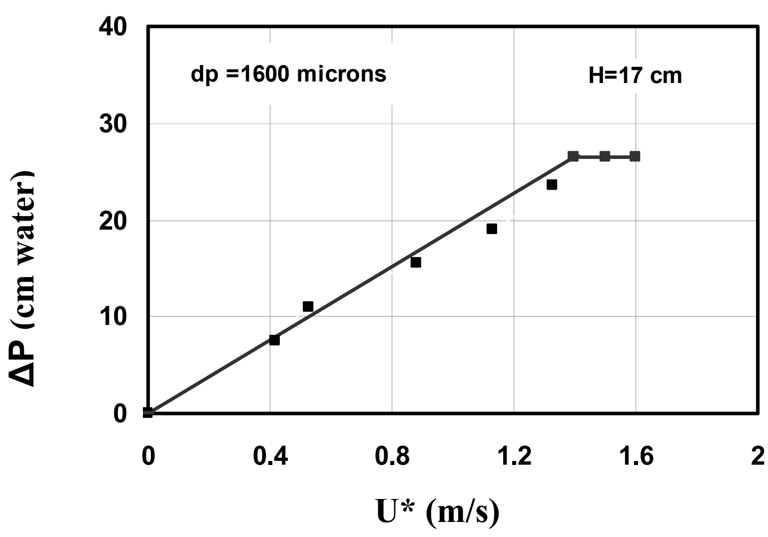

Fig. (7) Variation of pressure drop with air velocity for single tube immersed in a sand bed of d_p = 1600 microns and H = 17 cm

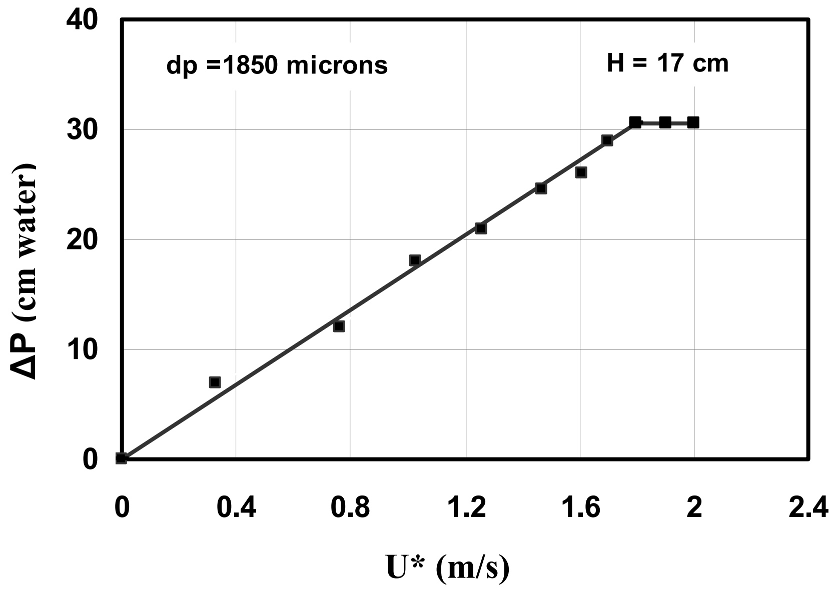

Fig. (8) Variation of pressure drop with air velocity for single tube immersed in a sand bed of d_p = 1850 microns and H = 17 cm

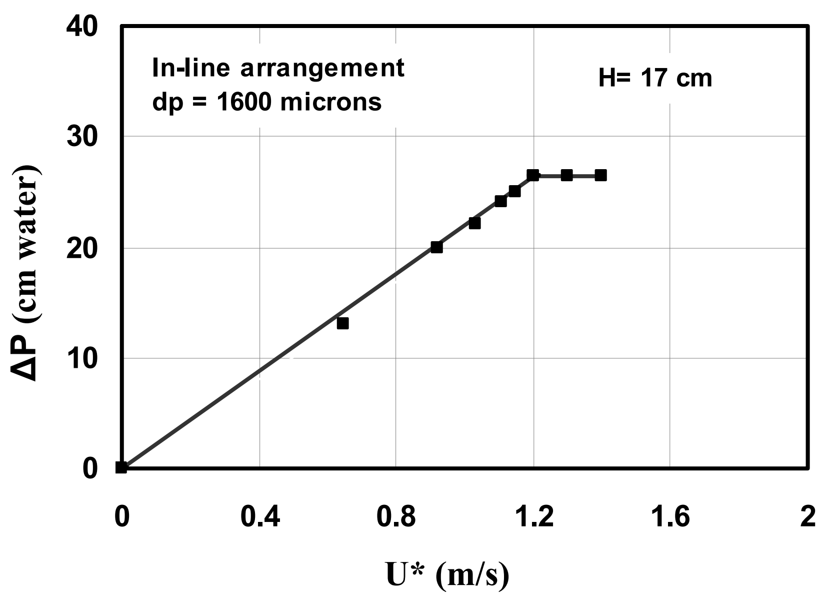

Fig. (9) Variation of pressure drop across the bed versus air velocity for in-line tube arrangement immersed in a sand bed of d_p =1600 microns and H =17cm

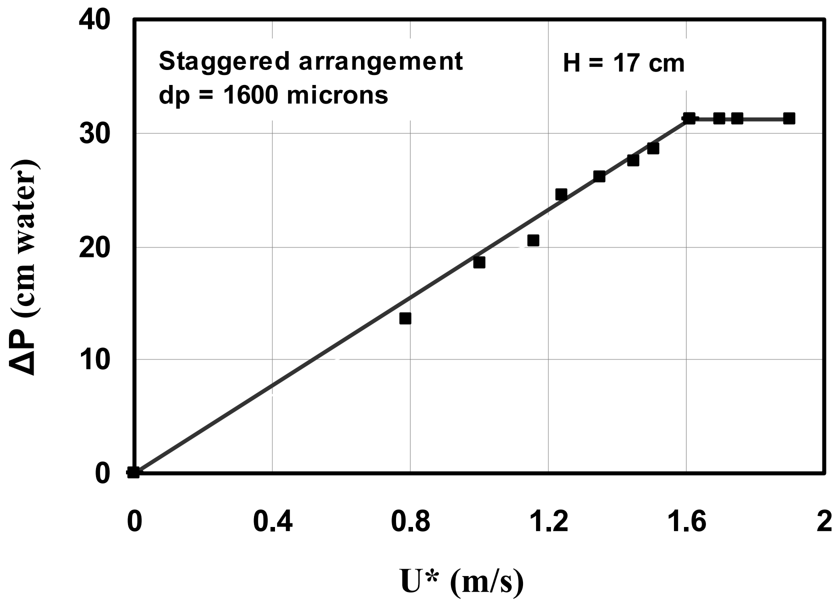

Fig. (10) Variation of pressure drop across the bed versus air velocity for staggered tube arrangement immersed in a sand bed of d_p =1600 microns and H=17cm

4.2. Heat Transfer Results

In the present experimental work, the effect of the particle diameter, air velocity, static head of the bed, and different tube arrangements on the average heat transfer coefficient in the bed are presented in the following sections.

