EXPERIMENTAL STUDY OF THE EFFECT BED DENSITY ON HEAT TRANSFER COEFFICENTS IN TOWERS WITH GAS-SOLID CIRCULATING FLUIDIZED BED

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Abstract:

Experimental works were carried out in gas- solid CFBs to investigate the steady state heat transfer between gas and solid and the surface immersed in the bed. The bed column was 76 mm in inside diameter and 1500 mm in height fitted with a horizontal heating tube with outer diameter 28 mm heated electrically with different power supplies (105 W). The fluidizing medium was air at different velocities (4.97, 5.56 and 6 m/s). Three different size of sand particle were employed (i.e 194, 295 and 356 μm). The initial bed height used experimentally with different values (15, 25 and 35 cm). The column of heat exchange provide with return line which content of the cyclone to separated the sand with type is High efficiency cyclone connected to standpipe to transport the separated sand to riser mixing region inside the riser. Heat transfer coefficients are found to increase with fluidized air velocity and, through clear, with heat flux, but, they show an inverse dependence on particle size, and direct proportional with initial bed height which representation the bed density.

KEYWORDS: CFB, Two phase flow Heat exchange, Fluidization, Gas-solid fluidization

دراسة تأثير كثافة الجريان المتميع على معاملات انتقال الحرارة في الأبراج ذات الطبقات المميعة المدورة (غاز– صلب).

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الخلاصة

في هذا البحث تم أجراء تجارب عملية في انتقال الحرارة بالحالة ألمستقره للطبقة المميعة وكان عمود التميع بـألا بعاد ٧٦ ملم قطراً داخليا و ١٥٠٠ ملم أرتفاعاً ومجهز بمسخن أنبوبي أفقي مغمور داخل ألحشوه ذات قطر خارجي ٢٨ ملم يسخن كهربائيا ، جهز بقدرة كهربائية (١٠٥ واط). وقد استخدمت دقائق الرمل كحشوه وبثلاث أقطار مختلفة (١٩٤، ٢٩٥، ٣٥٦ مايكرو متر). وحشوات الرمل المستخدم بارتفاعات أولية هي (١٥ ، ٢٥، ٣٥ سم). أستخدم الهواء كوسط للتميع وبثلاث سرع مختلفة (٤.٩٧ ، ٥.٥٦ ، ٦ م/ثا). كذلك أستخدم خط الإرجاع للمواد ألصلبه التي يتم ترسيبها داخل منظومة

ترسيب مخروطيه من نوع (High Efficiency Cyclone)، ومن ثم أعادتها الى منظومة التبادل الحراري عبـــر خـــط (Standpipe). وجدت معاملات انتقال الحرارة تزداد مع زيادة سرع التميع وكذلك مع زيادة معدلات ســريان الحـــرارة (Heat Flux)، وتتناسب عكسيا مع حجم الدقائق. تم دراسة تأثير تغير ارتفاع الحشوة على معاملات انتقال الحرارة حيــث وجــد أن زيادة ارتفاع الحشوة الابتدائية سوف يؤدي بشكل متناسب الى زيادة معاملات انتقال الحرارة، كونه يمثل زيادة كثافة الجريـــان وتقليل النفاذيه.

Nomenclature

Symbols	Description	unit
A_h	Heater surface area	m^2
d_p	Average particle diameter	m
Hi	Initial bed height	m
h_L	local heat transfer coefficient	$W/m^2.K$
q	Power supplied = VI	W
T_b	Bed temperature	K
T_s	Surface Temperature	K
u	Superficial velocity	m/s
u_{mf}	Minimum fluidization velocity	m/s
z	Axial distance in riser column	m
ε	Fluidization porosity(or bed voidage)	-

Introduction

Fluidization is the operation by which fine solid transferred into a fluid-like state through contact with gas or liquid. This method of contacting has a number of unusual characteristics , and fluidization engineering is concerned with efforts to take advantage of this behavior and put it to good use .The first major successful application of gas– fluidized bed techniques, however, was to the engineering of the catalytic cracking process (Botterill, 1975).

