TRENSENT HEAT TRANSFER IN CIRCULATING FLUIDIZED BED

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ABSTRACT:

The paper presents practical data from the experimental work on heat transfer between immersed heat transfer surface and Circulating Fluidized Bed (CFB) in special experimental rig with certain dimensions and with set of operation conditions at constant heat flux (120 W) supplied to the electrical heater. The fluidizing medium was air at different velocities (4 and 5.5 m/s)and two different size of sand particles were employed (161 and 257 µm), as well as , different initial bed heights were used in the experiments (i.e, 18 and 31 cm).

Heat transfer coefficients calculated instantaneously by instant temperature of bed and heater surface are found to increase with fluidized air velocity and with heat flux, but, they show an inverse dependence on particle size, and directly proportional with initial bed height which represents the bed density.

The stabilization case depended strongly on fluidization density, where the required time to reach steady state case of heat exchange increase with increasing initial bed height as well as increasing the particle size.

The experimental heat transfer coefficient data were correlated in terms of Nusselt number with other parameters. This correlation of the present study are assessed by comparing it with the available correlation in literature for the overall heat transfer coefficient. Comparison shows reasonable agreements in some results and large deviation in others. Also this correlation not to correspond with the experimental values in some conditions.

انتقال الحرارة بالحالة غير المستقرة في الأبراج ذات الطبقات المميعة المدورة

الخلاصة

في هذا البحث تم أجراء تجارب عملية في انتقال الحرارة بالحالة غير المستقرة للطبقة المميعة من خلال تصميم و تشغيل منظومة جريان مميع بمواصفات خاصة وتحت ظروف تشغيلية معينة لضمان حالة التدوير للمادة الصلبة المستخدمة (الرمل)حيث تم تجهيز المسخن الأنبوبي المغمور أفقاً داخل حشوه بقدرة ثابتة (120 واط)،حيث استخدمت دقائق الرمل كحشوه وبأقطار (161 و 257 مايكروميتر)، وهذه الحشوات بارتفاعات أبتدائيه (18و31 سم)،وأستخدم الهواء كوسط للتميع وبسرع مختلفة (4، 5,5 م\ثا . حيث وجد ان زيادة ارتفاع الحشوة الابتدائية سوف يؤدي بشكل متناسب الى زيادة معاملات انتقال الحرارة، كونه يمثل زيادة كثافة الجريان وتقليل النفاذيه. أضافه إلى ان حالة الاستقرار في النتائج تعتمد بشكل مباشر على الظروف

الهيدروليكية للجريان وبالأخص سرعة الجريان حيث ان الوقت ألازم لوصول عملية التبادل الحراري إلى حالة الاستقرار يزداد مع زيادة كثافة الجريان وكذلك مع زيادة معدل حجم حبيبات الرمل. من خلال معاملات انتقال الحرارة العملية تم إيجاد معادله رياضيه باستخدام القيم اللابعديه ، أجريت مقارنة لمعاملات انتقـــال الحـــرارة الماخوذه من هذه العلاقة مع معاملات انتقال الحرارة المحسوبة من أحدى العلاقات المتوفرة في الأدبيات حيث أظهرت النتائج وجود اختلاف مع هذه العلاقة. كذلك يلاحظ عدم التطابق بشكل تام بين العلاقة المستنتجة والقيم المأخوذة من التجربة.

الاستقرار يزداد مع زيادة كثافة الجريان وكذلك مع زيادة معدل حجم حبيبات الرمل. من خلال معاملات انتقال الحرارة العملية تم إيجاد معادله رياضيه باستخدام القيم اللابعديه , أجريت مقارنة لمعاملات انتقال الحرارة الماخوذه من هذه العلاقة مع معاملات انتقال الحرارة المحسوبة من أحدى العلاقات المتوفرة في الأدبيات حيث أظهرت النتائج وجود اختلاف مع هذه العلاقة. كذلك يلاحظ عدم التطابق بشكل تام بين العلاقة المستنتجة والقيم المأخوذة من التجربة.