4.2.1. Single tube

The variation of Nu_T with Re_T of bed for single tube with static head of 17 cm at different d_p is presented in Fig. (11). It is clear that the experimental results of the average Nusselt number, Nu_T, of the fluidized bed are significantly higher than that of the empty bed. It is concluded that, the value of Nusselt number is increased with the decrease of particle diameter for the same conditions of fluidized bed, within the Reynolds number range from 1860 to 2200. This result is due to that the net surface area of the particles in contact with the heated tube is higher for small solid particles than large particles [19]. It can be seen that, the Nusselt number of fluidized bed of particle diameter d_p =1400 µm is the highest one, while without sand (empty bed) is the lowest one.

In case of single tube with fluidized bed of different particle diameters (1400, 1600, and 1850 µm), the average heat transfer coefficient, h_w, increases with the increase of air velocity, while it decreases with the increase of particle diameter for the tested velocity range (from 1.33 m/s to 1.55 m/s) as shown in Fig. (12). Also, Fig. (13) shows the variation of the average heat transfer coefficient, h_w, with air velocity for single tube at different static heads of the bed with the same particle diameter (d_p =1400 μm). It is found that, the decrease of bed static head increases the average heat transfer coefficient for the tested velocity range (from 1.33 m/s to 1.99 m/s).

Figure (14) illustrates the variation of average Nusselt number, Nu_p, versus Reynolds number, Re_p, based on particle diameter for single tube. The results show that Nu_p for all tested particle diameters of the bed are higher than that of Zenz [29], this may be attributed to the difference in operating conditions. It is observed that, the average Nusselt number is increased with the decrease of particle diameter. Also, Fig. (15) indicates the comparison of the average Nusselt number, Nu_T, based on tube diameter for single tube with static head of 17 cm and modified Zenz correlation. It is clear that, the present results of single tube are higher than that of Zenz [29] for tested particle diameters d_p =1400 μm, and 1600 μm.

Figure (16) shows the comparison between the present work of single tube with Gunn (stated in Ref.[30]) modified correlation of Nusselt number, Nu_T, versus Reynolds number, Re_T, and the comparison suggest the flowing results: 1- The Nu_T of d_p =1400 μm is the highest one and in the same time slightly higher than that of Gunn [30]. 2- All the present work and Gunn [30] correlations have the same trend.

Figure (17) illustrates the comparison between the present work and Vreedenburg correlation [3]. Vreedenburg correlation was presented by a single line with percentage deviation of ±10%, as the correlation does not include d_p parameter. The results showed that Nu_T of d_p =1400 μm is the highest and Nu_T of d_p =1600 μm lies slightly higher than that of the Vreedenburg correlation, while Nu_T of d_p = 1850 μm is slightly lower than that of the Vreedenburg [3].

Finally, from the above results, it is concluded that the heat transfer coefficient for fluidized bed of all tested particle diameters is significantly higher than that of the empty bed. This can be a basis for design of a heat exchanger immersed in fluidized bed.

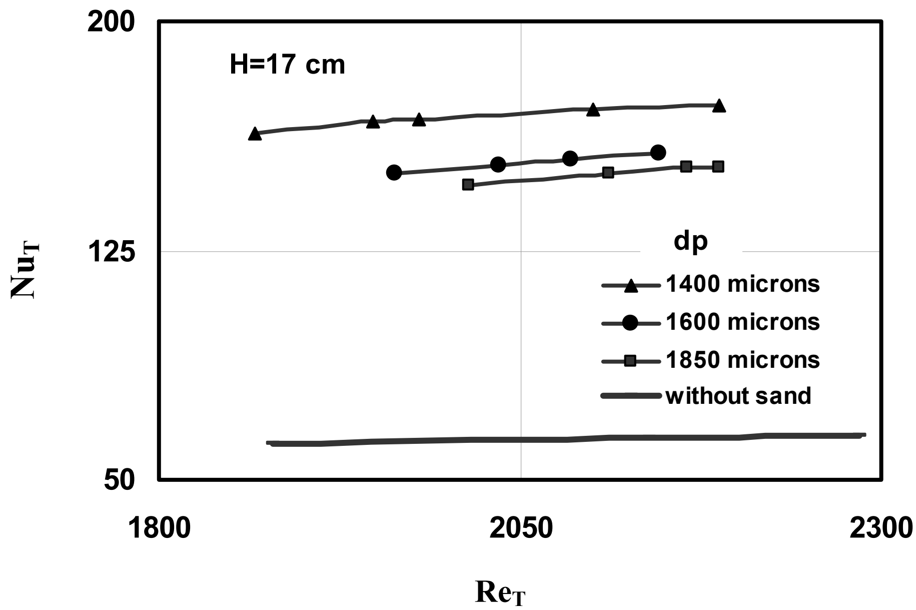

Fig. (11) Variation of average Nusselt number versus Reynolds number for fluidized bed (H=17cm) and empty bed having single tube

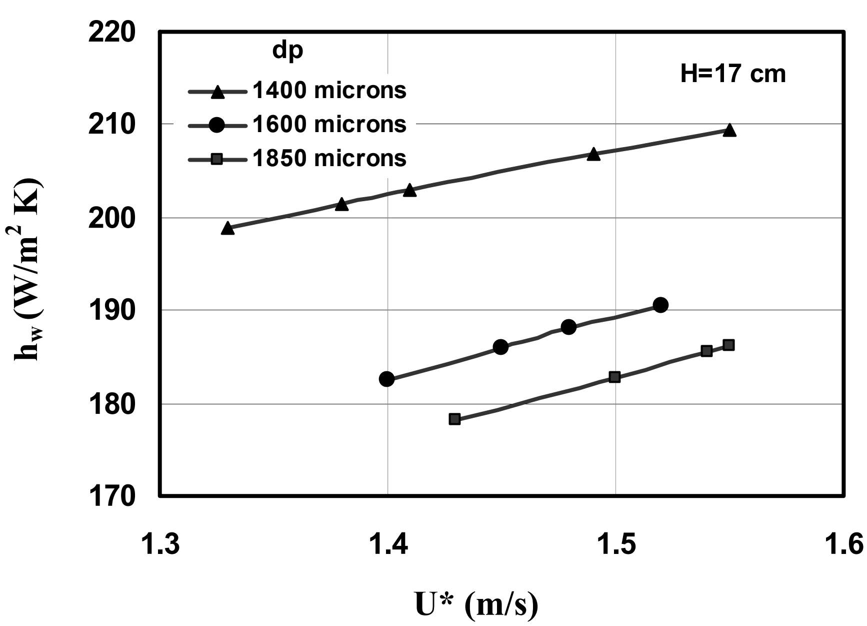

Fig. (12) Variation of the average heat transfer coefficient with air velocity of a single tube for a sand bed (H=17cm) at different particle diameters

