Investigation of Heat Transfer Parameters of a Bundle of Heaters in a Simple Bubble Column Reactor Using CFD Method

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Abstract

Bubble columns are gas- liquid contactors that are widely used in chemical and bio- chemical industries. High mixing that result in high heat and mass transfer rates are amongst their advantages. Heat transfer in a bubble column having a bundle of heaters investigated and the variation of heat transfer coefficient with variation in heaters pitch to diameter ratios in a bundle of heaters reported. The bubble column modeled in a three-dimensional in the Eulerian framework by using computational fluid dynamic simulation. It was 0.292 m in diameter and 1.5 m in height. Simulations performed for five heaters and for nine heaters in a bundle. The bubble column with a five heaters bundle was investigated in the pitch to diameter ratios of 1.5, 2 and 2.5. The bubble column with nine heaters bundle was investigated with the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46. The range of the gas velocity was 0.0025 to 0.04 m/s.

The heat transfer results show that, with decreasing the pitch to diameter ratios the heat transfer coefficient decreases. In bubble column with bundle of nine heaters at a constant superficial gas velocity, with decreasing the pitch to diameter ratio from 2.46 to 1.5 the heat transfer coefficient decreases; whereas in the pitch to diameter ratio of 1.25 heat transfer coefficient is high.

Keywords: Bubble column, Bundle of heaters, CFD, Heat transfer, Hydrodynamics, Pitch to diameter ratios

Introduction

Bubble Columns, BCs, are widely employed gas-liquid contactors as reactors in chemical, biochemical and waste treatment industries. The high applicability of bubble columns are due to;1) the absence of moving parts that leads to the ease of construction and operation ,2) the high mixing effect that leads to high gas- liquid interfacial area and 3)the relatively high rate of heat and mass transfer [1,2]. However, their applications as absorbers or strippers due to their high back mixing have been limited. Nevertheless their use for removal of CO_2 for the goal of reducing greenhouse effect has been reported [3].

A simple bubble column reactor is commonly a cylindrical vessel filled with a liquid up to a certain level. The gas phase is continuously dispersed in the batch liquid phase through a gas distributer which is placed at the lower part of the column. Non-uniform distribution of gas in the column causes radial and axial mixture density variation that leads to the production of large liquid circulation [4]. Depending on the gas superficial velocity three kinds of flow regime in bubble column that are homogeneous (bubbly), heterogeneous (churn turbulent) and slug flow, are recognizable. In homogeneous flow regime, the bubbles are almost uniform and their movements are distinct. This regime is observed in superficial gas velocities less than 5 cm/s for air-water systems. At superficial gas velocities greater than 5 cm/s the heterogeneous flow regime is prevailed where, bubbles coalescence and break-up rates [4,5,6] are high and the bubble size distribution is not uniform. Hence, design parameters such as gas holdup, heat and mass transfer coefficients, effective gas-liquid interfacial area and bubble size distribution depend on superficial gas velocity [7, 8].

Thermal control of endothermic or exothermic reactions necessitate the use of internal heat exchanger, which requires knowledge of heat transfer from an

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immersed object to the gas- liquid mixture or the use of external heat exchanger, which requires knowledge of heat transfer from gas- liquid mixture to the wall [5, 9]. It is understood that heat transfer coefficient depends on the bubble column hydrodynamics [4, 10].

Heat transfer in a simple gas- liquid bubble column reactor depends on the superficial gas velocity, liquid viscosity, axial and radial positions of the heat transfer probe and column dimensions. Studies showed that heat transfer coefficient increases with the increase of temperature, superficial gas velocity and in case of present solids it depends on the particles size. Increase of fluid viscosity and particle density decreases heat transfer coefficient [3, 11, 12]. Even though the increase of heat transfer coefficient with increasing solid concentration was reported [11, 12], its contradiction was also mentioned [13, 14]. Internal heat exchangers affect liquid circulation pattern of bubble columns. It was realized that insertion of the tubes bundle in BC has an intense effect on the decrease of the liquid turbulence kinetic energy [15-17].

Determination of heat transfer coefficient bubble column with a five tubes bundle of heater were performed [10]. The heaters pitch and the heat fluxes were 3.65 cm and 468, 607, 813 and 943 W, respectively. Superficial gas velocity was in range of 0.04 m/s to 0.42 m/s. The increase of heat transfer coefficient with increasing in gas superficial velocity up to 0.26 cm/s reported. But, with further increase of superficial gas velocity heat transfer coefficient remained constant and it is independent of the heat flux value. Rahimi and Tayebi [9] investigated the effect of size and exposure of a heater in a simple bubble column by CFD method. They showed that only vertical dimension of heater has an affect on the heat transfer coefficient. That finding was previously experimentally observed by Lewis *et al*. [18]. One should realize that heat transfer coefficient correlations for tall bubble columns are unable to predict heat transfer in shallow bubble columns, which their geometry and fluid dynamics are unique [19].

Heat transfer coefficient and gas holdup measurements for three-phase system in bubble column showed that the heat transfer coefficient depends on the superficial gas velocity as well as the solid concentration, and is independent of the particle size. Further it was concluded that heat transfer coefficient of a three- phase BC is higher than the heat transfer coefficient of two-phase BC. Heat transfer coefficient for an immersed seven tubes bundle in a slurry reactor that contained soft powders of iron oxide was reported to be higher than that of a single tube heat exchanger [16,17]. Schluter *et al*. [17] investigated heat transfer of the longitudinal and cross flow 890 tube bundles in two and three phase system (air-water-propylene glycol) bubble column with internals. It was reported that heat transfer coefficient of highly viscous liquid is less than the heat transfer coefficient of low viscous liquids and the tube pitch effect on a high viscous liquid is low. Also they compared seven of the fifteen correlations for estimation of heat transfer coefficients to report their wide predictions.

Lack of reported data on the heat transfer coefficient of bubble column and even contradiction of reported data justifies further study of heat transfer in bubble columns. Recent studies show the influence of design of internals on the column hydrodynamics and on the heat transfer coefficients, therefore proper design and modeling relies on experimentations [20, 22].

Availability of powerful computers has led to develop and improve the tools in field of CFD modeling of bubble columns [24-27]. Hence, in this article a CFD model developed and validated with the available experimental data. Further an insight is given to the hydrodynamics of bubble column. A design correlation for the heat transfer coefficient estimation as a function of number of tubes in a bundle and pitch to tube diameter ratio is provided.

2. Experimental

Experimental data that were obtained for heat transfer in a bubble column with a nine and with 5 heaters bundles [28] were employed to validate the CFD model. Data were obtained from 0.292 m in diameter and 1.5 m in height bubble column. Water and air were used as liquid and gas phases. Dispersed air temperature was 18 °C. Initial height of water was considered to be 0.9 m. Each heater was 60 mm in height and 30 mm in diameter. Heaters were located at distance of 0.75 m from gas distributor. Heat transfer coefficient was measured at the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46, and in range of superficial gas velocities of 0.01 to 0.04 m/s. The heat supply from each heater was 3kW[28].

3. CFD Model

Heat transfer was investigated in a 0.292 m in diameter and 1.5 m in height bubble column. Two internal heat exchangers were used. Bubble column with 5 heaters was investigated at the pitch to diameter ratios of 1.5, 2 and 2.5. Bubble column with 9 heaters was investigated at the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46. Nine heaters bundle arranged with square pitch in three rows and three columns. Tables (1) and (2) show the pitch to diameter ratios for bubble columns with five and with nine heaters, respectively. Air and water were used as gas and liquid phases, respectively.

Table 1: Pitch to diameter ratios for bubble column with five heaters

Pitch (m)	Pitch to diameter ratio
0.045	1.5
0.060	2
0.075	2.5

Table 2: Pitch to diameter ratios for bubble column with nine heaters

Pitch (m)	Pitch to diameter ratio
0.0375	1.25
0.045	1.5
0.0675	2.25
0.0738	2.46

Gas phase was dispersed in water by sieve plate sparger with 129 holes of 1mm in diameter. Initial height of water was considered to be 1m. Heat transfer coefficient is independent of liquid initial height [28, 23]. Rahimi (1988) [28] investigated the effect of water height on the heat transfer in BC. As the effect of liquid height in the range of 0.9 m and 1.2 m on the heat transfer coefficient was not confirmed experimentally taking a 1m liquid height for modeling is justifiable.

Heaters were located 0.65 m above gas distributor to eliminate the gas distributor effect. Heaters were 60 mm and 30 mm in diameter. The water and air temperatures were 25 °C. Superficial gas velocities of less than 0.04 cm/s assure bubbly flow regime.

In homogeneous regime the bubble diameters are almost equal in the range of 1 to 7 mm [9]. Therefore, the average bubble diameter in this work was set to 5 mm.