Introduction:

Circulating Fluidized Bed (CFB) technology has been widely used for various gas-solid reactions such as catalytic cracking ,combustion and other reactions which commonly require heat transfer during the reactions, and CFB is one of recently developed regimes of fluidization, which belongs to the group of so called transport regimes. The characteristics of CFB is continual entrainment of solids out the bed. The solids circulate along the circulation loop, which connects all parts of the CFB. The condition of CFB is remarked by relatively high solids concentration, agglomeration of solid into clusters and strands which are destroyed quickly and formed again by intensive back mixing of particles (Kunii and Levenspiel, 1977).

Radially nonuniformity of bed voidage, at least for smaller bed cross-section is widely accepted. There are some doubts about voidage distribution in large industrial unit, due to lack of data for large diameter bed. There is general consensus about the existence of a denser bed in the lower part and relatively dilute regions in the upper part of the reactor. Relative position of those phases in a system depends on gas velocity, particle circulation and pressure drop. The solid from group B of Geldart's classification (ρ_p =1400 ~ 4000 kg/m^3 and d_p= 45 ~ 500 µm)are usually used in CFB (Grace 1986).

Fundamental investigations of the CFB are behind it's practical application in industry. It is obvious, especially in the field of CFB hydrodynamics, because there is no unique model of hydrodynamics. Empiric nature of the published models limits their application on experimental conditions from which they are derived. The majority of hydrodynamic models of CFB consider axial profile of particle or gas phase concentration. Some models follow the investigations of bubbling fluidized beds and use some expressions from bubbling beds for some regions of CFB.

Heat Transfer Models of CFB

Many circulating fluidized beds involving combustion or other exothermic reactions commonly require heat exchange during the reaction, where the riser is the main column in which major reactions and heat transfer process occur. Heat transfer phenomena in the riser is therefore critical to the design, operation and control of CFB reactions. With CFB riser becoming more and more popular, especially in the last two decades (Grace et al, 1997), an accurate understanding of heat transfer in CFB is very important for the proper design of CFB reactors. Many studies have been carried out to test the effect of different design and operation parameters on heat transfer (Grace ,1990; Basu 1996).

Three main heat transfer process in CFB, as well as in bubbling beds occur, the most important of which is heat transfer between CFB and heat transfer exchange surfaces. Other two process, between particles and gas and heat transfer from one part of the bed to the other, are less important for investigation, because active zone of heat transfer at the entrance of the bed is so short, that the bed is considered isothermal throughout the volume. High heat transfer coefficients between heat exchange surfaces and a bed enable heat transfer and temperature drop to occur just near heat exchange surface, while the rest of the bed remains isothermal. Experimental investigation are conducted only on laboratory scale apparatuses. There are still not published results about experiments on large industrial units, and it is a good reason to suppose that such results can somewhat change existing understanding of heat transfer in CFB.

Heat transfer between suspension and heat exchange surface is the most important heat transfer process in CFB. Overall heat transfer coefficient suspension-surface can be divided into three separate components as follow:

$$h = h_{gc} + h_{pc} + h_r \qquad \ldots(1)$$

Models of heat transfer between suspension and heat exchange surfaces in CFB are based on the models of heat transfer in conventional fluidized beds. The base of those models is the "packet model" of Mickley and Fairbanks(1955) and the modifications of Baskakov (1978), who introduced additional thermal resistance of the gas film on heat transfer surface. In these models heat transfer by the particle convection is modeled as the process of unstationary conduction of the particle clusters which are in contact with heat exchange surface for a definite period of time, nearby the heat exchange surface there is a gas film which transfers heat by gas conduction. Particle convection heat transfer coefficient is defined as:

$$h_{pc} = \frac{1}{\frac{1}{h_w} + \frac{1}{h_s}} \qquad \ldots(2)$$

Heat transfer coefficient of the gas conduction though the gas boundary layer on the heat exchange surface depends on gas film thickness:

$$h_w = \frac{k_g}{\delta} \qquad \ldots(3)$$

The thickness of gas film has the values of (0.1 - 0.4) d_p (Wu et al., 1990). The component of particle convection h_s ,which comes from suspension or clusters is modeled depending on suspension hydrodynamics on heat exchange surface.

Although there is great variety of heat transfer models, the majority of them consider particle concentration, or the bed voidage as the most significant hydrodynamics factor in heat transfer between the bed and heat transfer surfaces.