In a fluidized bed ,there exist favorable conditions for rapid heat and mass transfer between the solids and the fluid , very rapid mixing of the solid generally occurs and the coefficients for the transfer of heat to boundary surface are very high . Hence fluidized beds are used both as heat exchangers and as chemical reactors particulate where close control of temperature is required and where a large amount of heat must be added to or removed from the system . In all fluidized bed applications , heat must be transferred to or from the fluidized bed , thus heat transfer in gas- fluidized beds is an important operation in the chemical process industries (Kunii 1977).

Circulating fluidized beds (CFB) are gas-solid contacting systems that operate at high gas velocity . Although their design varies according to the application ,the typical element of such unit can be seen in **Figure(1)**. A CFB is constituted of a riser , a cyclone , a standpipe and a solid injection system . Solids at the bottom of the riser lifted upward by the riser gas . Upon exiting the riser, the gas and solids are separated by the cyclone located at the top of riser. The separated solid fall into the standpipe or downcomer, where they flow into the downcomer and are subsequently re-injected into the riser. To help the smooth circulating of solid, gas is generally added at the solid injection valve (secondary air duct) to drag the solid into the riser . In this way, the gas velocity and solid mass flux in the riser are independent of each other. Globally, there is a pressure balance established in a CFB. The pressure drop in the riser balance the head of solid as well as the transfer of energy to the upward flowing particles . This is also an important pressure drop across the solid injection valve and in the cyclone. The pressure gain in the dowencomer balances pressure drops (Basu, p. and Nag, P. K., 1996).

The circulating fluidized bed operate at gas velocities at which significant solid entrainment occur. While and Cankurt (1978) defined the CFB as a fluidized bed operating at gas velocities exceeding the minimum transport velocity, a more recent definition by Berruit et al.(1995) says that a CFB can be defined as any system involving solid recirculation and operation at gas velocities exceeding transition velocity from bubbling /slugging to turbulent fluidization (u_c).Thus a CFB may be operated at gas velocities higher than those used for bubbling fluidized beds ,i.e. $(u > 30u_{mf})$,where the flow successively enter the turbulent fluidization , fast fluidization and pneumatic fluidization regimes (levenspiel,1999).

Literature Survey

Many investigators have explored heat transfer in CFB furnaces using bench-scale units. These units are operated either at wide rang of temperatures from room temperature (20-60^{o}C) to combustion (600-900 ^{o}C) temperatures. Heat transferring surfaces ranged from small probes (10 mm) diameter to long panels(1500 mm long).

The earliest study on heat transfer to fluidized beds where solid exiting the column recycled to its base was made by Mickley and Trilling (1949). They conducted numerous experiments using different superficial gas velocities and particle size. A strong effect of particle size may be a reflection of very dense bed (> 300 kg/m^3) close to bubbling used in their experiments.

Information about heat transfer to the horizontal cylinder immersed in the gas-solid suspension of a circulating fluidized bed is scarce in the open literature. The review paper by Saxena *et al.*(1992) has summarized most of the limited amount of pervious work. Nag *et al.*(1992) reached a heat transfer coefficient of 450W/m^{2}K for a suspension density of 70 kg/m^3 using sand , and investigated the effect of fins on the overall heat transfer. More work has done concerning heat transfer to vertical cylinder [e.g. Ahn *et al*. (1997)]. Experiments carried out on a bubbling bed, can be of some interest to this work as the freeboard region of fluidized bed is analogous to dilute flow in a CFB riser.

Biyikli *et al.*(1983) investigated the heat transfer coefficient in the freeboard and they showed the variation of the heat transfer coefficient along the surface of the probe. The heat transfer coefficient was shown to exhibit a maximum on the sides of the probe and a minimum at the bottom.

Grace (1986) reviewed the surface to suspension heat transfer coefficients in the CFB. Two approaches, a single particle bombardment model and a finite pocket model, were discussed. It was concluded that surface to suspension heat transfer coefficients increased strongly with increasing solid concentration and decreases with increases particle diameter. The heat transfer coefficient was also found to vary inversely positions along the height of the column.

