Experimental Measurements and Theoretical Prediction for the Volumetric Heat Transfer Coefficient of a Three-Phase Direct Contact Condenser

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ABSTRACT

The volumetric heat transfer coefficient of a three-phase direct contact heat transfer condenser has been investigated analytically and experimentally. The experiments were carried out utilising a column of 70 cm in total height and 4 cm inner diameter. The active column height throughout the experiments was taken to be equal to 48 cm. Vapour pentane with three different initial temperatures (40°C, 43.5°C and 47.5°C) was used as a dispersed phase, while tap water at a constant temperature (19°C) was used as a continuous phase. The variation of the volumetric heat transfer coefficient along the height of the column was measured experimentally and predicted analytically. The effects of the initial dispersed phase temperature, the dispersed mass flow rate and the continuous mass flow rate on the volumetric heat transfer coefficient were tested. The results indicate that the volumetric heat transfer coefficient decreases upon moving up the column, while it increases with an increase in the mass flow rate of either the dispersed phase or the continuous phase. No considerable impact of the dispersed initial temperature on the volumetric heat transfer coefficient was observed under the experimental conditions considered here. Finally, an excellent agreement was achieved between the analytical model and the experimental results.

Keywords: direct contact condenser; volumetric heat transfer coefficient; modelling

1. Introduction

In general, direct contact condensers (DCC), as a type of direct contact heat transfer equipment, have many advantages over a surface type condenser (i.e. a shell and tube condenser), namely lower cost, a closer temperature approach (in some cases less than 1°C), , a high heat transfer rate, and the elimination of problems with fouling and corrosion. They are also efficient in extracting non-condensable gases, smaller in size and because of the improved thermal performance, less cooling medium is often required (up to 60% less than that needed in surface type condensers), which makes them a preferred option in geothermal power production plants. The main drawback facing the direct contact condenser is the problem of the mixing between the condensate produced and the cooling fluid because of the potentially high cost of separation and the requirement for special equipment. However, this difficulty can be solved by using two immiscible liquids. Therefore, direct contact condensers can be found in different industrial applications such as water desalination, geothermal power production, solar energy applications and recently in bio fuel condensation [1].

Direct contact condensers can be classified into two practical types: single component and two-component. In the former, the same species is used on both sides the condensation process, e.g. condensation of steam in water, while the latter is concerned with condensation using a second, different immiscible liquid, for instance the condensation of pentane vapour in water. Much attention has been paid to the one-component direct contact condenser in the literature. On the other hand, studies of two-component direct contact condensers mainly concentrate on understanding the direct condensation phenomena through investigation of the heat transfer and the hydrodynamics of a single bubble [2-13] and a train of bubbles [14,15]. The resulting lack of understanding has severely limited the practical implementation of such condensers.

Only the work of Sideman and Moalem [16], who developed a mathematical model describing the condensation of multi-bubble in an immiscible liquid, is directly relevant to the topic considered herein. Although this study is not representative of a real three-phase direct contact condenser, it has improved the general understanding of the condensation of a swarm of two phase bubbles in an immiscible liquid. More recently, Mahood et al. [17-20] have undertaken a research programme concerned with the investigation of a bubble type three-phase direct contact condenser. Different parameters, which have an effect on the characteristic behaviour of a three-phase direct contact condenser have been investigated experimentally and theoretically.

In this investigation, experimental measurements and an analytical model of the variation in the volumetric heat transfer coefficient along a bubble type three-phase direct contact condenser are developed. The measurements and the prediction assume steady – state operation for different light hydrocarbon (pentane) vapour to water (continuous phase) mass flow rate ratios, and different initial vapour temperatures.

2. Experimental Study

2.1. Experimental setup

Figure (1) shows the experimental setup employed in the present investigation. It consisted of two main separated loops with auxiliary equipment. The first loop was for the dispersed phase, while the second for the continuous phase. The former loop comprised the test column (DCC), the vaporising or heating vessel, the circulation pump, the connecting tubes and the working fluid (pentane) with its small storage tank. The direct contact condenser was a 70 cm vertical cylindrical column of internal diameter 4 cm, connected to the heating vessel through a short (10 cm length and 6 mm internal diameter) externally heated tube. The heating vessel was a cubical shape, stainless steel pool type boiler of 100 cm height and about 50 cm width.