Yoshida et al.(1969) derived a model for bubbling bed from the following equation which represents the phenomenon they discussed:

$$\frac{\partial T}{\partial t} = \frac{k_e}{\rho_e . Cp_s} \frac{\partial^2 T}{\partial x^2} \qquad 0 < x < I_e \qquad \ldots(4)$$

Where, I_e is the effective thickness of emulsion layer. A mechanism of heat transfer between fluidized beds and wall surfaces was proposed which includes both steady stat conduction of heat through an emulsion layer to the wall and the unsteady state absorption of heat by emulsion elements. They developed a criterion suggesting which mechanism controls, this criterion determined the controlling step in the heat transfer process was given as:

$$Z = \left[\frac{k_e . t_{cl}}{\rho_s . Cp_s (1 - \varepsilon_{mf})} \right]^{0.5} \Big/ I_e \qquad \ldots(5)$$

(i) Unsteady absorption into emulsion elements dominates when Z<1.
(ii) Steady state transfer across the emulsion layer dominates when Z>1.

Also, some measured data found to be consistent with the fluidized bed proposed by Kunii and Levenspiel (1977).

Gabor (1982) used the method presented by Botterill (1963). The differential equation for solid phase and gas phase in unsteady state conditions were solved numerically. Two models are presented for unsteady stat heat transfer from a wall to a bed of particles that may be either fluidized or packed.

An expression for estimating the heat transfer coefficient in fluidized beds has been developed by Ehung et al. (1993) based on the surface renewal and penetration concept. The expression was:

$$Nu = \frac{13.5\left[b^2 + 0.4 \frac{k_g}{k_s} \left(b^2 + 5b - \frac{d_p^2}{2\alpha\tau} \right) \right]}{\left(b + 1.6 \frac{k_g}{k_s} \right)} \qquad \ldots(6)$$

Where;

$$b = \frac{d_p}{\sqrt{2\alpha\tau}} \coth \frac{d_p}{\sqrt{2\alpha\tau}} - 1$$

$$\tau = \frac{k_s . t}{\rho_s Cp_s R^2}$$

They concluded that the only conditions under which the conductivity of particles may have influence on the heat transfer coefficient are when both $\frac{d_p}{\sqrt{2\alpha\tau}}$ and (k_s/k_g) are very large.

Basu et al.(1988) proposed a model of heat transfer in circulating bed, based on the theory of clusters, where total heat transfer coefficient (include radiation) is defined as:

$$h = h_{cl}\delta_{cl} + (1-\delta_{cl})h_d + h_r \qquad \ldots(7)$$

Where δ_{cl} is the part of the heat exchange surface covered by clusters:

$$\delta_{cl} = \frac{x(1-\varepsilon_{cl}) - y}{1-\varepsilon_{cl} - y} \qquad \ldots(8)$$

In the equation (8) y is the volume fraction of solid in the dilute phase (out of clusters) and x is the ratio of volume concentration of solid on the wall to that averaged over the cross-section of the bed. Particle convective component from clusters of voidage ε_{cl} , which are in contact with heat exchange surface during the period of time t_{cl} is :

$$h_{cl} = \frac{1}{\frac{d_p}{10k_g} + \sqrt{\frac{t_{cl}}{4k_{cl}\rho_s(1-\varepsilon_{cl})}}} \qquad \ldots(9)$$

Particle convection component from clusters is directly proportional to the concentration of solid in the clusters $C_{cl} = (1-\varepsilon_{cl})$. Since the component h_{cl} is dominant in the great part of CFB operation range, the influence of particle concentration is very significant to heat transfer between immersed heat transfer surfaces and the circulating fluidized bed. It means that particle concentration increase directly increases total heat transfer coefficient.

The model of Molerus (1993), primary developed for conventional fluidized beds, can be applied on CFB, since it is valid in the wide range of concentrations from C_{mf} to $C \to 0$ and high fluidization velocities, that cover CFB. Martin model predicts strong dependence of heat transfer on average solid concentration, which is in agreement with experimental results:

$$\frac{h_{pc}d_p}{k_g} = CZ\left[1 - \exp\left(Nu_{pc} / fZ\right)\right] \qquad \ldots(10)$$

The draw back of this model is that it does not take into account the average concentration of the bed cross-section, and not the one around the heat transfer surfaces.