Furchi , *at el*,(1988) measured local heat transfer coefficient in a 6 m high, 72 mm ID riser. The phase consisted of propane combustion products with very high excess air temperature up to 250^{o}C. The solid particles were glass spheres with diameter of 109, 196 and 269 μm. The heat transfer coefficients were measured by using six water jackets attached along the height of the riser. Superficial gas velocity were maintained in the 5.8 m/s to 12.8 m/s range, and solid circulation rate were maintained between 0 to 80 kg/m^{2}s. Results indicated that heat transfer coefficients ranged between 30 to 150W/m^2 K. Their values increased as solid circulation rates were increased and decreased as the particle increased. Gas velocities were found to have a weak effect on the heat transfer coefficient. The local heat transfer coefficient when plotted versus column height reached a maximum values at 1.5 m and then decreased. The value of heat transfer coefficient at 2.5 m and above appeared to be more or less constant. The only exception occurred in the case of the 269 μm particles where a slight increase in the heat transfer coefficient was observed near the top of the pipe tested.

Poolpol (1992) measured the wall-to-bed heat transfer coefficients in the cold model, using bed material made of 250μm silica sand. In this study, three different sizes of silica sand were used to observe the effect of particle size on heat transfer coefficients. Also, the geometry of the riser was modified to include a small diameter section under the heater wall section and a flat top exit to the primary cyclone. This geometry is common in commercial CFB units in which a lower section of the riser, where the main combustion reaction take place, is constricted and covered by a thick layer of refractory. A flat top exit is also preferable as it enhances the internal recirculation of solid particles and heat transfer to the water walls.

Grace (1997) reviewed the effect of equipment geometry on particle and gas flow in CFB riser. Items discussed included the influence of riser diameter, gas distributor design, solid re-entry and feeding configuration and level , secondary air entry the level, conical bottom section , corner , wall roughness, and exit geometry. Grace included that the exit geometry strongly influenced hydrodynamics and heat transfer to a wall or immersed surface. Its effect was fell not only at the top but for a considerable distance back down the riser. The measured local solid concentrations 0.4 m below the bottom of exit for both an abrupt and smooth exit were also examined by Senior (1992). Senior found that at a superficial gas velocity of 6.5 m/s and a solid circulating rate of 60 kg/m^2 s, local solid concentrations next to the riser wall with the abrupt exit ranged from 5 to 10 %, while they only ranged from 1 to 3% next to the riser wall with the smooth exit.

Experimental Procedure

Experiments were carried out by the experimental rig shown in **Figure(2)** and a photograph for whole experimental set up is shown in **Figure (3).** to show the effect of different operation conditions on the heat transfer in the CFB. Such condition are the superficial air velocity(main pipe air velocity),mean particle size, initial bed height, and heat input, where all experimented operations takes place at constant air velocity second air line which used to push the separated sand to mixing region in the riser ,the velocity is (5.7133 m/s)with volume flow rate of (0.007788 m^3/s). The selected experimental values are presented in **Table(1).**

Results and Discussion

The heat transfer coefficient investigated in this section. The local heat transfer coefficient is calculated on the basis of the local bed temperature, according to the following equation:

$$h = \frac{q}{A_h (T_s - T_b)} \qquad (1)$$

A set of typical results can be seen in **Figures. 3 to 5** showing the effect of the superficial gas velocity on the value of local heat transfer coefficient for different axial distances along the riser. The figures show decreasing in local heat transfer coefficient values as axial distance increases.

The local heat transfer coefficient are presented as a function of fluidizing velocity in Figures. 3 to5. The qualitative variation of the surface-to-bed heat transfer coefficient on fluidizing velocity in each case is in agreement with reported trends observed by various investigators. The value of heat transfer coefficient increases with increasing fluidizing velocity. The rate of increase of heat transfer coefficient is larger at low fluidizing velocities than at high velocities, because the rate of replacement of solid in the vicinity of the surface rises, and particle residual time at the heat transfer surface decreases. However, the particle density (1-ε) close to heat transfer surface decreases with further increases in fluidizing velocity. This increase in heat transfer coefficient is due to the effect caused by an increase in the initial bed height or flow density.

From same Figures below it is clearly noticed that the heat transfer coefficient decreases with an increase in solid particle sizes. It is noticeable that the curve for the small particle size (i.e. 194 μm) is higher that those of the larger particle sizes. All the curves have the same trend which is in agreement with reported finding in the literature (Grewal, 1981; Saxena, 1989). This trend is expected since smaller particle size has higher contacting frequency on the heat transfer surface and reducing the particle will reduced the effective gas path (or gas film) between immersed surface and particles which is dependant upon the mean particle diameter (calculated as 0.1 d_p as by Wu *et al*, 1990). The decrease in the gas path decreases the resistance to heat flow. In addition, the particle surface area per unit volume of the bed is larger for smaller particles, therefore small particles are more efficient in exchanging heat with the surface. This effect agrees well with Ying L.M, (1998) who found that the heat transfer coefficient was approximately inversely proportional to mean solid particle diameter raised the power (1.2). The shape and surface state of the solid particles influenced the heat transfer rate; heat transfer coefficient is somewhat higher for rounded and smother particles (Botterill, 1975).