It contained a long, (6 m) small diameter (6 mm) copper coil used for carrying, warming and vaporising the dispersed phase (pentane). This coil was completely immersed in water which filled the heating vessel and was warmed by two electric heaters (total power 6 kW). These heaters were controlled by a thermostat to avoid the overheating. The coil was connected to the liquid pentane storage tank via the circulation pump. The liquid pentane flow rate was controlled by a ball valve, and a non return valve was included to avoid liquid pentane back flow due to the pressure head of the heating vessel, especially during a shut down period between two runs. The outlet of the coil was connected to the DCC test section via a short externally heated tube. A trace heater with temperature controller was used to maintain an approximately constant dispersed phase (vapour) temperature t through the short injection tube. The pentane vapour flow rate was controlled by a ball valve, using a similar technique to that described in [17, 21-23], and its temperature and pressure were measured just before injection, via a sparger, into the test column using a calibrated K-type thermocouple and a pressure gauge. The dispersed phase condensate was collected as a separated layer, due to gravity, from the top of the DCC test section. It was then sent back to the liquid pentane storage tank for continuous operation or stored and used in another run.

The continuous phase loop consisted of a large (160 l) water tank, pump, flow meters and connection pipes. Two rotameters (one on the water inlet stream and another on the water outlet stream) were used to measure and control the continuous phase flow rate and to maintain water level into the DCC column.

The temperature along the test section was measured using five calibrated K-type thermocouples fixed at positions from the bottom column: TC1 @ 0 cm, TC2 @ 12 cm, TC3 @ 24 cm, TC4 @ 36 and TC5 @ 48 cm (which measured the water inlet temperature). Three more thermocouples were used for measuring the condensate temperature (at a height of 52 cm), the temperature of the injected vapour and the temperature of the dispersed phase at the

exit of the heating vessel. All of these thermocouples were connected to an eight channel data logger to display the measured temperatures directly on a PC.

2.2. Experimental procedure

A constant temperature cooling water (about 19 °C) was prepared in the large water storage tank and circulated through the test section (DCC). This ensured that the temperature was uniform within the column. The short individual run time and the large storage tank size helped to maintain the cooling phase temperature at a constant value. The water level of the cooling water, or continuous phase (48 cm), was achieved by adjusting the inlet and outlet water flow rate. The heating vessel was warmed up to the desired temperature. The heating temperature was controlled and increased gradually until the desired temperature inside the heating vessel was achieved. The liquid pentane was then pumped to the heating vessel where it boiled and formed a vapour (the normal boiling point of pentane is 36.5 °C). When it reached the desired temperature inside the coil, the injection valve was opened by a carefully calibrated amount. Immediately after the vapour was injected in to the DCC column, the direct contact condensation process is considered to have begun and the temperature along the column was recorded on the PC. The vapour injection temperature and pressure were maintained nearly constant throughout the experiments, where they were measured by a dedicated thermocouple and pressure gauge respectively. Finally, the dispersed phase mass flow rate was determined using a mass balance, where the collected condensate was weighted and divided by the experimental duration (about 3 minutes). To recover nearly all of the condensate, a conical separating flask was used, where the water was drained out while the condensate was returned back to the liquid pentane storage tank or collected and used again in another run.

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Fig. 1. Schematic diagram of the experimental rig.

3. Modelling

Mathematically, the volumetric heat transfer can be derived from the surface heat transfer coefficient using the following relation [24]:

$$U_v = n_o \pi D^2 h_d \quad (1)$$

Where n_o , D and h_d represent the initial number of bubbles, the bubble diameter and the surface heat transfer coefficient respectively.

154 The number of bubbles may be found as a function of holdup ratio (ϕ) as:

$$155 \quad n_o = \frac{\phi}{\frac{\pi D^3}{6}} \quad (2)$$

156 Therefore, Eq.(1) can be written as:

$$157 \quad U_v = \left(\frac{6\phi}{D}\right) h_d \quad (3)$$

158 Mahood et al.[17] have derived the surface heat transfer coefficient in a three-phase direct
159 contact condenser, and used it to model the size of a two-phase bubble swarm [18] as:

$$160 \quad h_d = \frac{2k_c}{\sqrt{3}\pi} \left(\frac{\beta U_r}{a\epsilon}\right)^{0.5} \quad (4)$$