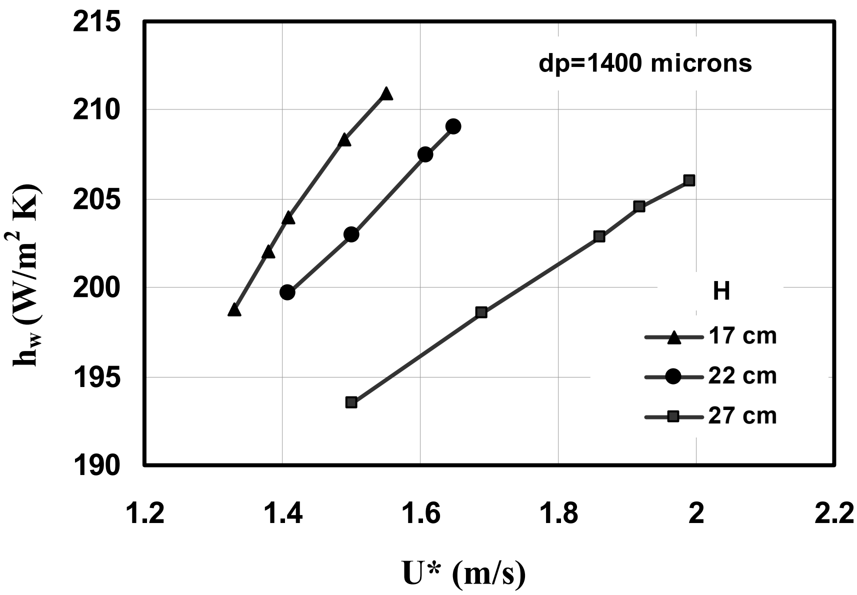

Fig. (13) Variation of the average heat transfer coefficient with air velocity for single tube immersed in a sand bed with different H and d_p = 1400 microns

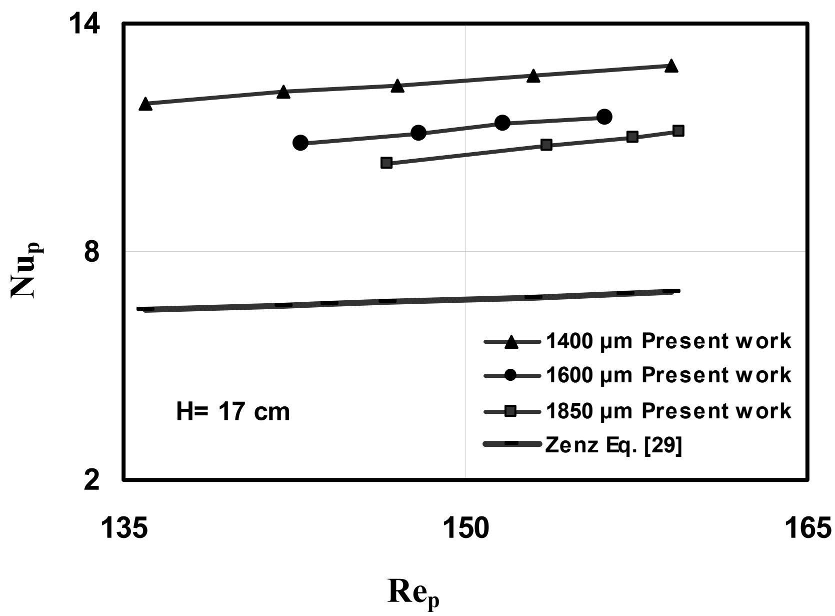

Fig. (14) Comparison between present work of Nu_p – Re_p for single tube immersed in a sand bed (H=17 cm) and that of Zenz [29]

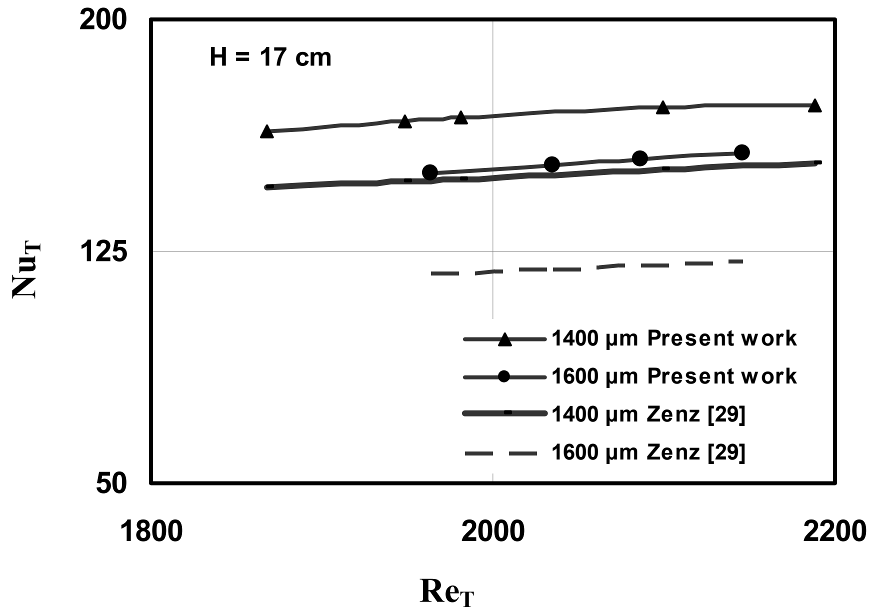

Fig. (15) Comparison between present work of Nu_T –Re_T for single tube immersed in a sand bed (H=17 cm) and that of Zenz [29]

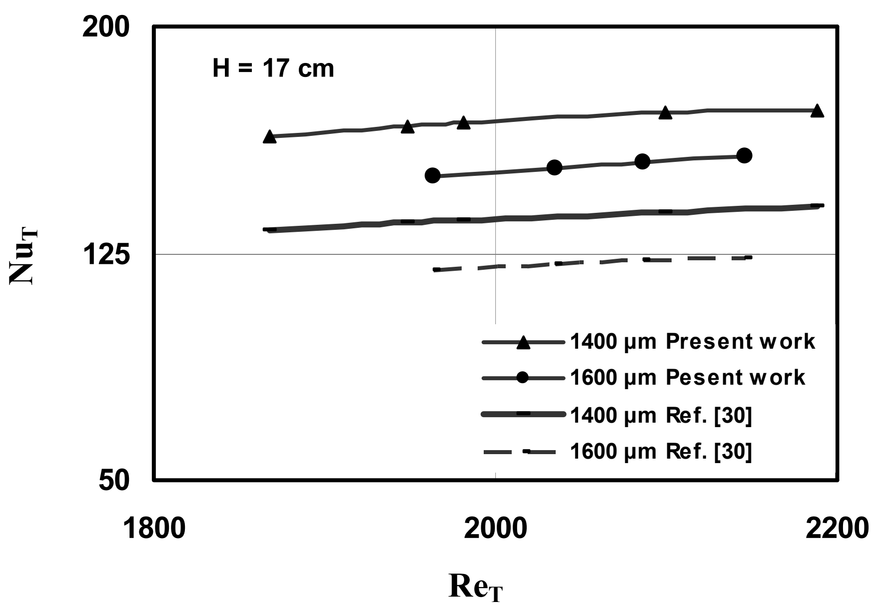

Fig. (16) Comparison between present work of Nu_T – Re_T of single tube immersed in a sand bed (H =17 cm) for different particle diameters with that of Gunn (Ref. [30])