Experimental Investigations

An experimental investigation of heat transfer in CFB was conducted on laboratory scale apparatuses made of Plexiglas. Column or a riser has 76mm inner diameter and 1.5m height. On the outlet of the column, there is a cyclone 220mm in diameter and 880mm in height. Its effect is the recovery of the solids through the recirculation column, or return-leg and through the flow-valve back to the circulating bed as shown in Figure (1). Flow valve is a non-mechanical V-valve, or loop seal , designed as a double chamber. Solids from the return-leg come to first chamber and flow over the second, where bubbling fluidization is affected by fluidization air. Fluidized solids are rising to the top of the chamber and returned to the column through the connecting pipe. Heat exchange surface is a copper tube with electric heater(120 W)inside, and thermocouples on outer surface. Fluidization air is divided on main stream for CFB and secondary air for bubbling fluidized bed in the flow valve. By secondary air quantity the dosing of solids into the bed is regulated. The power of electric heater is regulated by voltage variation with a variable transformer. Air temperature is measured by (k-type) thermocouples. The bed temperatures measured at six points through the column starting at height 0.65m above the distributor plat.

Heat transfer coefficient between heat exchange surface and the CFB is estimated as:

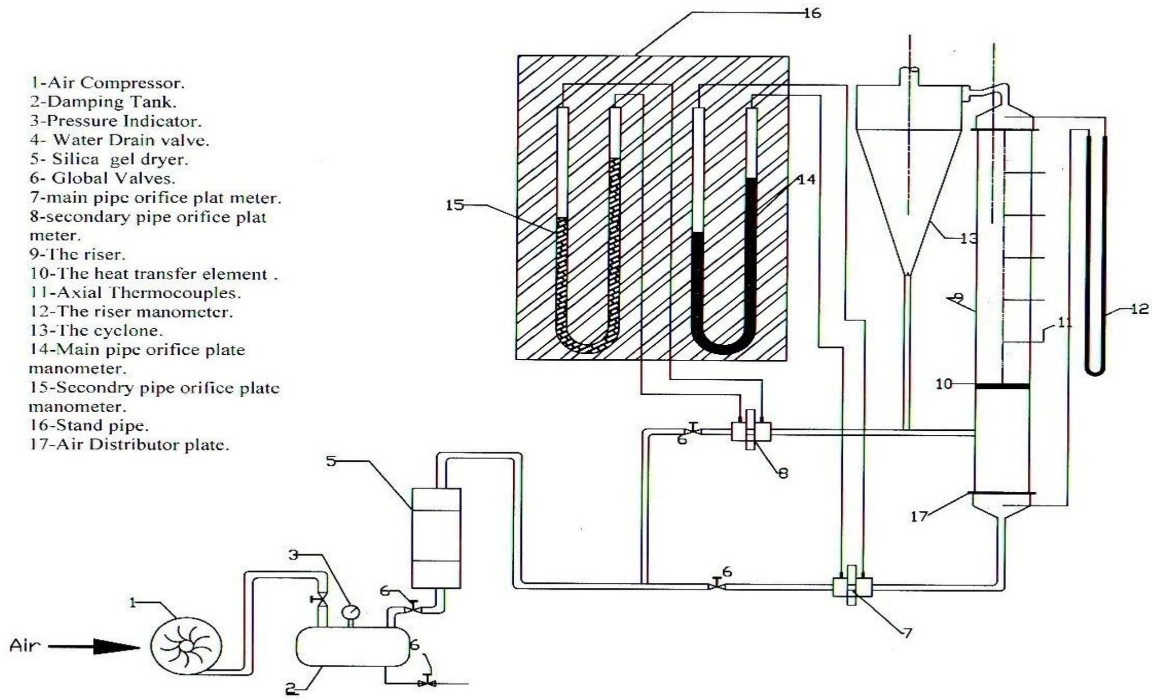
$$h = \frac{q}{A_h (T_s - T_b)} \qquad \ldots(11)$$

The sand which was used in the experiment has the following properties:

- mean particle diameter d_p = 161 and 257 µm.

- bulk density ρ_p = 2213 and 2267 kg/m^3.