Conclusions

1- The axial heat transfer coefficient profile are clearly recognized with large values of initial bed heights (bed flow density) and low gas velocities. At small initial bed heights and high gas velocities, the beds seem to be homogenous with nearly no temperature profile exists.

2-The particle size of the solid materials used in fluidization have an important effect on the heat transfer coefficient. Best heat transfer coefficient values can be obtained by using small particle size of solid material, which its effect on bed voidage fraction and bed density.

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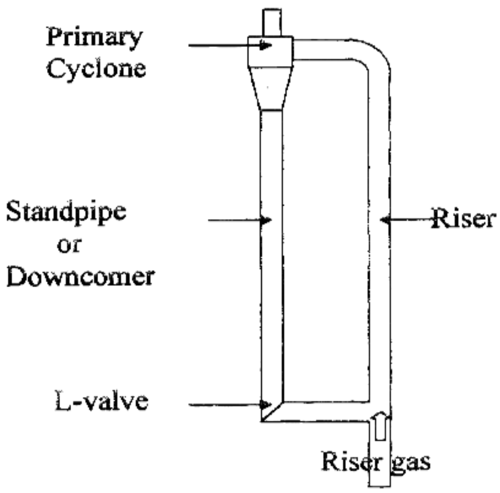
Table (1): The values of Operation Conditions used in Experimental.

Operation conditions	First value	Second value	Third value
Superficial air velocity, u (m/s)	4.97	5.56	6
Mean particle size, d_p (μm)	194	295	356
Initial bed height , Hi (cm)	15	25	35
Heat flux input , q (W)	50	105	190



Secondary Air Flow

Main Air Flow

Figure (1): A Simplified Representation of a Circulating Fluidized Beds.

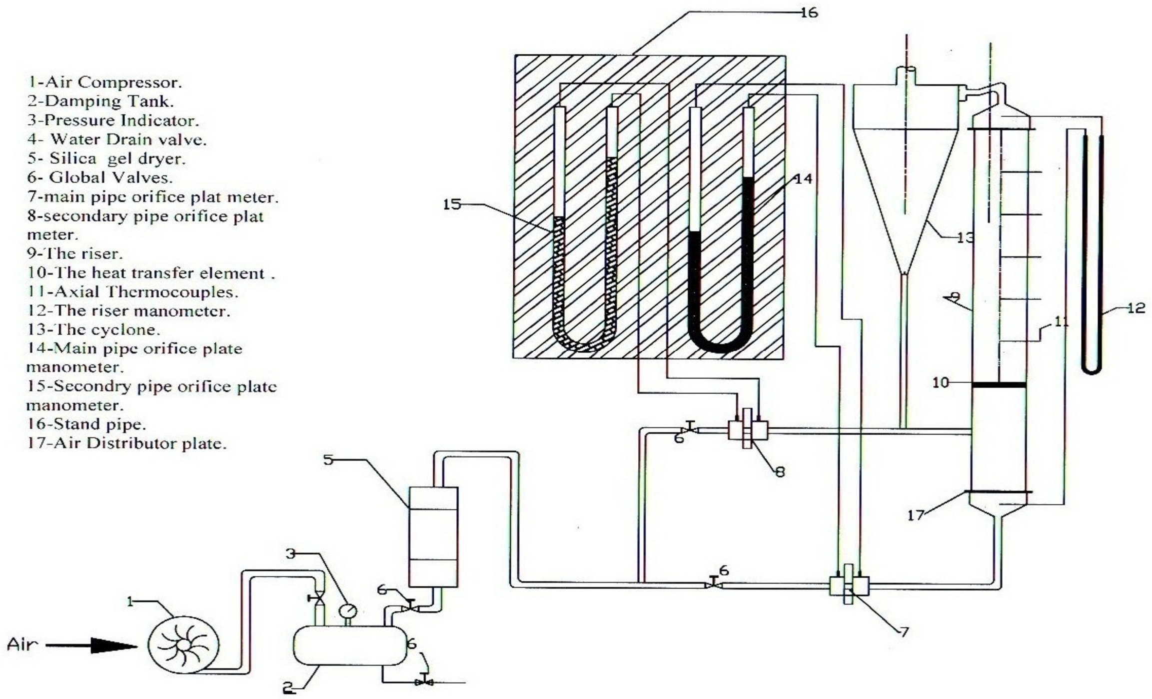


Figure (2): Schematic Diagram of the Experimental Setup.