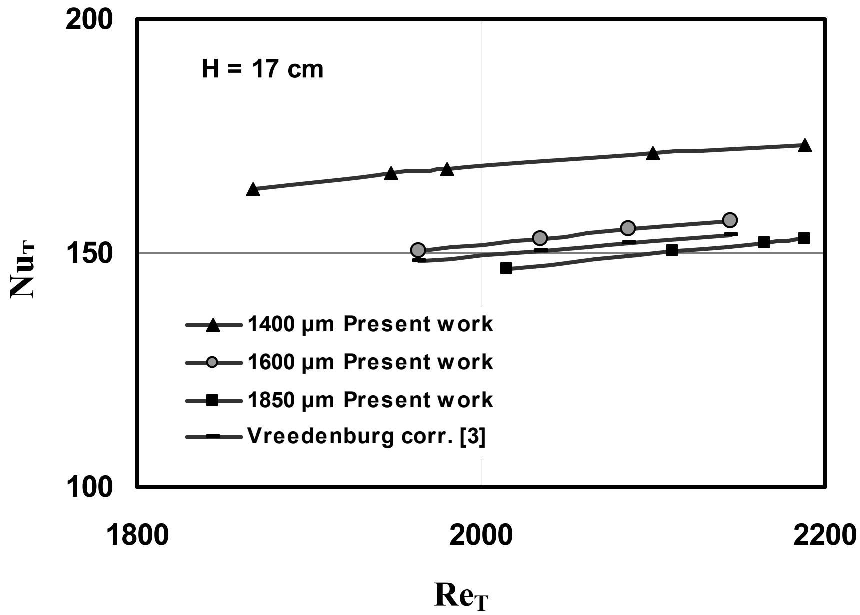

Fig. (17) Comparison between NuT – Re_T for single tube immersed in a sand bed (H =17 cm) of different particle diameters with that of Vreedenburg [3]

Moreover, an empirical correlation using the experimental results of single tube is obtained by the least square method for the average Nusselt number, Nu_T, based on tube diameter, as follows:

$$Nu_T = 34.548\left(\frac{\rho_s}{\rho_a}\frac{\mu_a^2}{d_p^3\rho_s^2 g}\right)^{0.258} \mathrm{Re}_T^{0.374}\,\mathrm{Pr}^{0.3} \tag{6}$$

This correlation, Eq. (6) satisfies the present experimental data within ±2% maximum deviation, as shown in Fig. (18) for the ranges of Reynolds number ($1800 \le Re_T \le 2200$), and particles diameter ($1400 \le d_p$ (µm) ≤ 1850).

4.2.2. Tube banks arrangement

Sand particles of diameters (1400, 1600, and 1850 µm) are used in all experiments with static head of 17 cm. The variation of the average heat transfer coefficient, h_w, with the air velocity for each of in-line and staggered tube arrangement at different particle diameters are indicated in Figs. (19) and (20), respectively. The results showed that the average heat transfer coefficient is increased with the increase of air velocity and with the decrease of particle diameter for both tested arrangements.

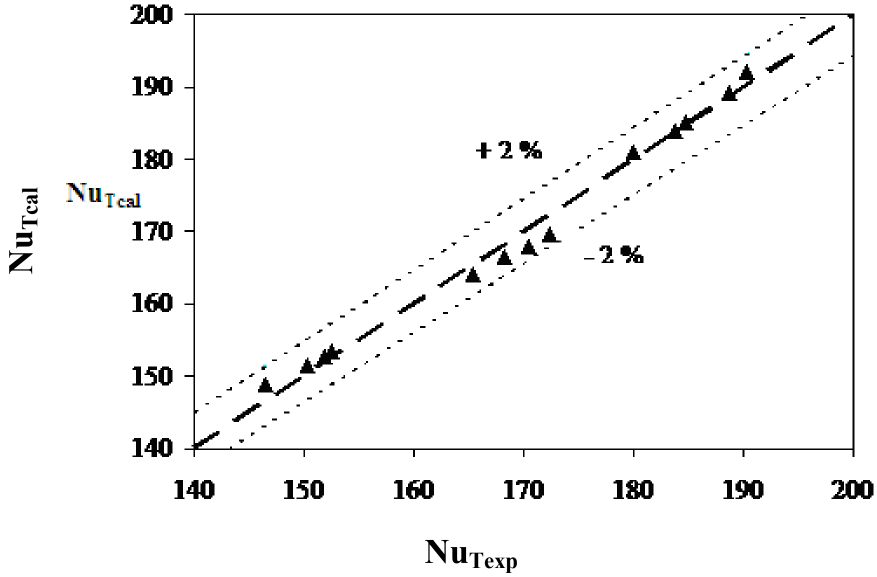

Fig. (18) Nu_{Tcal} against Nu_{Texp} for single tube immersed in a sand bed

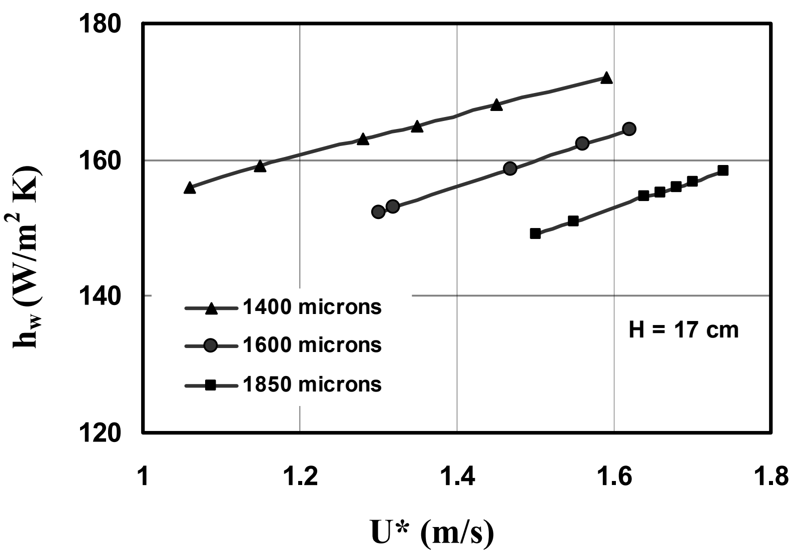

Fig. (19) Variation of average heat transfer coefficient with air velocity for in-line tube arrangement immersed in a sand bed (H=17 cm) of different particle diameters