The simulation has been done under transient condition. The initial values that are required for an unsteady run were produced by an initial steady state run so that its procedure reduces computing time substantially. The convergence criteria that simulation reached to steady state were 0.0001. For solver control high resolution scheme was used. The second order backward Euler method was used to solve equations numerically. This scheme for volume fraction calculation was second order [24] whereas for solving turbulent equation it was first order.

The total number of unstructured nodes that were used for the bubble column with five heaters and for the bubble column with nine heaters was 205510 and 302242, respectively. The column mesh structure and its geometries are shown in Figures1 and 2.

A pressure boundary condition was applied to the top of the column. No slip and free slip boundary condition for liquid and gas phases assumed, respectively. A CFX 11.0 software on three PC's with 4 GHz CPU and 4 GB RAM were used to solve the equations for the two fluid mixtures.

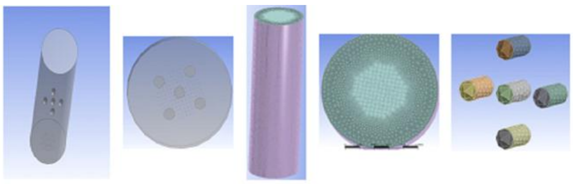


Fig. 1

Figure 1: Geometry and column mesh for bubble column with five heaters

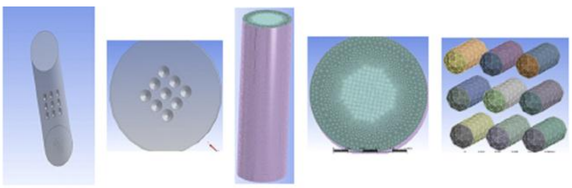


Fig. 2

Figure 2: Geometry and column mesh for bubble column with nine heaters.

3.1. Model equations

The gas and liquid phases in this work were modeled in the Eulerian- Eulerian frameworks. The dispered gas and the continous liquid phases currents were assumed to be laminar and turbulent, respectively. The continuity, momentum and energy equations were solved simultanously. The set of governing equations and model assumptions for unsteady state, incompressible, fully developed two-phase flow in bubble column in Cartesian coordinate system is represented by the following equations of change [24]:

Continuity equation:

$$\frac{\partial(\omega_k \rho_k)}{\partial t} + \nabla.(\omega_k \rho_k u_k) = 0 \qquad (1)$$

where ρ, u and ω_k are density, velocity and volume fraction of the k^{th} phase(gas or liquid), respectively.

Momentum equation:

$$\frac{\partial(\omega \rho u)_k}{\partial(t)} + \nabla\left(\rho_k \omega_k u_k - \mu_k \omega_k (\nabla u_k)^T\right) = -\omega \nabla P + M_k + \rho_k g \qquad (2)$$

P is the pressure, μ_k , viscosity of the k^{th} phase, M_{kl} , interfacial momentum exchange between phase *k* and phase *l*. ω_c is the volume fraction of continuse phase. One should notice that , $M_{kl} = M_{lk}$. Interfacial momentum exchange is given by Eq. 3.

$$M_{kl} = \omega_c F_D + \omega_c F_V + \omega_c F_L + \omega_c F_T + \omega_c F_W \qquad (3)$$

where, F_D, F_V, F_L, F_T and F_W are drag, virtual, lift, turbulence dispersion and wall lubrication forces, respectivly. Those forces but the drag force, F_D , are negligible. Therefore; $M_{kl}=\omega_c F_D$ [25].

$$M_{kl} = \left[\frac{3}{4}\frac{C_D}{d_b}\rho_l\right]\omega_c (U_G - U_L)|U_G - U_L| \qquad (4)$$

where the subscript L denotes the continuous liquid phase and G denotes the disperse gas phase. $|U_G - U_L|$ is slip velocity between the gas bubbles and the liquid phase. d_b is bubble diameter and ρ_l is density of the liquid phase.

The drag coefficient, C_D, depends on the flow regime and the properties of the continuous phase. The interphase drag coefficient was calculated by using Schiller Nauman'srelation, Eq. 5. This model originaly has been developed for solid spherical particles or fluid. It could be employed when bubbles are small enogh to be assumed as rigid spherical particles [25].

$$C_D = \frac{24}{\mathrm{Re}}(1 + 0.15\,\mathrm{Re}^{0.687}) \tag{5}$$

Energy equation:

$$\frac{\partial}{\partial t}(\omega_k \rho_k h_k) + \nabla \cdot (\omega_k \rho_k U_k h_k) = -\nabla \cdot q + (\tau_k . \nabla U_k) + \omega_k \frac{Dp_k}{Dt} + Q_{lk} + S_k \tag{6}$$

where h_k the specific entalpy is defined by Eq. 7 and S_k is entalpy source.

$$h_{ki} = \int_{Tref}^{T} C_{pki} dT \qquad h_k = \sum_{i=1}^{c} m_{ki} h_{ki} \tag{7}$$

C_{pk} is the specific heat of species k, at constant pressure. q is a flux of enthalpy andcan be written in terms of temperature gradient asEq. 8 [25]:

$$q = -k\nabla T \tag{8}$$

Q_{lk} is the energy transfer between the lth and kth phase. Heat transfer between the phases should satisfy the heat balance condition. The rate of energy transfer between phases is given by Eq.9 .

$$Q_{lk} = h_{lk}(T_l - T_k) \tag{9}$$

Where,h_{lk},represents heat transfer coefficient and ($T_l - T_k$) is the temperature driving force [25].

Ranz Marshall Equation, Eq. 10, has been used for interphase heat transfer coefficient. This equation is employed on the continuous phase side and for spherical particles.

$$Nu = 2 + 0.6\,\mathrm{Re}^{0.5}\,\mathrm{Pr}^{0.3} \qquad 0<Re<200 \qquad 0<Pr<250 \tag{10}$$

The Reynolds stressmodel, Eq. 11, uses an eddy viscosity hypothesis.

$$-\rho\left(\overline{u_i u_j}\right) = \mu_T\left(\frac{\partial U_i}{\partial x_i} + \frac{\partial Uj}{\partial x_i}\right) - \frac{2}{3}\delta_{ij}\left(\mu_T \frac{\partial U_k}{\partial x_k} + \rho k\right) \tag{11}$$

Where$e$,μ_T, the turbulent or eddy viscosity depends on the local state of flow or turbulence. Turbulent viscosity related to the characteristic velocity, u_T and length scales of turbulence , l_T.

$$\mu_T \propto \rho u_T l_T \tag{12}$$

Most simple models which estimate characteristic length and velocity scales are called zero equation models [25]. The zero equation model was employed for the dispered gas phase.

The k-ε model has been employed for turbulency of liquid phase. k and ε are turbulent kinetic energy and rate of energy dissipation. sThe k-ε model are presented by equations (13) and (14), respectively. Equations 15 to 19 are definition of parameters used in the k-ε model

$$\frac{\partial}{\partial t}(\omega_l \rho_l k) + \frac{\partial}{\partial x_i}(\omega_l \rho_l U_{l,i} k) - \frac{\partial}{\partial x_i}(\omega_l(\mu_{l,lam} + \frac{\mu_{l,tur}}{\sigma_k})\frac{\partial k}{\partial x_i}) = \omega_l(G - \rho_l \omega) + S_{l,k} \tag{13}$$

$$\frac{\partial}{\partial t}(\omega_l \rho_l \varepsilon) + \frac{\partial}{\partial x_i}(\omega_l \rho_l U_{li} \varepsilon) - \frac{\partial}{\partial x_i}(\omega_l(\mu_{l,lam} + \frac{\mu_{l,turb}}{\sigma\varepsilon})\frac{\partial \varepsilon}{\partial x_i}) = \omega_l \frac{\varepsilon}{k}(C_1 G - C_2 \rho_l \varepsilon) + S_{l,\varepsilon} \tag{14}$$

$$\mu_{turb} = C_\mu \rho_l \frac{k^2}{\varepsilon} \tag{15}$$

$$S_{l,k} = -A_D \frac{\nu^t{}_L}{\omega_L \omega_G \sigma_\omega}(U_L - U_G).\nabla\omega_G + 2A_D(C_t - 1)k \tag{16}$$

$$S_{l,\varepsilon} = 2A_D(C_t - 1)\varepsilon \tag{17}$$

$$A_D = \frac{3}{4}\frac{\omega_G \rho_L C_D}{d_b}\left|U_L - U_G\right| \tag{18}$$

$$G = \mu_L(\nabla U_L + \nabla U_L^T) : \nabla U_L \tag{19}$$

The constant values of C_1 and C_2 of Eq. 14 are 1.44 and 1.92, respectively [25]. ρ_l is liquid density. $S_{l\varepsilon}$ and $S_{l,k}$ are inter phases exchange terms that are qiven by Eqs. 16 and 17. Their values are assumed to be zero. G, in Eq. 19, is the generation of turbulent kinetic energy.