Figure (3): General View of Experimental Rig.

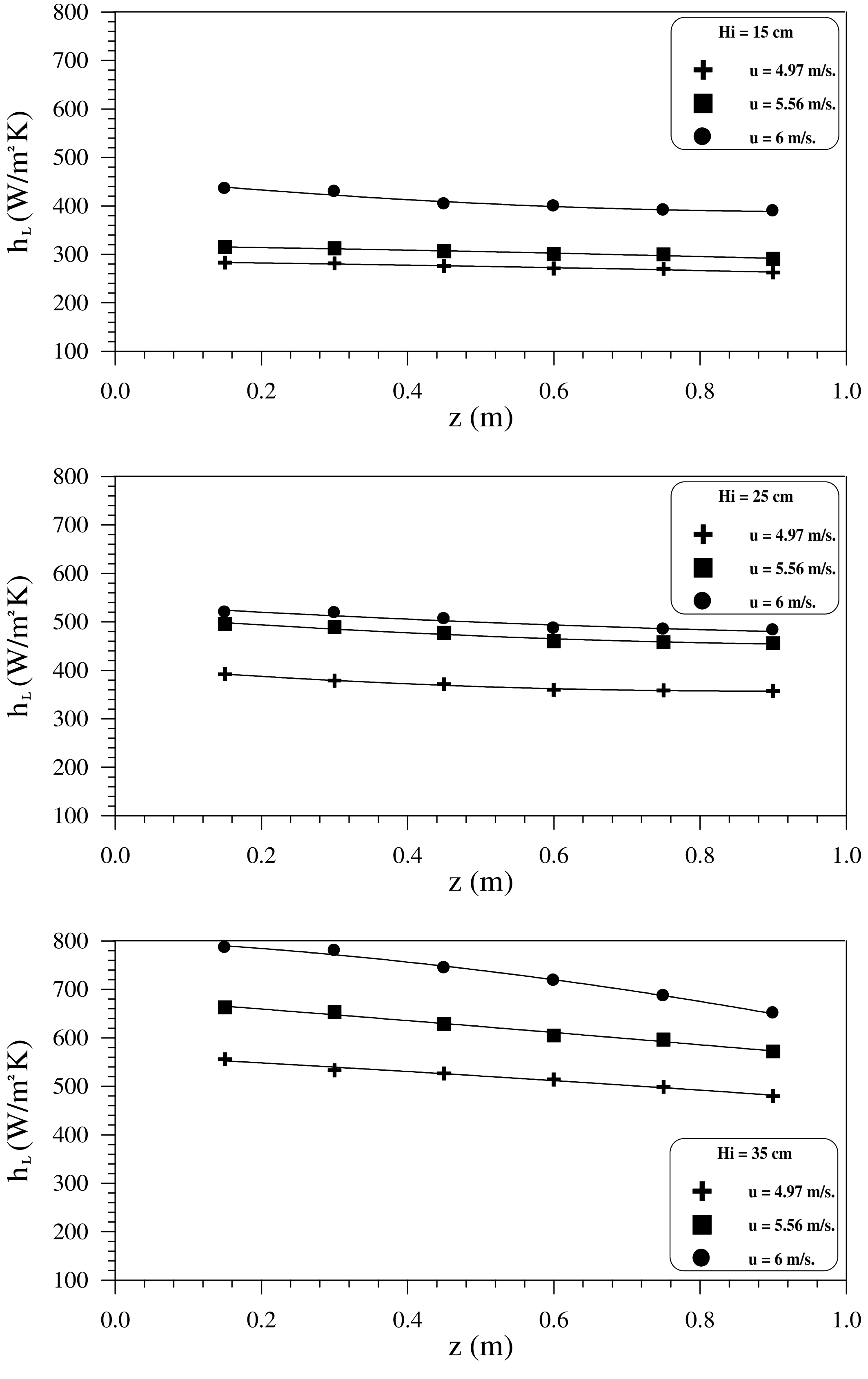


Figure (4): Local Heat transfer Coefficient Distribution.