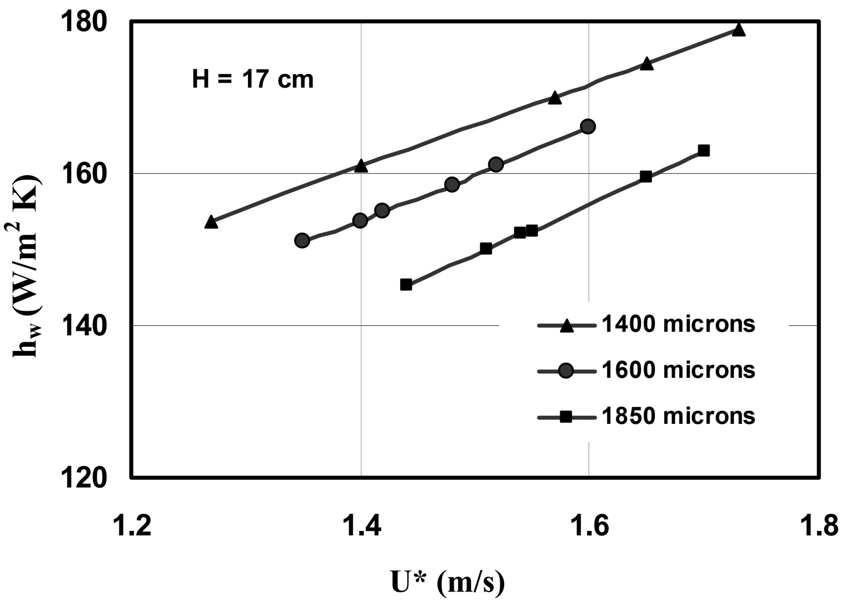

Fig. (20) Variation of average heat transfer coefficient with air velocity for staggered tube arrangement immersed in a sand bed (H=17 cm) of different particle diameters

Figures (21) and (22) show the variation of average Nusslet number, Nu_T, versus Reynolds number, Re_T, for beds of different d_p and empty beds of in-line and staggered arrangements, respectively. It is found that, the average Nusselt number is increased with the increase of Reynolds number and with the decrease of particle diameter for both tested arrangements. Also, the results of, Nu_T, for the three tested particles diameter are significantly higher than that of the empty bed.

The variation of average Nusselt number, Nu_p, with Reynolds number, Re_p, based on particle diameter at different tube bank arrangements for the three particle diameters of d_p =1400, 1600, and 1850 μm is shown in Figs. (23), (24), and (25) respectively. For d_p =1400 μm, it is found that the Nusselt number for staggered tubes is higher than that of the in-line arrangement by about 3.7% and 4.9% at Re_p=130 and Re_p = 161 respectively. Also, for d_p=1600 μm, it is observed that the Nusselt number for staggered tubes is higher than that of the in-line arrangement by about 1.3% and 5.8% at Re_p = 133 and Re_p = 151 respectively. Finally, for d_p=1850 μm it is noticed that the Nusselt number for staggered tubes is higher than that of the in-line arrangement by about 3.4% and 2.5% at Re_p = 167 and 174 respectively. It is concluded that, the average Nusselt number in case of staggered tube bank is higher than that in case of the in-line tube arrangement. This increase is due to the change of the bed hydrodynamic and the increase in the possibility of particles touch the wall of the heated tube and transfer more heat from it [2].

A comparison between the present data of in-line and staggered tube arrangements with Zenz correlation [29] are presented in Figs. (26), and (27), respectively. The following conclusions can be obtained: The Nusselt number (Nu_T) for d_p =1400 μm is closely lower than that of Zenz [29], specially at low Re_T. The Nu_T for d_p =1600 μm is slightly higher than that of Zenz [29] with the same trend.

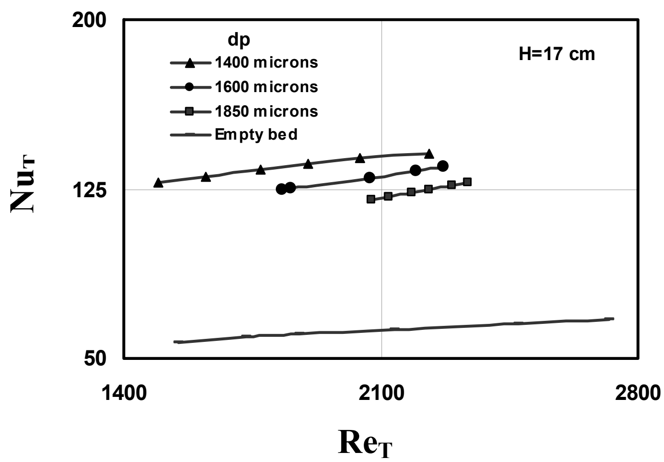

Fig. (21) Variation of Nu_T with Re_T for in-line tube arrangement immersed in a sand bed (H=17 cm) of different particle diameters

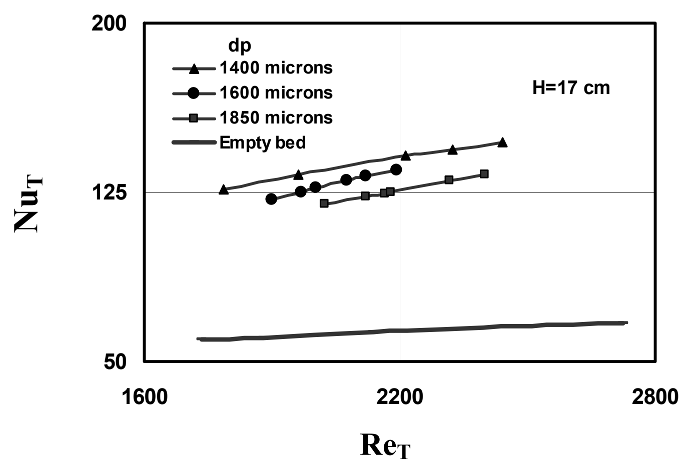

Fig. (22) Variation of Nu_T with Re_T for staggered tube arrangement immersed in a sand bed (H=17 cm) of different particle diameters