- Geldart's classification group B.



Figure(1): Schematic Diagram of the Experimental Setup.

Results and Discussion

From local instantaneous bed temperature measured by (*k-type*) thermocouples, the heat transfer coefficient calculated according to Eq.(11). Figure(2) shows the results of these measurement, which represent the clear picture of relation of transient heat transfer coefficient with operation variables such as particle diameters, initial bed height or (gas-solid fluidized bed density), and fluidization velocity. These heat transfer coefficients are directly proportional with initial bed height and fluidization velocity, and inversely proportional with particle diameter. From each curve in Figure(2) it is noted that the differences of heat transfer coefficient of different initial bed height not large clear in first minute approximately, because the rapid mixing of flow in heat transfer region of column (the region around heat transfer element). Also the recorded time to reach the steady stat heat transfer coefficient depends upon the hydrodynamic behaviors of circulating fluidized bed and this complexity of this flow, where the heat transfer in circulating fluidized bed is one of the anomalies of the fluidization process because there is an unknown and complicated interaction between various system and operating parameters which make it impossible to design a finite set of experiments to establish uniquely the role and influence of each of parameters on the main element in heat transfer process, i.e. the heat transfer coefficient. The reach to steady state heat transfer became quickly with reduction of initial bed height and small bed diameter.

The experimental data of transient heat transfer coefficient of the present study can be used to predict an empirical correlation using dimensionless group to correlate the transient heat transfer coefficient as a function of effective parameters such as particle diameter, fluidization velocity, and initial bed height. Such relation is shown in the following equation.

$$Nu = \frac{0.2082\left(F.\mathrm{Re}/Ar\right)^{0.955}}{\left(\mathrm{Re}^{0.232}/Ar^{1.325}\right)+0.3113F^{3.35}} \qquad \ldots(12)$$

where; $F = \dfrac{k_s.t}{\rho_s.Cp_s.d_p.Hi}$

The above empirical correlation was determined by using the " STATISTICA" method with Proportion of variance accounted for (R^2 = 0.855939310).

The comparisons of the present Nusselt number with values obtained from Eq.(6) (Ehung model) and original experimental values to explained the differences between these values of Nusselt numbers are presented in Figures. 3 and 4. These comparisons are shown the clear differences between the Nusselt number computed from original experimental values and its value computed from above empirical correlation because the difficulty to make the experimental values as empirical correlation because the complexity of transient heat transfer phenomena in CFB.

The comparison shows a reasonable agreement in some parts and high deviation in others. The present experimental heat transfer coefficient values tend to increase with increasing of air velocity more than that shown by the correlation (Ehung model). Also the existing correlation did not deal with the variation of initial bed height and fluidization velocity but use the particle diameter instead because this model depending upon the surface renewal model where this model consider the particle convection (h_p) has larger part in heat transfer coefficient value. The deviations in values are due to the different experimental conditions adopted by different researchers and also to the strong anomalies present in the fluidization processes.

Conclusions

1- The heat transfer coefficient is directly proportional with time and initial bed height (or flow density) also , the change of fluidization velocity, but these coefficient inversely proportional with particle diameter.
2- Conversion difficulty the experimental values of heat transfer coefficient to empirical correlation by dimensionless group because the complexity of transient heat transfer phenomena in CFB.
3- The differences in the values of present Nusselt number profile and profiles results from another model at applied same the recent experimental conditions because the assumptions difference for each case.

NOMENCLATURE

Symbols	Description	unit
A_h	Heat transfer element outside surface area	m^2
Cp	Specific heat at constant pressure	J/kg.K
d_p	Average particle diameter	m
h	Total heat transfer coefficient	$W/m^2.K$
h_r	Radiation heat transfer coefficient	$W/m^2.K$
Hi	Initial bed height	m
I	Electrical Current	Amp.
k	Thermal conductivity	W/m.K
q	Power supplied = VI	W
R	Radius of column or (Riser)	m
t	Time	sec
T_b	Bed temperature	K
T_s	Surface Temperature	K
u	Superficial velocity	m/s
V	voltage across the heater	Volt
Greek Letters		
ε	Bed voidage	-
ρ_g	Density of gas	kg/m^3
ρ_s	Density of solid	kg/m^3
μ	Fluidizing gas viscosity	$N.s/m^2$
α	Thermal diffusivity	m^2/s
δ	Gas film thickness	m
Dimensionless Group		
Ar	Archimedes number $\left(\frac{d_p^3\rho_g(\rho_s-\rho_g)g}{\mu^2}\right)$	-
Nu	Particle Nusselt number $\left(\frac{h.d_p}{k_g}\right)$	-
Re	Reynolds number based on particle diameter $\left(\frac{\rho_g u d_p}{\mu}\right)$	-
Subscripts		
e	Effective	
g	Gas	
gc	Gas convection	
mf	Minimum fluidization	
p	Particle	
pc	Particle convection	
s	Solid	
b	Bed	
w	Wall	
cl	Cluster	