$q = 105$ W, $d_p = 194$ μm.

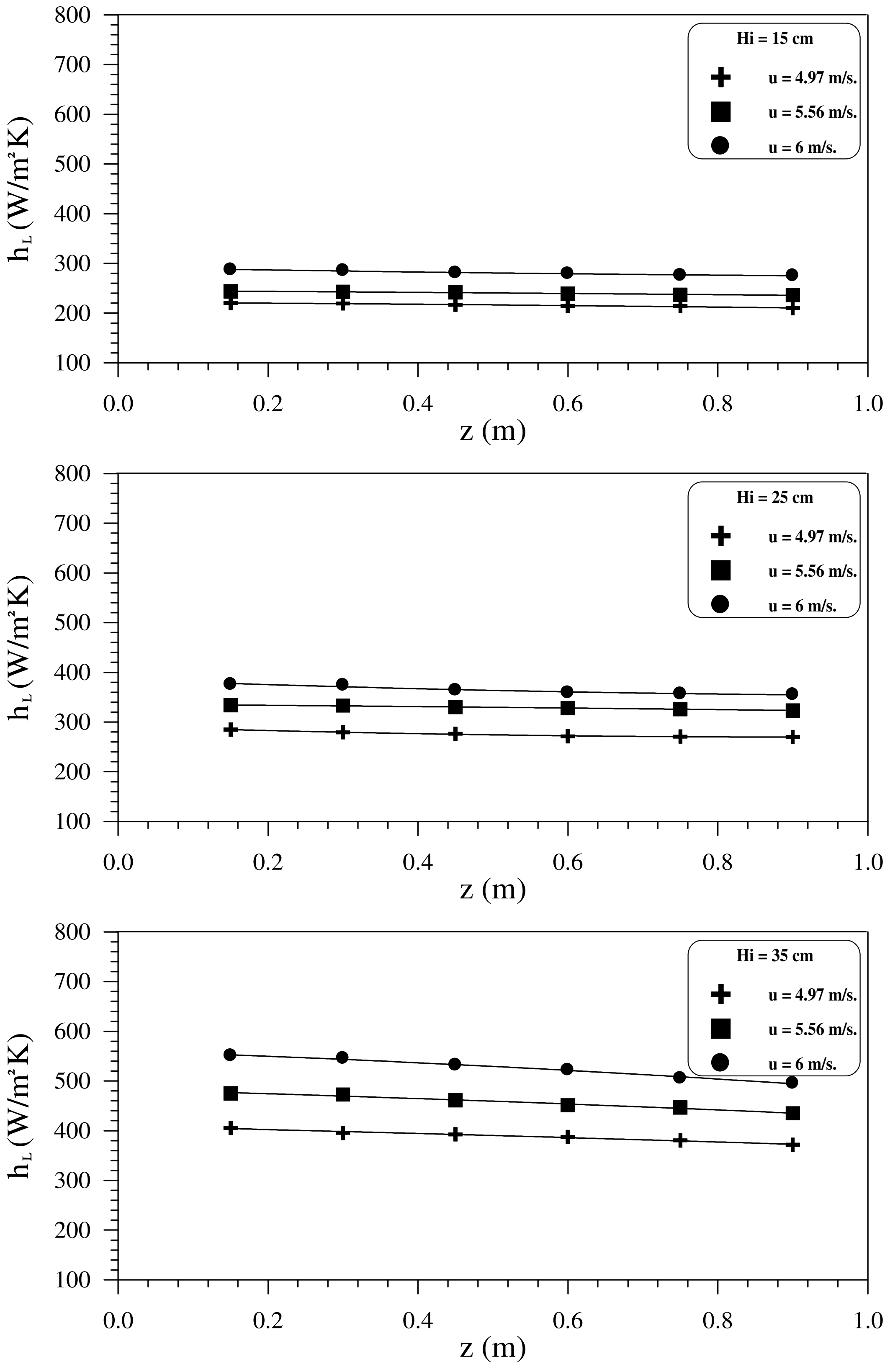


Figure (5): Local Heat transfer Coefficient Distribution.