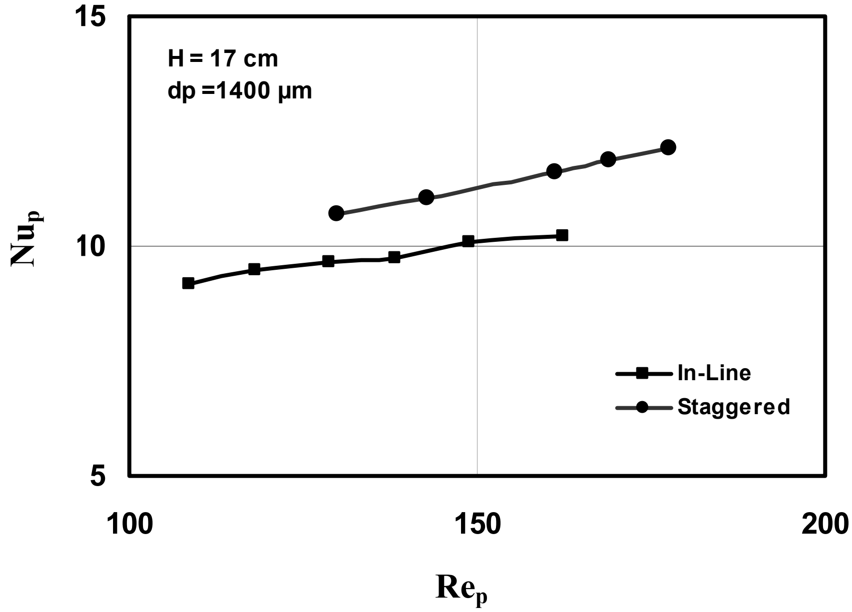

Fig. (23) Variation of Nu_p with Re_p for different tube arrangements immersed in a sand bed (H=17 cm) of d_p = 1400 microns

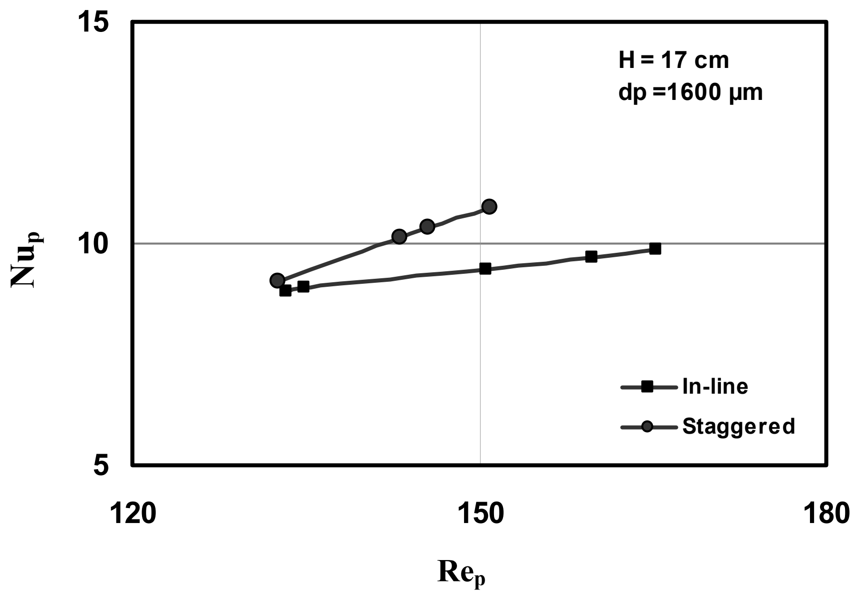

Fig. (24) Variation of Nu_p with Re_p for different tube arrangements immersed in a sand bed (H=17 cm) of d_p = 1600 microns

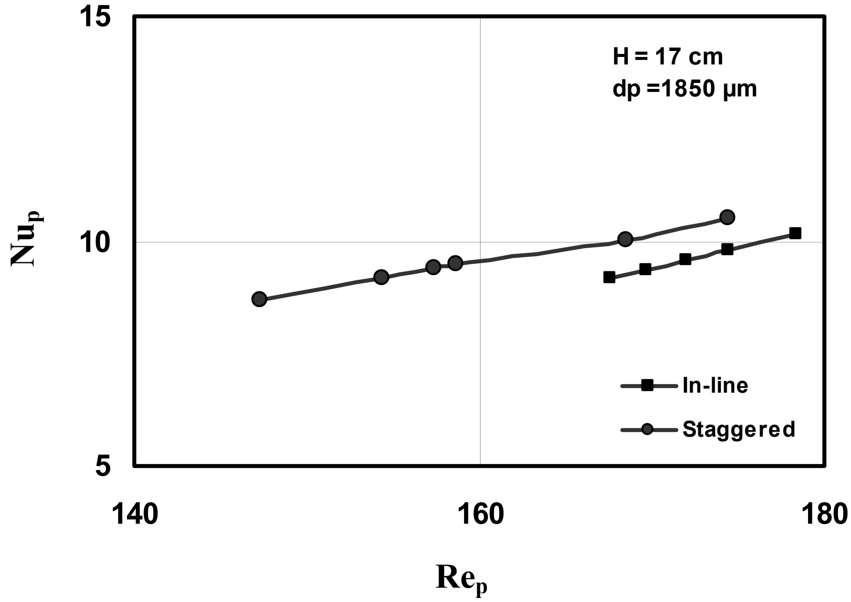

Fig. (25) Variation of Nu_p with Re_p for different tube arrangements immersed in a sand bed (H=17 cm) of d_p = 1850 microns

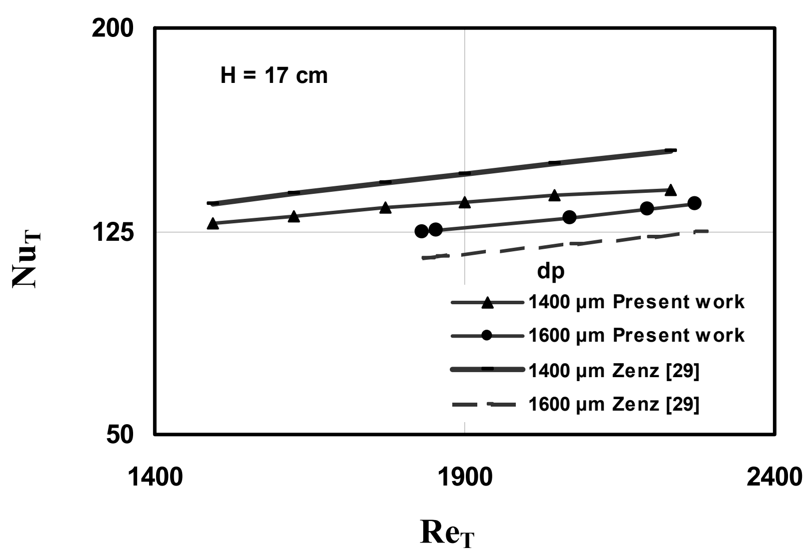

Fig. (26) Comparison between in-line tube arrangement data for different sand particle diameters (H=17 cm) with that of Zenz correlation [29]