σ_ω is a turbulent Prandtl number, set to 1. ν_L^t is turbulent kinematic viscosity for liquid phase. C_t is a turbulent response coefficient. It is defined by Eq. 20:

$$C_t = \frac{u_G}{u_L} \tag{20}$$

u_G and u_L are gas phase velocity and liquid phase velocity, respectively. C_t approaches a constant value of 1 [25,26].

4. Results and discussions

4.1. Hydrodynamics

Simulation runs achived their steady states after 15 s. One of the important hydrodynamics parameters that affect operation of columns is gas hold-up. Gas hold-up contours at different times are shown in Fig. 3 and 4 for bubble column with five and nine heaters at the pitch to diameter ratio of 1.5 and at superficial gas velocity of 0.04 m/s. It is clear form the count or plots that location of gas accumulation varies with time. This dependancy was not studied further in this article.

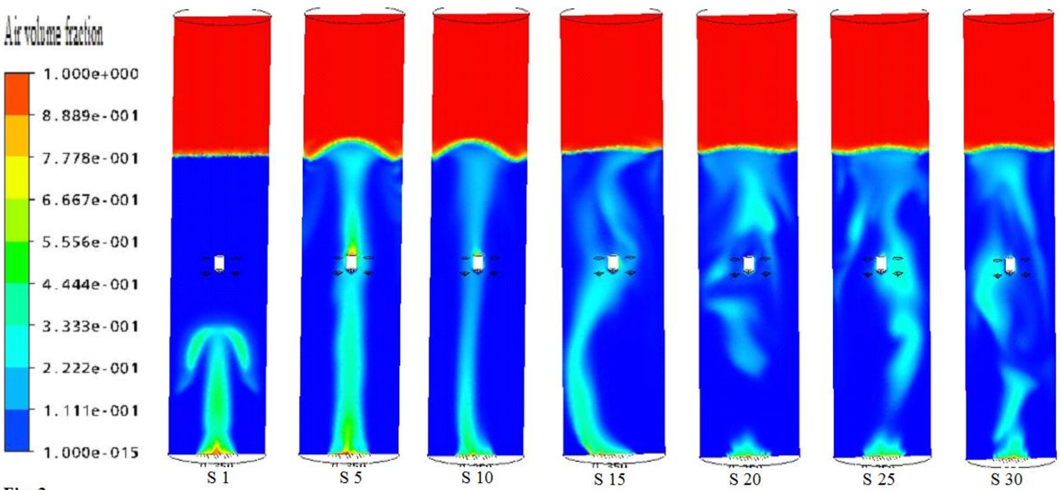


Fig. 3

Figure 3: Contour of gas holdup at different times in bubble column with five heaters Ug=0.04 m/s

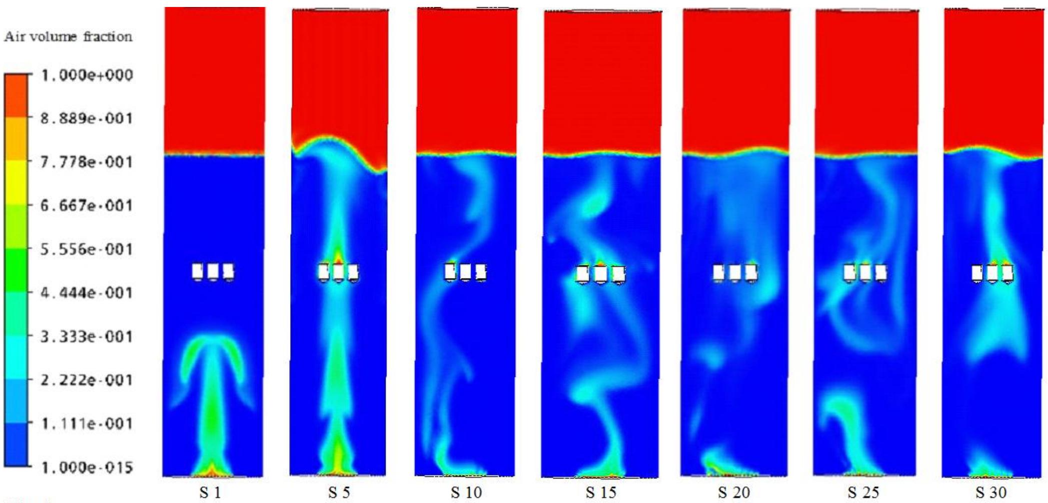


Figure 4: Contour of gas holdup at different times in bubble column with nine heaters Ug=0.04 m/s

Fig. 5 shows variation of gas holdup for bubble column with five heaters at the pitch to diameter ratiosof 1.5, 2 and 2.5. Superficial gas velocities varied in the range of 0.0025 to 0.04 m/s. Gas velocity covers the bubbly flow. The reasults show that with increasing superficial gas velocity or at a constant superficial gas velocity with increasing the pitch to diameter ratios from 1.5 to 2.5, the gas holdup increases. That is caused by the increasing in distance between the heaters which provide a wider passage for the liquid that can hold the gas phase and so increase the gas holdup.

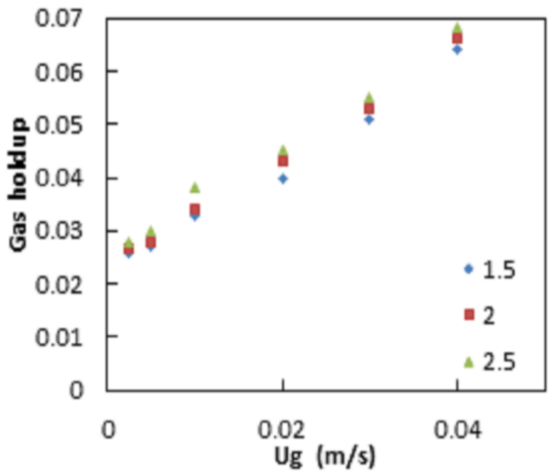


Figure 5: Variation of gas holdup versus superficial gas velocity for bubble column with five heaters at the pitch to diameter ratios of 1.5, 2 and 2.5

Fig. 6 shows variation of gas holdup for a bubble column with nine heaters and at the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46 versus superficial gas velocity. The reasults show that, with increasing superficial gas velocity, the gas holdup increases. At a constant superficial gas velocity, with increasing the pitch to diameter ratio from 1.5 to 2.46 the gas holdup increases. With decreasing the pitch to diameter ratio the distance between the heaters decreases. The volume of liquid passing around the heaters does not remain constant and decreses, so its velocity. On theother hand the heaters block free movments of the bubbles and some accumulation of bubbles occurs. Thus, the lowest pitch to diameter ratio of 1.25 gave a higher gas holdup than other ratios. Again the forgoing argument is validated in Fig. 7.

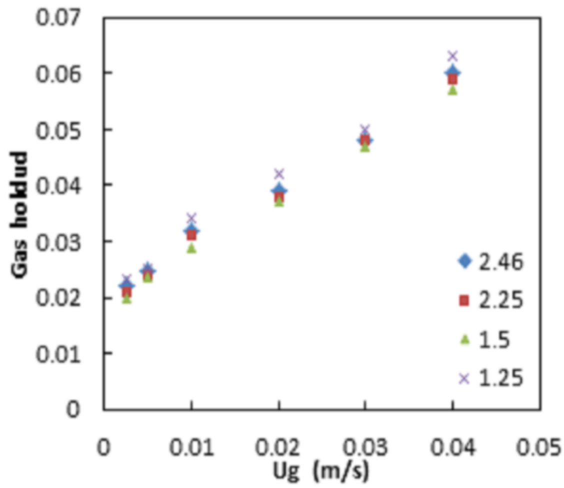


Figure 6: Variation of gas holdup with superficial gas velocity for bubble column with nine heaters at the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46

In Fig. 7 the gas holdup contour is shown for bubble column with nine heaters in the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46. It is clear that with decrease of the pitch to diameter ratio from 2.46 to 2.25 or 1.5 the distance between the heaters decreases. But with further reduction in the pitch to diameter ratio from 1.5 to 1.25 the bundle of heaters has become into an inter locking and integrated bundle of heaters. Thus, when bubbles passed around the heaters, they faced with a reduction in sectional area. So they move towards the wall of column but near to the heaters which causes a higher gas velocity.

Variation of gas hold up with the variation of superficial gas velocity for the 5 heaters and for the 9 heaters bundles at the constat pitch to diameter ratio of 1.5 of the heaters are shown in Fig. 8. At a constant gas velocity increase of the number of heaters reduces the gas hold up.