$q = 105$ W, $d_p = 295$ μm.

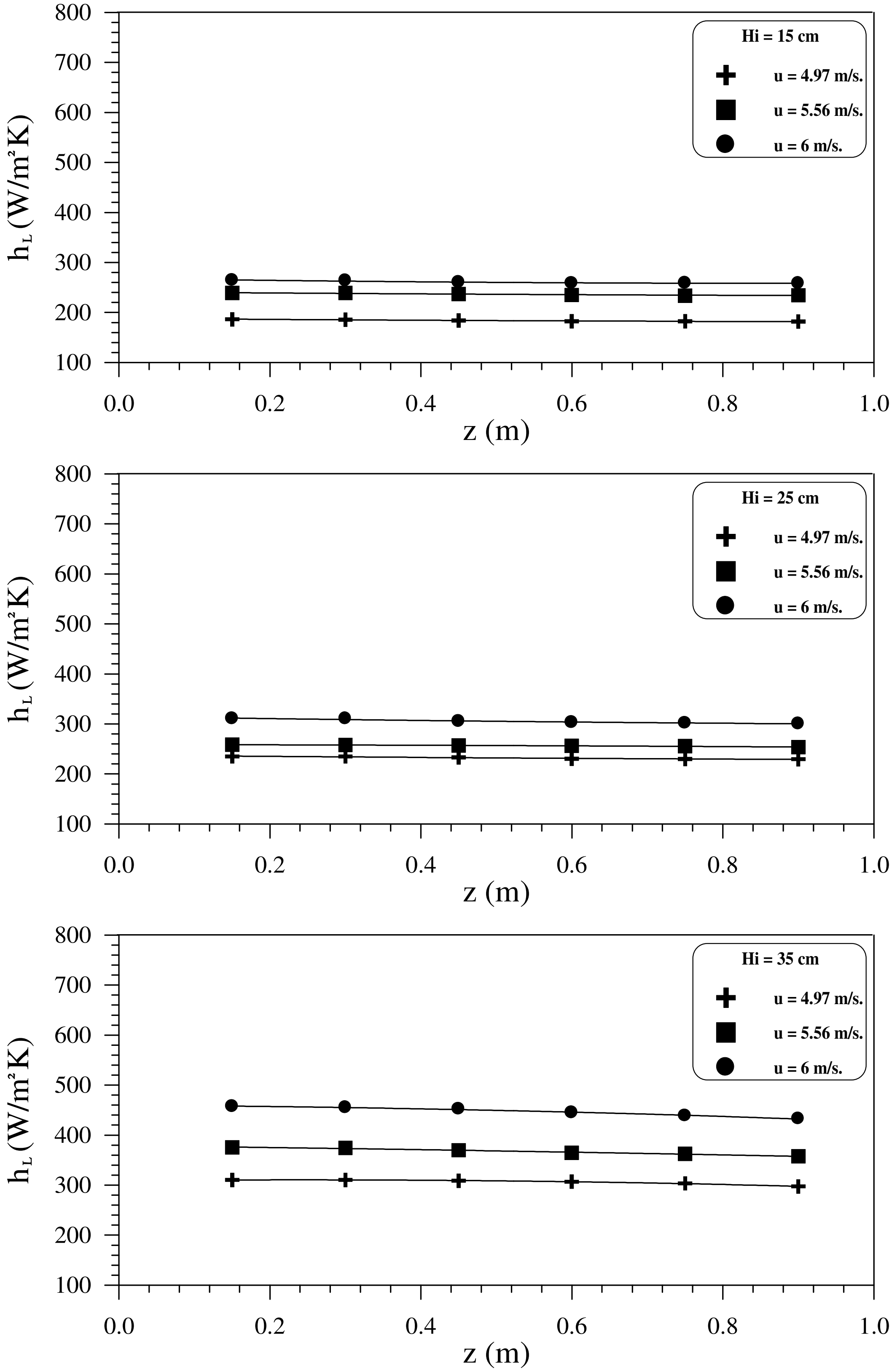


Figure (6): Local Heat transfer Coefficient Distribution.

$q = 105$ W, $d_p = 356$ μm