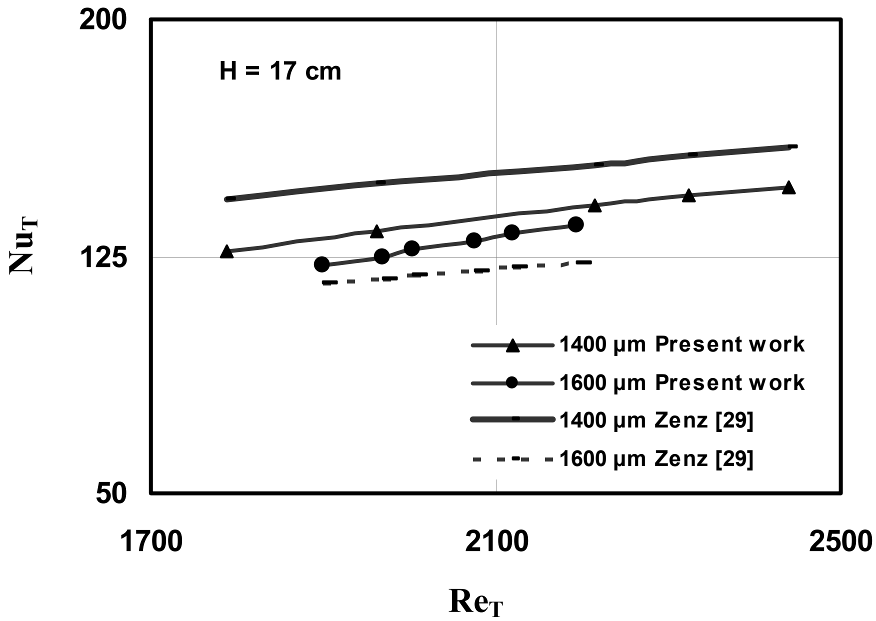

Fig. (27) Comparison between present $Nu_T - Re_T$ data of staggered tube arrangement immersed in a sand bed (H =17 cm) of different particle diameters with that of Zenz correlation [29]

Also, a comparison between the present results of in-line and staggered tube arrangements, with Vreedenburg correlation [3] are presented in Figs. (28) and (29), respectively. Vreedenburg correlation is represented by a single line, as the correlation does not contain d_p parameter. The results of the two figures are: The Nu_T of d_p = 1400 µm is higher than that of d_p = 1600 µm and in turn it is higher than that of d_p = 1850 µm and all of them are lower than that of the Vreedenburg correlation [3]. All the present results of Nu_T are ranged inside the domain of percentage deviation of ±10% of the Vreedenburg correlation [3].

Finally, the Experimental results are fitted using the power regression and the correlation of the Nu_T for in-line and staggered arrangements are expressed as the following:

- **For in-line tubes**:

$$Nu_T = 36.9\left(\frac{\rho_s}{\rho_a}\frac{\mu_a^2}{d_p^3\rho_s^2 g}\right)^{0.156} \mathrm{Re}_T^{0.278}\,\mathrm{Pr}^{0.3} \tag{7}$$

for $1400 \le Re_T \le 2300$ and $1400 \le d_p\,(\mu m) \le 1850$

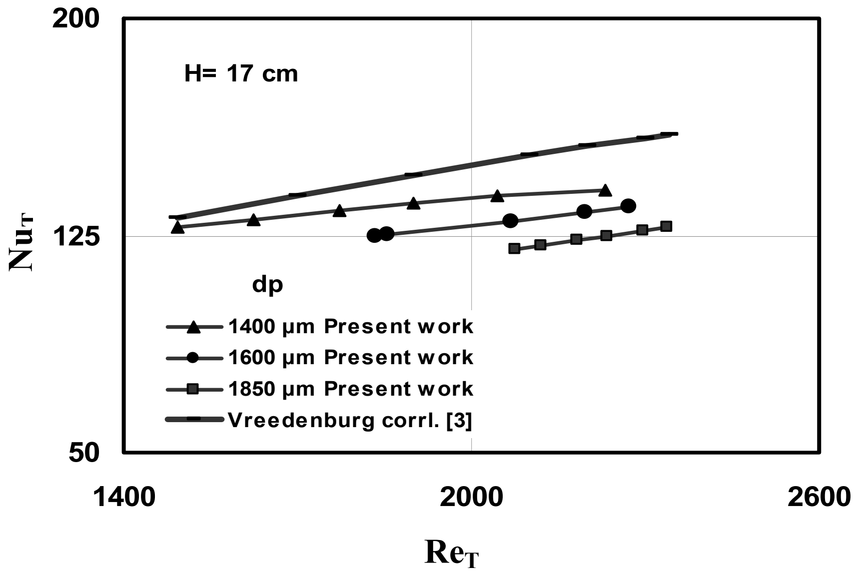

Fig. (28) Comparison between present $Nu_T - Re_T$ data of in-line tube arrangement immersed in a sand bed (H =17 cm) of different particle diameters with that of Vreedenburg correlation [3]

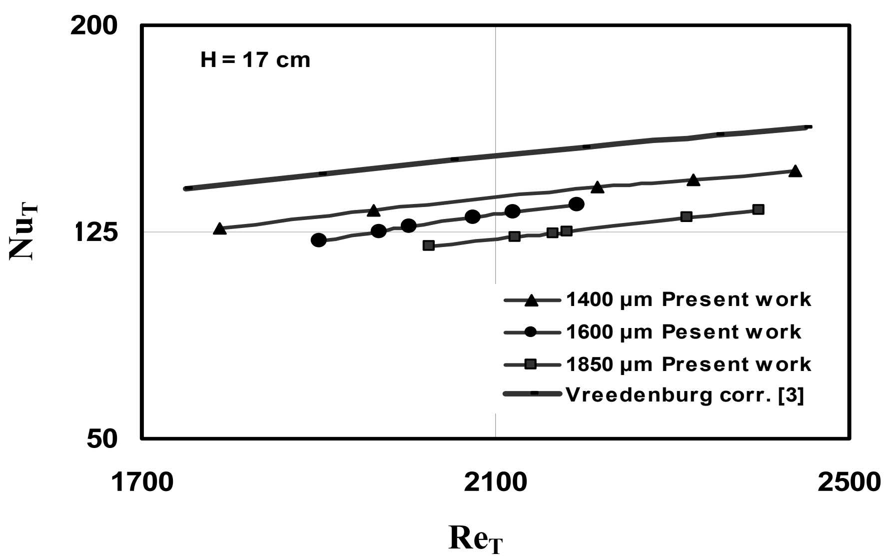

Fig. (29) Comparison between present $Nu_T - Re_T$ data of staggered tube arrangement immersed in a sand bed (H = 17 cm) of different particle diameters with that of Vreedenburg correlation [3]

The calculated average Nusselt number (Nu_{TCal}) from Eq. (7) is plotted versus experimental average Nusselt number (Nu_{TExp}) in Fig. (30). As shown from this figure the maximum deviation between the experimental data and the correlation, Eq. (7) is ±2%.