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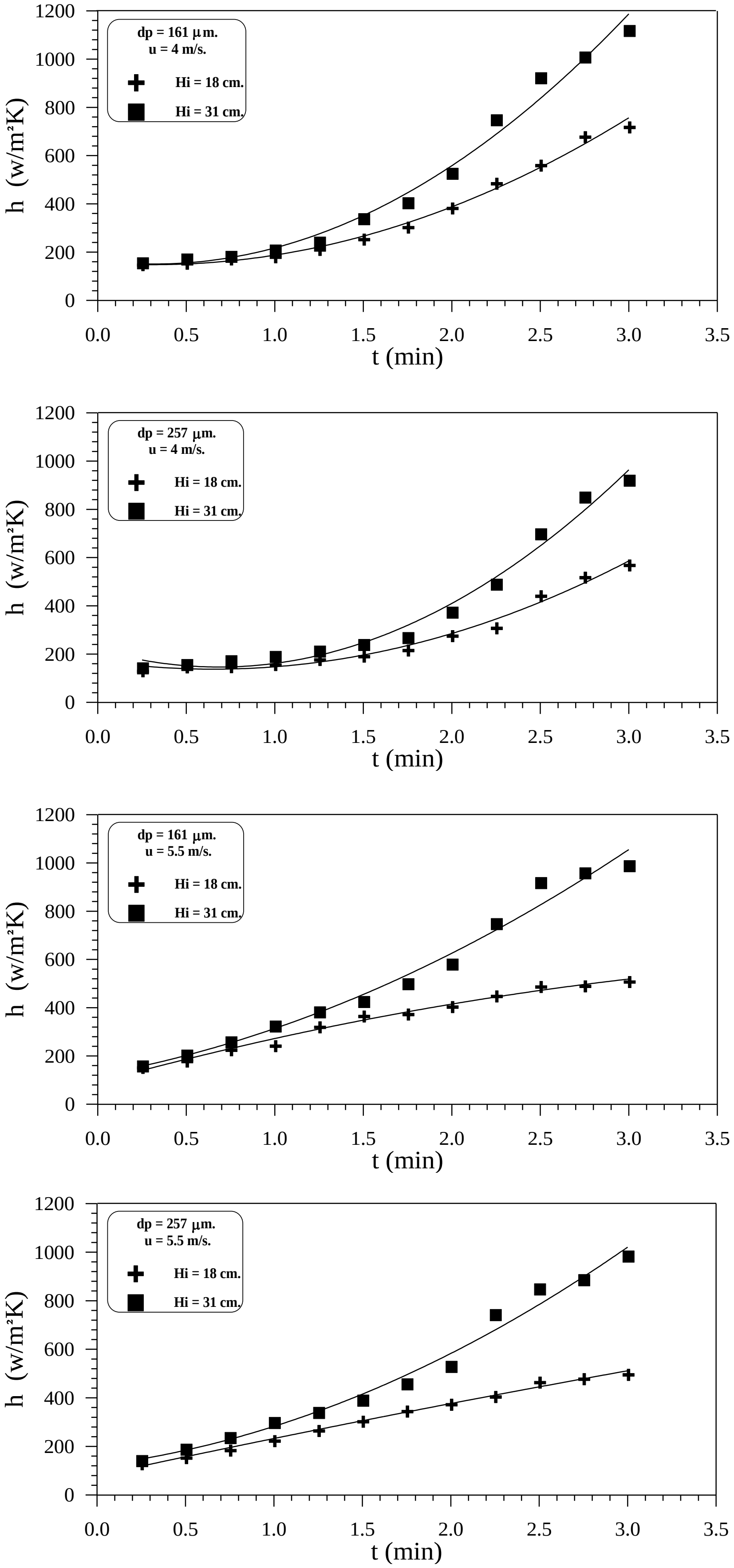

Figure (2): The Experimental Total Heat Transfer Coefficients Values as Function with Variables Operated Conditions.