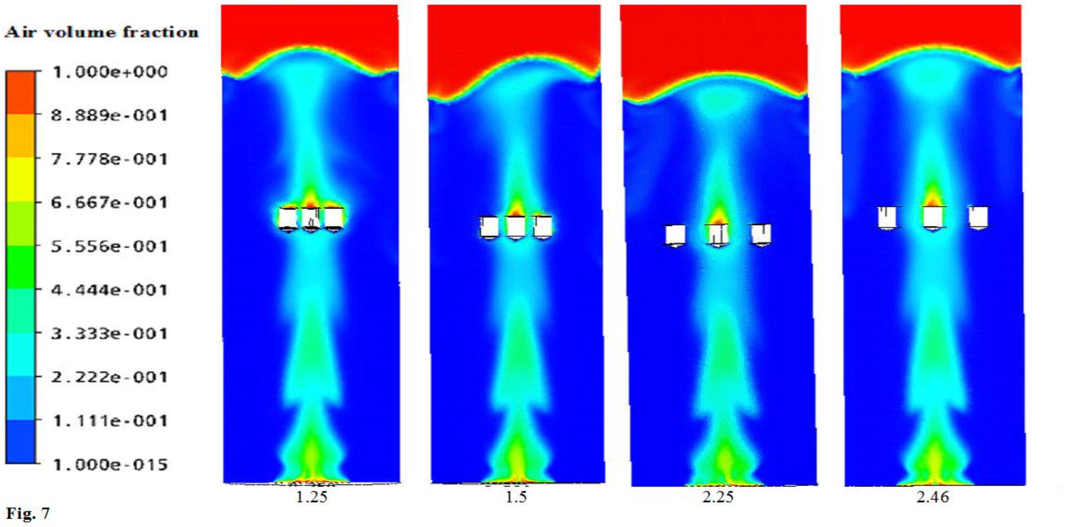


Figure 7: Contour of gas holdup for bubble column with nine heaters at the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46

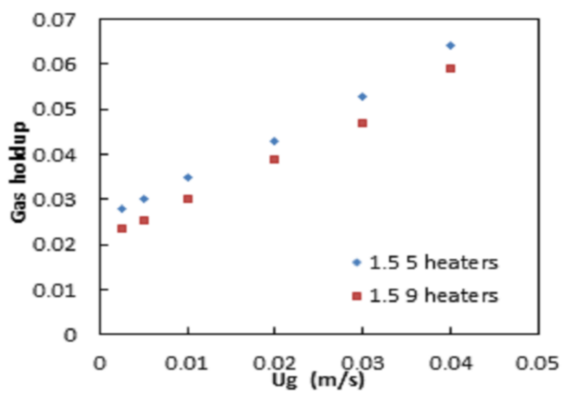


Figure 8: Comparison of the gas holdup in bubble columns with nine and five heaters at the pitch to diameter ratio of 1.5.

Radial distribution of gas holdup and liquid velocities with respect to radial distance from column axis at different superficial gas velocity and at height of 0.5 m from gas sparger in bubble column with five and nine heaters bundels of pitch to diameter ratio 1.5 are shown in Fig.9. Results indicate that with increasing superficial gas velocity, bubbles tend to concentrate in central zone of the columnbut, with increasing distance from column axis, concentration of bubbles decreases. This decreasing of bubble concentration makes a rigorous down flow of water near the column wall. Also upward liquid movment along the column axis and downward liquid movment near the column wall leads to a liquid circulation in bubble column. With increasing number of heaters radial distribution of gas holdup and radial distribution of liquid velocity decreases. Increasing the number of heaters in the column reduces the space available for passage of liquid. Impact of the rising bubbles with the heater surface causes bubble breakage, which in turn deccreases the radial distribution of gas. These effects cause less turbulancy in the column that have an impact on the heat transfer coefficient.

4.2. Heat transfer

4.2.1. Effect of heaters pitch to diameter ratios

One of the important parameter on study of operation of a bubble column is the knowledge of heat transfer coefficient. To consider the effects of heaters bundle geometry, location and operational parameters on the heat transfer coefficient the results reported after 45 s of CFD simulation run for column to reach quasi-steady state. Heat transfer coefficient with variation of pitch to diameter ratios of 5 and 9 heaters tube bundle and with variation of gas superficial velocity were investigated. The heater bundles were positioned either in centeral area of the column or near to the column wall.

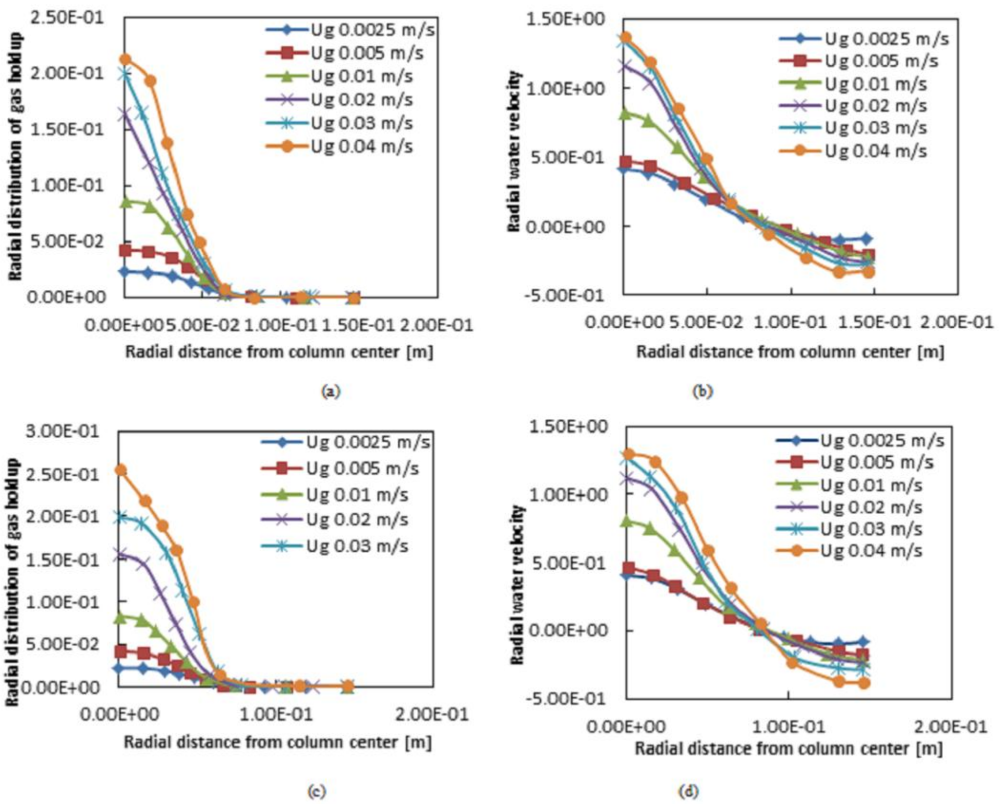


Figure 9: Radial gas holdup versusdimensionless distance from the axis at a superficial gas velocity for bubble column with (a)nine heaters and (c) five heaters

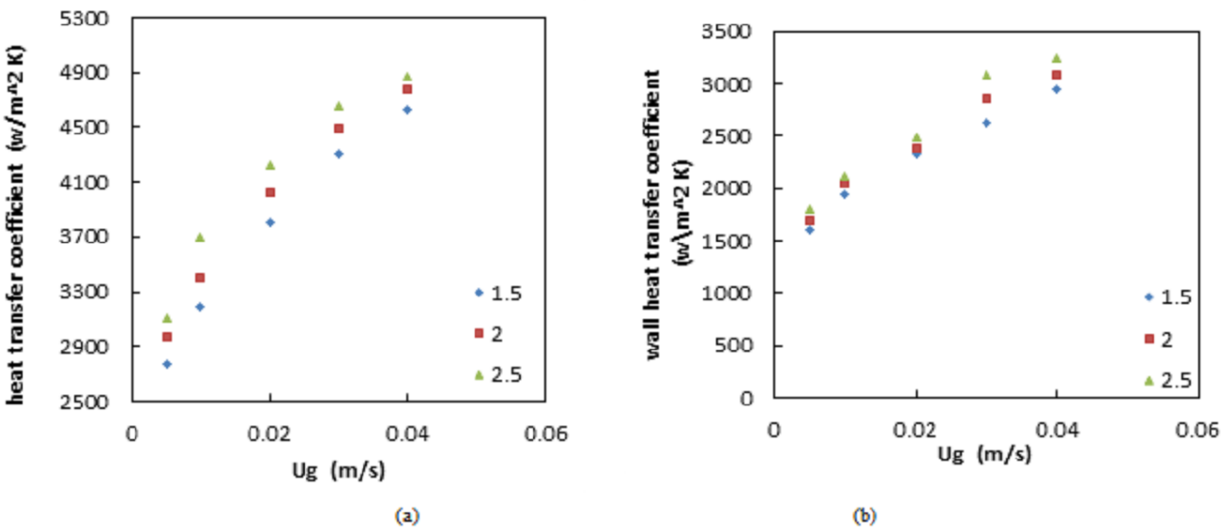
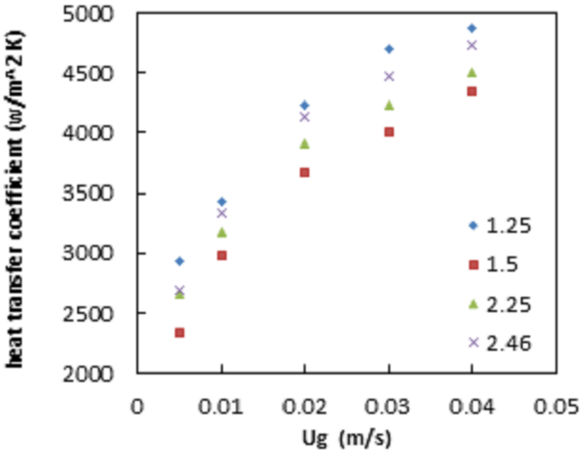