- **For staggered tubes**:

$$Nu_T = 4.97\left(\frac{\rho_s}{\rho_a}\frac{\mu_a^2}{d_p^3\rho_s^2 g}\right)^{0.124} \mathrm{Re}_T^{0.519}\,\mathrm{Pr}^{0.3} \tag{8}$$

for $1700 \le Re_T \le 2400$ and $1400 \le d_p\,(\mu m) \le 1850$

The calculated average Nusselt number (Nu_{TCal}) from Eq. (8) is plotted versus experimental average Nusselt number (Nu_{TExp}) in Fig. (31). As shown from this figure the maximum deviation between the experimental data and the correlation
is ±2%.

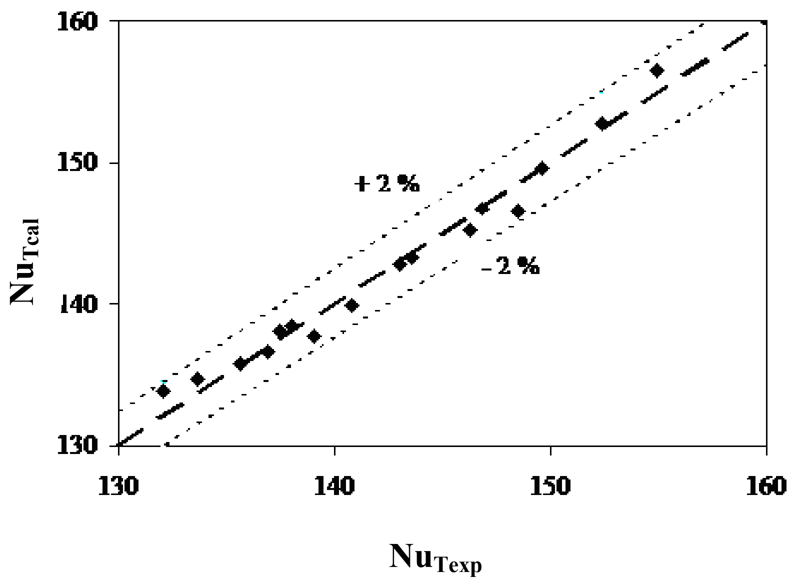

Fig. (30) Nu_{Tcal} against Nu_{Texp} for in-line tubes immersed in a sand bed

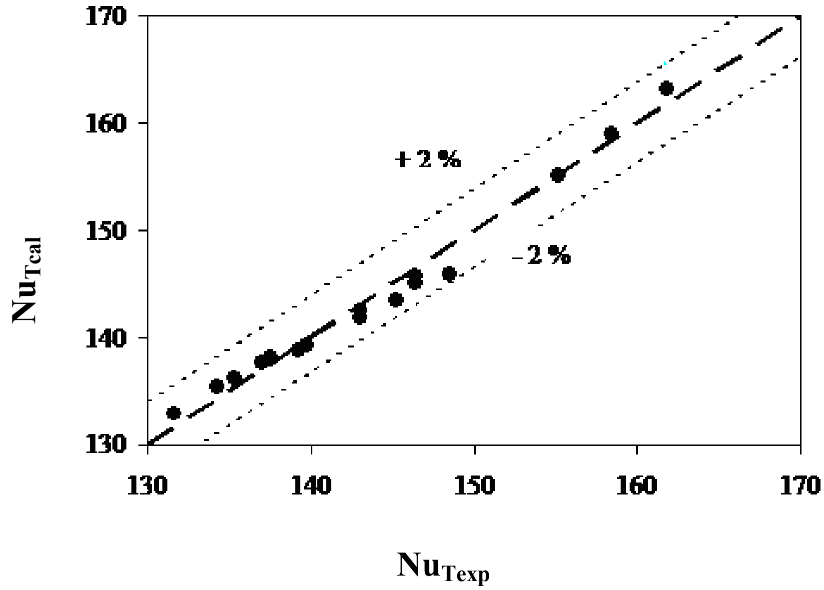

Fig. (31) Nu_{Tcal} against Nu_{Texp} for staggered tubes immersed in a sand bed

5. CONCLUSIONS

This research study the heat transfer characteristics around a single horizontal heated circular tube, in-line circular tubes arrangement and staggered circular tubes arrangement immersed in a gas fluidized bed. The results of this research can be summarized as:

1. The average heat transfer coefficient increases with the increase of the fluidizing velocity.
2. The average heat transfer coefficient increases with the decrease in particles diameter.
3. The increase of bed particles diameter increases the minimum fluidized bed velocity at the same static head.
4. The Nusselt number in case of fluidized bed is greater than that of empty bed.
5. The Nusselt number increases with the decrease of particle diameter.
6. The minimum fluidized bed velocity in staggered arrangement is greater than that in-line arrangement.
7. The average heat transfer coefficient in case of staggered is higher than that of the in-line tube arrangement.

8. Empirical correlations for the average Nusselt number of the tested arrangements (single tube, in-line and staggered tubes arrangements) using the experimental data are presented

NOMENCLATURE

A	Surface area of heated tube, m^2
C_{pa}	Specific heat at constant pressure of fluidizing air, J/kg K
C_{ps}	Specific heat at constant pressure of solid particles, J/kg K
CFB	Circulating fluidized bed
d_p	Mean particle diameter, m
D_T	Outside diameter of heated tube, m
g	Acceleration due to gravity, m/s^2
G	Air mass velocity, $kg/m^2.s$
h_w	Average heat transfer coefficient, W/m^2 K
H	Static bed height, m
k_a	Thermal conductivity of air, W/m K
Nu_p	Nusselt number based on particle diameter = $h_w d_p/k_a$
NU_T	Nusselt number based on tube diameter = $h_w D_T/k_a$
Nu_{Tcal}	The calculated average Nusselt number
Nu_{Texp}	The experimental average Nusselt number
Pr	Prandtl number = $\mu C_{pa}/k_a$
Q	Electrical power supplied to heater, W
Re_p	Reynolds number based on particle diameter = $\rho_a U^* d_p/\mu_a$
Re_T	Reynolds number based on tube diameter = $\rho_a U^* D_T/\mu_a$
T_b	Average bed temperature, K
T_w	Average wall temperature of heated tube, K
U	Mean velocity of air under the distributor plate, m/s
U^*	Average velocity of air inside the test section, m/s

Greek Letters

ΔP	Pressure drop across the fluidized bed, cm water
ρ_a	Density of air, kg/m^3
ρ_s	Density of solid particle, kg/m^3
μ_a	Dynamic molecular viscosity of air, N S/ m^2

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