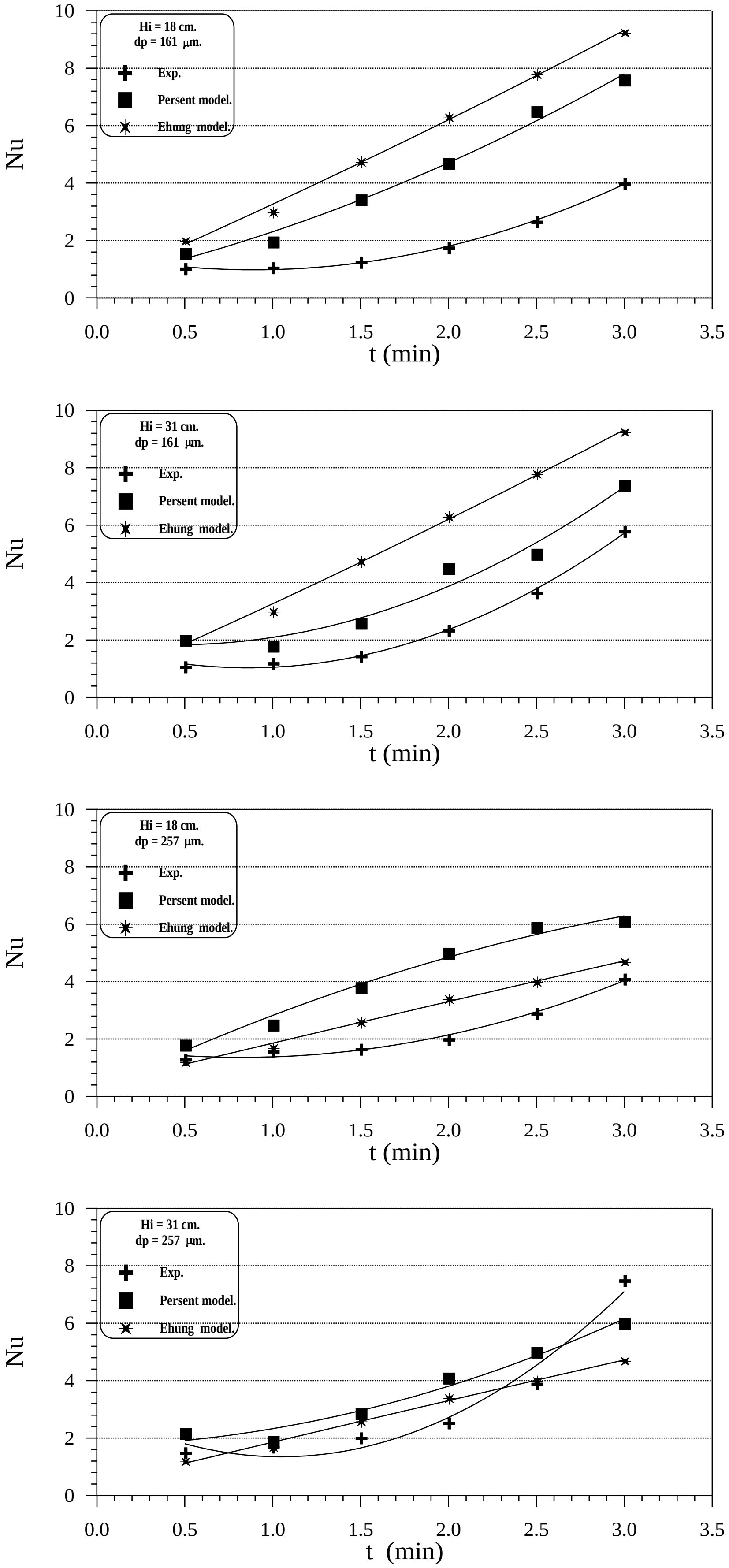

Figure (3): Comparison of Experimental and Computational Heat Transfer Coefficient with Ehang et.al(1993) Model, at u = 4 m/s .