Figure 10: Heat transfer coefficient versus superficial gas velocity at different pitch to diameter ratios for 5 heaters tube bundel posined at (a) axis and (b) near wall



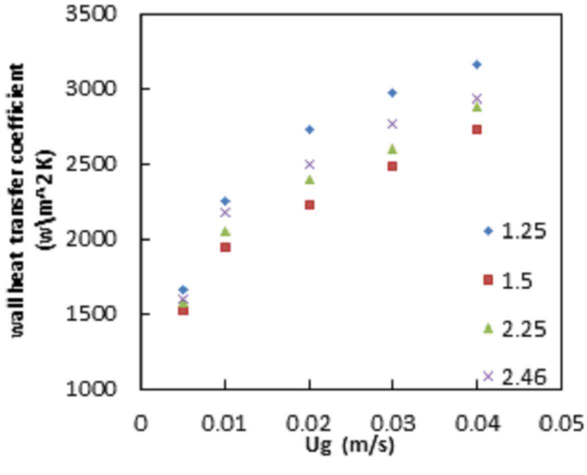


Figure 11 : Heat transfer coefficient versus superficial gas velocity at different pitch to diameter ratios for 9 heaters tube bundle poisoned at (a) axis and (b) near wall

Fig. 10 shows variation of centeral and wall heat transfer coefficients with variation of superficial gas velocity in bubble column with five heaters at the pitch to diameter ratios of 1.5, 2 and 2.5. The reasults show that, with increasing superficial gas velocity the heat transfer coefficient increases. This is in part due to the production of larger bubbles[27] and hence higher liquid circulation velocity. At a constant superficial gas velocity, with increasing the pitch to diameter ratio from 1.5 to 2.5 the heat transfer coefficient increases. With increasing the distance between heaters the quantity of liquid and gas bubbles passing through the heaters would increase. Thus the heat transfer coefficient with increasing the pitch to diameter ratio increases.

Fig. 11 shows variation of centeral and wall heat transfer coefficients with variation of superficial gas velocity in bubble column with nine heaters at the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46. The results show that, with increasing superficial gas velocity the heat transfer coefficient increases. By narrowing the distance between the heaters, the quantity of liquid passing through the heaters decreases but, the velocity of the liquid passing through the passage may increase. However a balance between amount of liquid passing through the passage and pressure drop establishes. So, at a constant superficial gas velocity, with decreasing the pitch to diameter ratio from 2.46 to 1.5 a reduction in the heat transfer coefficient observed. Although in the lowest pitch to diameter ratio of 1.25 heat transfer coefficient was high.

That observations were validated with experimentally available data [28] in Fig. 12. The results of CFD simulation with 9.75% difference are in good agreement with the results of experimental data.

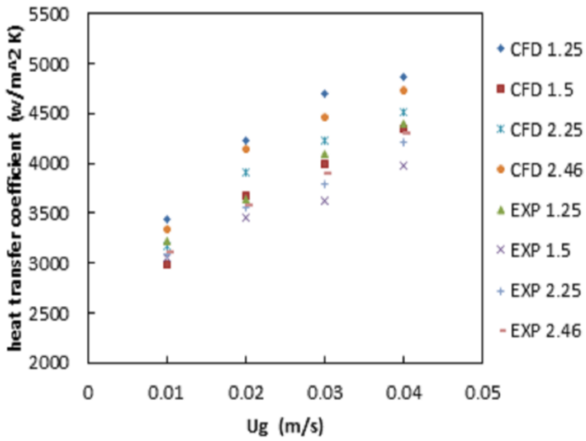


Figure 12: A comparison between CFD simulation and experimental data for bubble column with 9 heaters

Comparison between the wall and the centeral heat transfer coefficients for bubble column with five and nine heaters at the pitch to diameter ratio of 1.5 are shown in Fig. 13. The tendency to rise of bubbles near to the wall or near to the axis of the column are for small and large bubbles, respectively. As the contribution of larger bubbles for heat transfer are more than small ones [27], heat transfer coeficient for central heaters is higher than the wall heaters.

Fig. 14 is a comparison between the heat transfer coefficient of a bubble column with five heaters and a bubble column with nine heaters, with pitch to diameter ratios of the heaters of 1.5. The heaters are placed at the column axis and at the column wall.

Fig. 15 shows the measured [28] and CFD determined heat transfer coefficient for bubble column with 3 by 3 and 5 by 5 hearers bundels. CFD simulation data of the bubble column with nine heaters with the pitch to diameter ratio of 1.25 are included. It shows that with increasing the number of heaters the heat transfer coefficient decreases. By increasing the number of heaters in the column the available space for the passage of liquid and gas bubbles flow decreases which causes a reduction in the liquid velocity or reduces turbulancy. Therefor reduction of heat transfer coefficient can be anticipated.

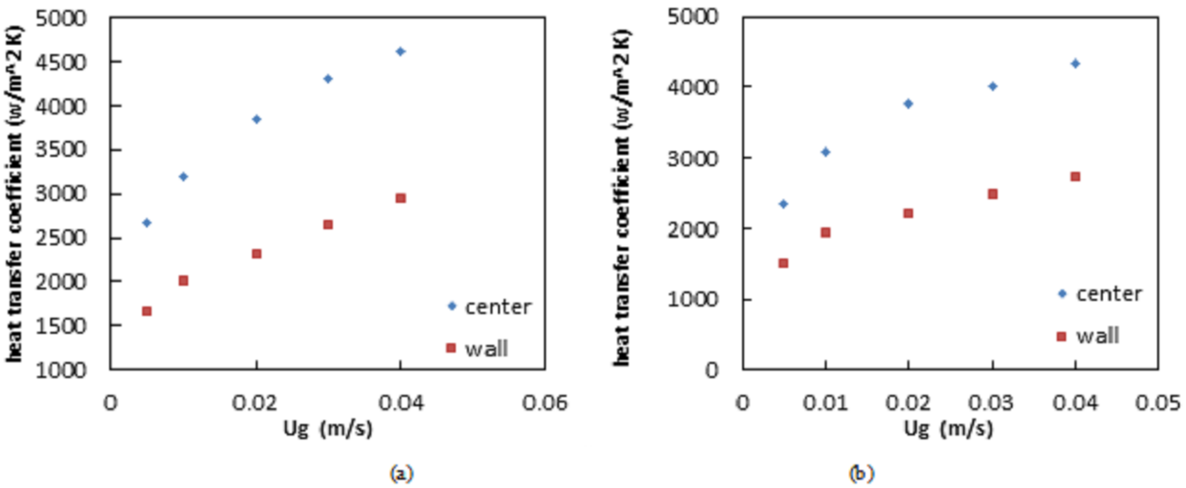


Figure 13: A comparison between wall and centeral heat transfer coefficient for bubble column with (a) five heaters, (b) nine heaters