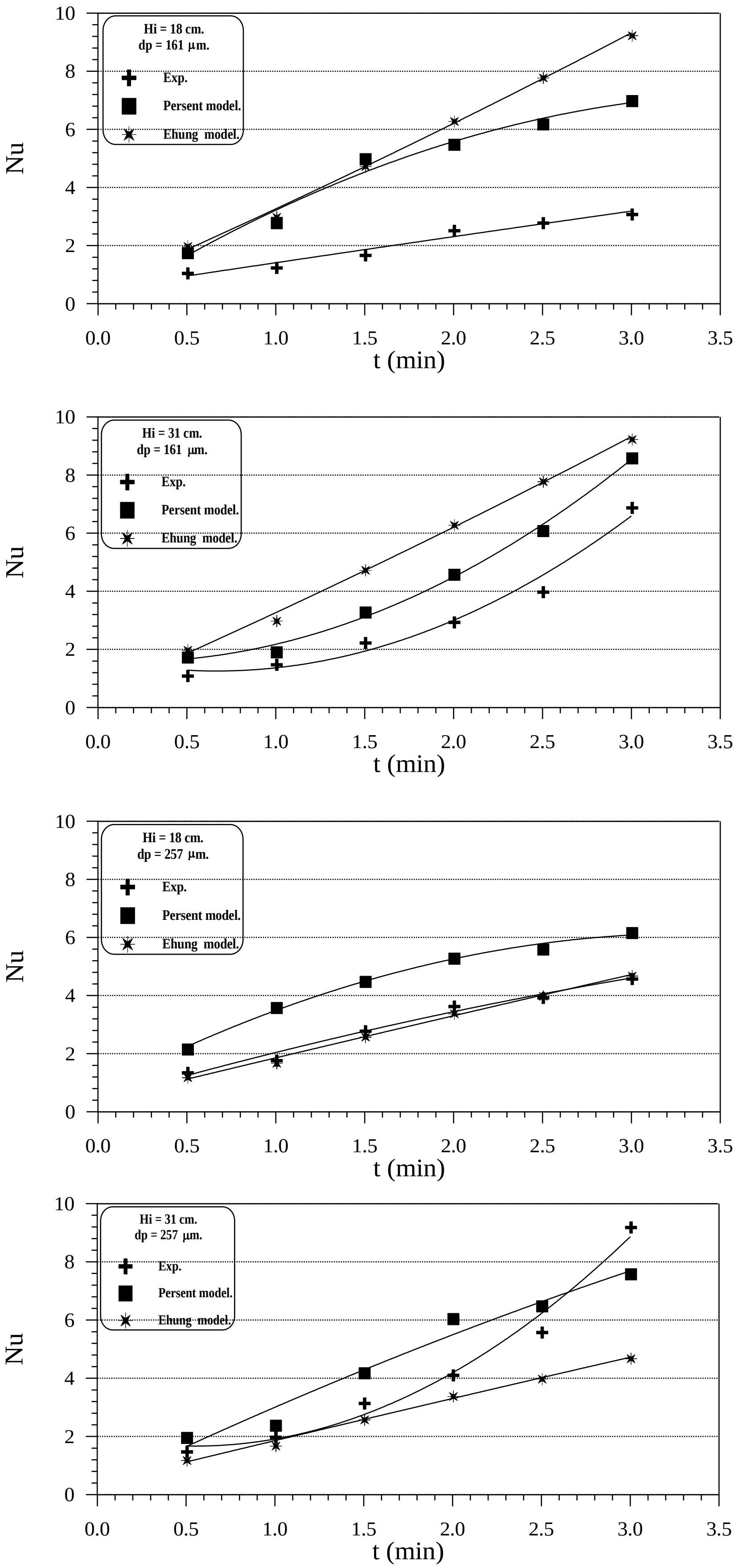

Figure (4): Comparison of Experimental and Computational Heat Transfer Coefficient with Ehang et.al(1993) Model, at u = 5.5 m/s .