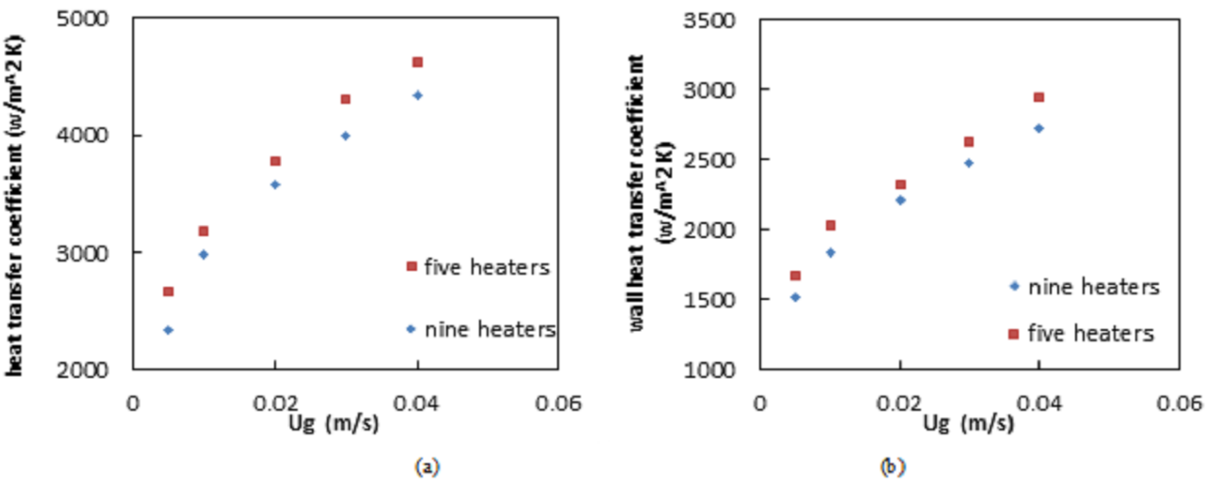


Figure 14: A comparison between heat transfer coefficient for 5 and 9 heaters tube bundel positioned (a) at the column axis and (b) positioned at the wall

4.2.2. Empirical correlations and CFD simulation dataon heat transfer coefficient

Table 3 gives a number of reported correlations for calculation of heat transfer coefficent. Fig. 16 shows estimated heat transfer coefficient with the variation of superficial gas velocity. The CFD simulation data for 5 and 9 heaters tube bundles with pitch to diameter ratio of 1.5 are included. CFD data are in good agreement with the estimated values from Fair*et al.*[22] correlation. The wide differences between estimation of different corrlations are obvious.

Table 3: A number of literature heat transfer coefficient correlations

$h=8850Ug^{0.22}$	Fair [22]
$St=0.125(Re.Fr.Pr^{2.4})^{-0.25}$	Hart[29]
$St=0.11(Re.Fr.Pr^{2.48})^{-0.23}$	Burkel [30]
$St=0.1(Re.Fr.Pr^{2})^{-0.25}$	**De**ckwer[31]
$St=0.124(Re.Fr.Pr^{2.5})^{-0.22}$	Kolbel[32]
$St=0.1(Re.Fr.Pr^{2})^{-0.22}$	Kast.[33]

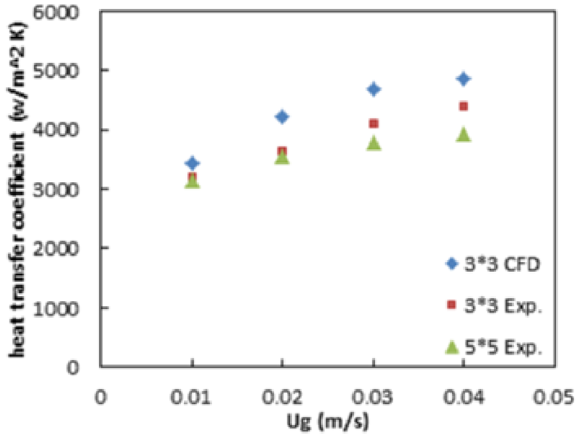


Fig. 15

Figure 15: A comparison between heat transfer coefficient of bubble columns having 25 and 9 heaters with experimental data [22] and with CFD simulation data atpitch to diameter ratio 1.25

Table 4: Comparison of heat transfer coefficient correlations

$St=0.0021\times Ug^{-0.78}$	Fair [22]
$St=0.0023\times Ug^{-0.75}$	Hart [29]
$St=0.0027\times Ug^{-0.69}$	Burkel [30]
$St=0.0022\times Ug^{-0.75}$	Deckwer[31]
$St=0.0035\times Ug^{-0.66}$	Kolbel [32]
$St=0.0035\times Ug^{-0.66}$	Kast[[33]
$St=0.0022\times Ug^{-0.76}$	This work five heater
$St=0.0026\times Ug^{-0.76}$	This work nine heater

In Tables 4, 5 and 6 the results of CFD simulation with empirical correlations and experimental data have been compared. The comparison is based on the variation of heat transfer coefficient as Stanton number,St., with superficial gas velocity in frlational form; $St \propto Ug^{n}$. Data of Table 4 are obtained from emperical correlation. Data in Tables 5 and 6 showeffect of 9 and 5 heaters tube bundles positions in centeral or wall area on the heat transfer coefficient.

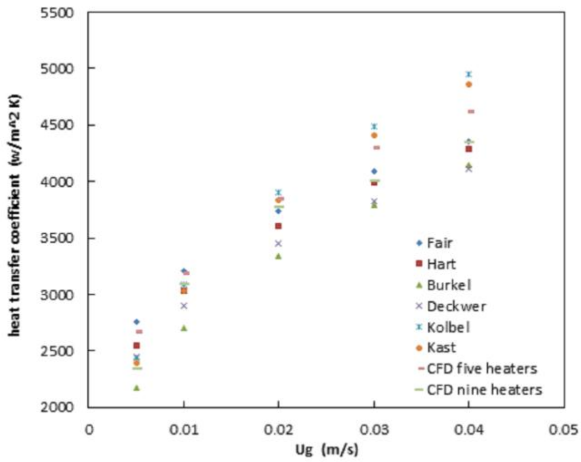


Fig. 16

Figure 16: A comparison between heat transfer coefficient resulted from CFD simulation and literature correlation

The average value of n is between -0.7 to -0.8. This means that increase of heat transfer coefficient with the increase of Ug is not sharp. Indeed for high gas velocities in the churn turbulent flow regime heat transfer coefficient becoms independent of superficial gas velocity [16, 22].

Heat transfer coefficint is found to be a function of number of heaters, N, pitch, t, and diameter of heaters tube, d, as well as superficial gas velocity and liquid velocity. That dependency correlated by using curve fitting method. The resulted correlations are reported in Table 7. The reported correlations had considred the effct of number of hetears, Pitch and heater dimension and in that regard are new.

Table 5: Comparison of heat transfer transfer coefficient correlations for bubble column with nine heaters. Heater position and pitch to diameter ratio is given

$St=0.0028\times Ug^{-0.74}$	CFD, center, pitch to diameter 1.25
$St=0.0026\times Ug^{-0.76}$	CFD, center, pitch to diameter 1.5
$St=0.0025\times Ug^{-0.75}$	CFD, center, pitch to diameter 2.25
$St=0.0026\times Ug^{-0.75}$	CFD, center, pitch to diameter 2.46
$St=0.0021\times Ug^{-0.7}$	CFD, wall, pitch to diameter 1.25
$St=0.0015\times Ug^{-0.73}$	CFD, wall, pitch to diameter 1.5
$St=0.0017\times Ug^{-0.72}$	CFD, wall, pitch to diameter 2.25
$St=0.0018\times Ug^{-0.72}$	CFD, wall, pitch to diameter 2.46
$St=0.0022\times Ug^{-0.77}$	Exp.,center,pitch to diameter 1.25
$St=0.0017\times Ug^{-0.82}$	Exp.,center pitch to diameter 1.5
$St=0.00196\times Ug^{-0.79}$	Exp.,center pitch to diameter 2.25
$St=0.0021\times Ug^{-0.77}$	Exp. center pitch to diameter 2.46

Table 6: Comparison of heat transfer coefficient correlations for bubble column with five heaters. Heater position and pitch to diameter ratio is given

$St=0.0022\times Ug^{-0.76}$	CFD, center, pitch to diameter 1.5
$St=0.0024\times Ug^{-0.75}$	CFD, center, pitch to diameter 2
$St=0.0027\times Ug^{-0.74}$	CFD, center, pitch to diameter 2.5
$St=0.0016\times Ug^{-0.75}$	CFD, wall, pitch to diameter 1.5
$St=0.0018\times Ug^{-0.71}$	CFD, wall, pitch to diameter 2
$St=0.002\times Ug^{-0.71}$	CFD, wall, pitch to diameter 2.5

4.2.3. Column temperature variations

High degree of mixing in BCs that results in the uniformity of liquid or column temperature is one of the advantages of using bubble columns in chemical and bio-chemical industries [7]. Nevertheless, this proposal was investigated. It is obvious that heating of liquid at adiabatic situation increases its temperature as time passes. The effect of superficial gas velocities on temperature rise of BC with time for bubble columns having five and nine heaters tube bundle were noted. The pitch to diameter ratios of 2.5 and 2.46 are used for the 5 heaters and for the nine heaters tube bundles, respectively. The results are shown in Fig. 17 a and b.

Table 7: Proposed heat transfer correlation for bubble column

$St=0.175[Re.Fr.Pr^2]^{-0.30}\times F(N)\times F(P)$ $F(N)=N^{(-0.13+2.5e(-6)\times Re)}$ $F(P)=[\frac{t}{d}]^{(0.22-1.6e(-5)\times Re)}$	CFD,BC with five heater
$St=0.14[Re.Fr.Pr^2]^{-0.292}\times F(N)\times F(P)$ $F(N)= N^{(-0.13+1.2e(-5)\times Re)}$ $F(P)=[\frac{t}{d}]^{(-0.024+1.4e(-5)\times Re)}$	CFD,BC with nine heater
$St=0.145[Re.Fr.Pr^2]^{-0.28}\times F(N)\times F(P)$ $F(N)= N^{(-0.13+2.8e(-6)\times Re)}$ $F(P)=[\frac{t}{d}]^{(-0.0202+1.04e(-5)\times Re)}$	Exp. BC with nine heater[28]
$St=0.16[Re.Fr.Pr^2]^{-0.28}\times F(N)\times F(P)$ $F(N)= N^{(-0.13+2.8e(-6)\times Re)}$ $F(P)=[\frac{t}{d}]^{(-0.0202+1.04e(-5)\times Re)}$	General Equation for BC

Fig. 17 shows that, the column temperature increases with time at a constant superficial gas velocity. Also with increasing superficial gas velocity the column temperature and heat transfer coefficient increases. Heat transfer coefficient is basically a function of Reynolds, Re, and Prandt, Pr, Numbers. Re number is a function of velocity but both Re and Pr are dependent on the liquid viscosity. As a result, the combined effect is that with increasing superficial gas velocity the column temperature and heat transfer coefficient increases.

The effect of number of heaters and their pitch on liquid temperature of BC, at a constant superficial gas velocity of 0.02 m/s, with time is presented in Fig. 18. The results show that, with increasing number of heaters the column temperature increases.

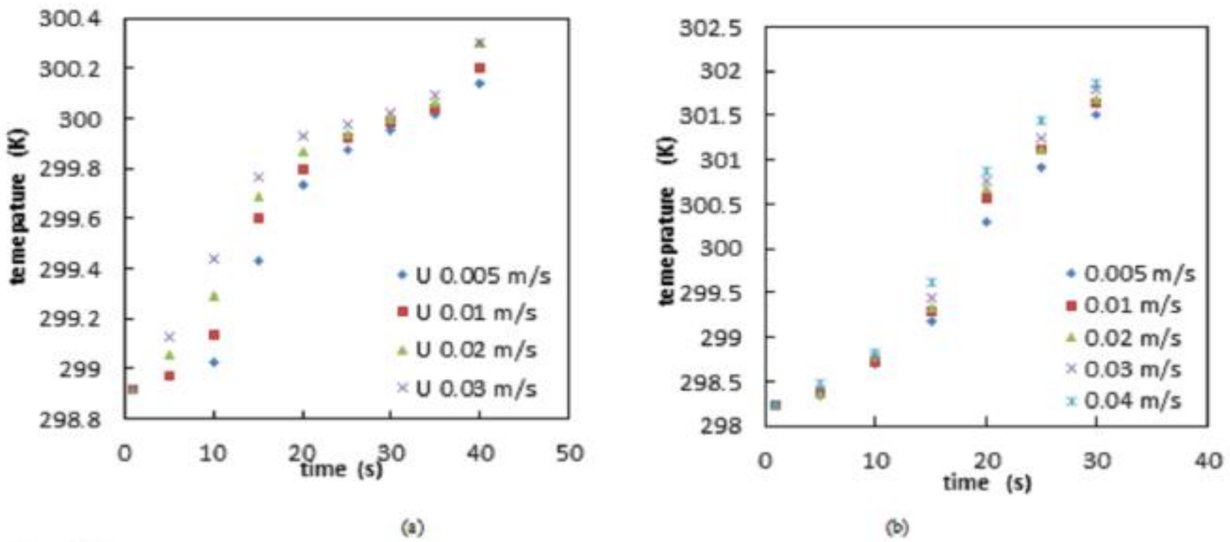


Fig. 17

Figure 17: (a) Temperature variations of bubble column with five heaters at the pitch to diameter ratio of 2.5, (b) Temperature variations of bubble column with nine heaters at the pitch to diameter ratio of 2.46

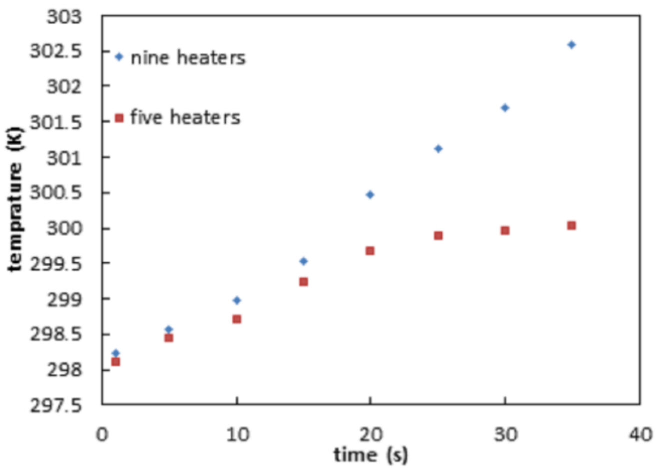


Fig. 18

Figure 18: Column temperature comparison ofa bubble columns with fiveheaters and a bubble column with nine heaters

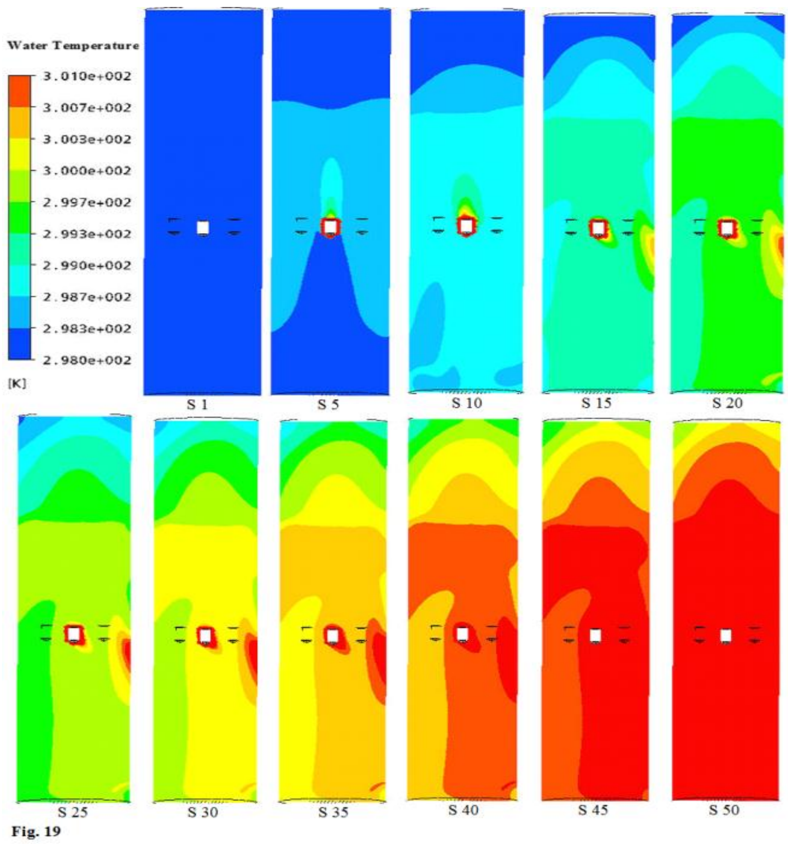


Figure 19: Contour of the water temperature in bubble column with five heaters

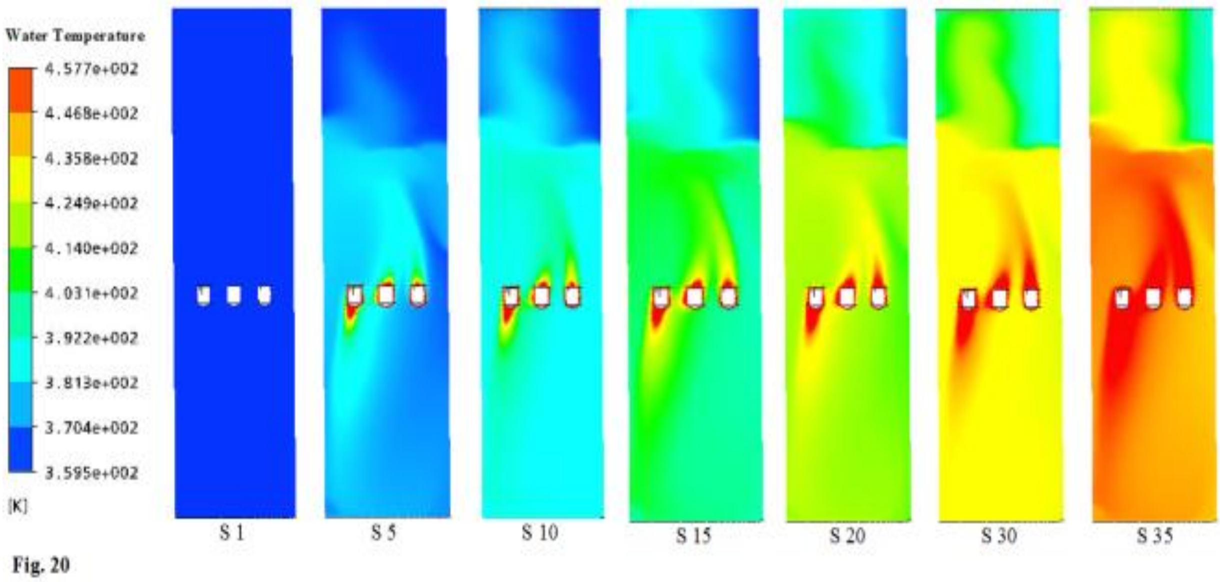


Figure 20: Contour of the water temperature in bubble column with nine heaters

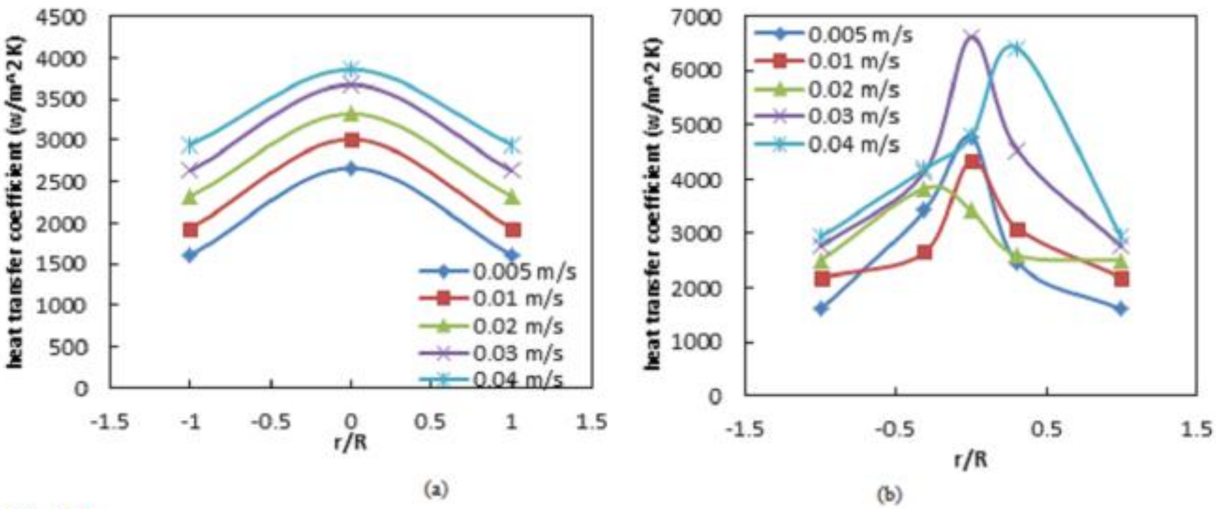


Fig. 21

Figure 21: (a) Variations of radial heat transfer coefficient in bubble column with five heaters, (b) Variations of radial heat transfer coefficient in bubble column with nine heaters

Figs. 19 and 20 show the contour of water temperature variations for bubble column with five heaters and bubble column with nine heaters at Ug=0.01 m/s. Those figures show that due to the high heat transfer coefficient and high degree of mixing that exist in bubble column the temperature uniformity is achieved in a short period of time.

4.2.4. Radial distribution of heat transfer coefficient

To better understand the system from thermal point of view the variations of radial heat transfer coefficient at different superficial gas velocities at distance 0.65 m from the gas distributor for bubble column with five heaters and bubble column with nine heaters at the pitch to diameter ratio of 1.5 are presented in Fig. 21.

In bubble columns the small bubbles tend to move in near the column wall and large bubbles tend to move in column axis. Large bubbles increase the heat transfer coefficient in column. Therefore, the heat transfer coefficient at the column axis is more than the heat transfer at the wall [27]. In bubble columns the bubbles may move in radial positive direction or in radial negative direction. In the same direction that the gas bubbles moves the heat transfer coefficient is higher. Increase of the number of heaters introduces more obstacle to the free movement of liquid and bubbles. That is shown by more scattering of data presented by Fig. 21b for nine heaters than Fig. 21a for five heaters.

5. Conclusions

Variation of pitch to diameter ratio and the number of heaters to find their mutual effects on the heat transfer coefficient of BC were investigated. The investigation carried out by 3D CFD modeling of BC in Eulerian- Eulerian framework. For this intent the simulation has been done in two bubble columns, one of them with five heaters and the other with nine heaters. The bubble column with five heaters was investigated with the pitch to diameter ratios of 1.5, 2 and 2.5. Bubble column with nine heaters was investigated with the pitch to diameter ratios of 1.25, 1.5, 2.25 and 2.46. Results show that:

- With decreasing the number of heaters the heat transfer coefficient increases.
- In bubble column with five heaters at a constant superficial gas velocity by increasing the pitch to diameter ratio from 1.5 to 2.5 the heat transfer coefficient decreases.

- In bubble column with nine heaters at a constant superficial gas velocity, with decreasing the pitch to diameter ratio from 2.46 to 1.5 the heat transfer coefficient decreases. But at lowest pitch to diameter ratio of 1.25, heat transfer coefficient increases. That shows an optimum value may exist.
- Away from the wall heat transfer coefficient increases.
- The heat transfer coefficients that calculated were in the range of 2000 to 5000 w/m^{2}K.

Correlations that take effect of element size, number and pitch were provided.

6. Nomenclature

C_1, C_2, C_μ	[-]	Constant of k-ε model
C	[-]	Constant of the Deckwer correlation
C_D	[-]	Drag coefficient
C_{pk}	[kj/kg.K]	Heat capacity of kth phase
C_t	[-]	Turbulence response coefficient
D	[m]	Diameter of heater
$Fr=(u_g)^2/(gd_b)$	[-]	Froude number ∝ (inertia force)/(gravity force)
G	[-]	Generation of turbulent kinetic energy
G	[m.s^{-2}]	Gravitational acceleration
H	[w/m^{2}K]	Heat transfer coefficient
k	[j.kg^{-1}],[J/(m.s.^{o}K)	Turbulent kinetic energy, thermal conductivity
M	[kg.m^{-2}.S^{-2}]	Interphase momentum exchange
M_{kl}	[kg.m^{-2}.S^{-2}]	Interphase momentum transfer
N	[-]	Number of heaters
Nu =hd/k	[-]	Nusselt number
P	[N.m^{-2}]	Pressure
Pr=ν/α	[-]	Prandtl number
Q	[w.m^{-2}]	Heat flux
$Re=(\rho_l u_g d_b)/\mu_l$	[-]	Reynolds ∝ (inertia force/viscous force)
St	[-]	Stanton
S_{lk}, $S_{l,\varepsilon}$	[-]	Inter phases exchange terms
S_{pk}	[-]	Transport rate from p phase to k phase
S_{kp}	[-]	Transport rate from k phase to p phase
T	[s]	Time
T	[K]	Temperature
Ug	[m/s]	Superficial gas velocity

Greek Symbols

$\alpha=k/C_p\rho$	[m^2.s^{-1}]	Thermal diffusivity
ω_k	[-]	Volume fraction of kth phase
μ_l	[kg.m^{-1}.s^{-1}]	Viscosity of liquid
μ_k	[kg.m^{-1}.s^{-1}]	Viscosity of gas
$\upsilon_l=\mu/\rho$	[m^2.s^{-1}]	Kinematic viscosity
Σ	[kg.s^{-2}]	Surface tension coefficient
σ_k, σ_ε, σ_ω	[-]	Turbulent Prandtl numbers for k-ε
Ε	[w.kg^{-1}]	Dissipation rate of k
Ρ	[kg.m^{-3}]	Density
ρ_k	[kg.m^{-3}]	Gas density
ρ_l	[kg.m^{-3}]	Liquid density
Δρ	[kg.m^{-3}]	Density difference between the phases

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