161 where

$$162 \quad \beta = \frac{(\phi+1/2)}{(1-\phi)} \quad (5)$$

163 However, for a single two-phase bubble condensing in another immiscible liquid, $\phi \rightarrow 0$, and
164 Eq.(4) becomes:

$$165 \quad h_d = \frac{2k_c}{\sqrt{6}\pi} \left(\frac{U_r}{a\epsilon}\right)^{0.5} \quad (6)$$

166 The two-phase bubble diameter (D) appearing in Eq.(3), can be found using an energy
167 balance over the whole two-phase bubble. In this context, Moalem et al. [25] developed the
168 following expression for a two-phase bubble condensing in another immiscible liquid:

$$169 \quad \frac{da}{dt} = -\frac{k_c \Delta T}{\rho_{dv} h_{fg}} \left(\frac{2U_r k_v}{\pi a \epsilon}\right)^{0.5} \quad (7)$$

170 and

$$171 \quad k_v = 0.25 Pr^{-\frac{1}{3}} \quad (8a)$$

$$172 \quad \Delta T = \Delta T_{min} + (\Delta T_{max} - \Delta T_{min}) \frac{Z}{H} \quad (8b)$$

173 Eq. (8b) represents a linear temperature difference along the direct contact column [24, 26].

174 Where Z and H are the local position and the active column height respectively.

175 Integration of Eq.(7), with initial condition, $t = 0$, $a = a_o$ results in:

$$D = \left[D_o^{3/2} - \left(\frac{3 k_c}{\rho_v h_{fg}} \right) \left(\frac{kv}{\pi \epsilon} \right)^{0.5} U_r^{-0.5} \cdot \frac{Z^2}{H} \right]^{\frac{2}{3}} \quad (9)$$

where

$$D = 2a$$

Substituting Eq.(6) into Eq.(3) and using Eq.(9), results in:

$$U_v = 6\phi C_m U_r^{0.5} \left[D_o^{\frac{3}{2}} - C_{m1} U_r^{-0.5} \frac{Z^2}{2H} \right]^{\frac{2}{3}} \quad (10)$$

where

$$C_m = \frac{2k_c}{\sqrt{3\pi\epsilon}} \quad (11)$$

and

$$C_{m1} = \left(\frac{3k_c}{\rho_v h_{fg}} \right) \left(\frac{kv}{\pi \epsilon} \right)^{0.5} \quad (12)$$

4. Results and Discussion

4.1. Experimental results

The continuous phase temperatures along the column height was measured at four points along the column as well as at the inlet and outlet ports, while the dispersed phase temperature was only measured at the inlet and outlet. The direct contact condenser's height was divided into four sub-heights or sub-volumes, depending on the number of thermocouples. The volumetric heat transfer coefficient along the direct contact column was estimated for each sub-volume depending by considering the inlet and outlet temperature of the dispersed and continuous phases.

The difficulty in measuring the dispersed phase temperature along the direct contact condenser height was solved by using Antoine Equation to estimate the variation in the saturation temperature due to hydrostatic effects. Consequently, for each sub-volume, the inlet and the outlet dispersed phase temperature have been updated. Furthermore, the change in the

dispersed phase mass flow rate ($\dot{m}_{di}$) for each sub-volume due to the condensation progress along the direct contact condenser height is calculated using an energy balance. In this case the actual latent heat transfer along the condenser is updated.

Based on the following assumptions, a calculation procedure was created:

- The sensible heat is small in comparison with the latent heat of the condensing vapour; therefore, it is negligible in the calculations. The saturation temperature (T_{di}) of each column sub-volume is calculated depending on the local saturation pressure (Antoine Equation).
- Both continuous phase and dispersed phase mass flow rates are assumed constant along the column. This can be made reasonably acceptable by means of a constant holdup ratio along the column height, which has been demonstrated for the direct contact evaporator [27-29].
- The heat losses from the direct contact column to the environment are ignored.

The experimental data is analysed using a simple energy balance, exploiting the assumptions above. The simple energy balance for latent heat dominating in three-phase direct contact condenser is:

$$Q_i = \dot{m}_c C_{pc} (T_{co} - T_{ci})_i = \dot{m}_{di} h_{fg} \quad (13)$$

Where Q_i represents the heat transfer rate for each sub-volume of the direct contact condenser. The volumetric heat transfer coefficient is obtained as:

$$U_{vi} = \frac{Q_i}{A \Delta Z_i (\Delta T_{lm})_i} \quad (14)$$

Where $(\Delta T_{lm})_i$ denotes the log-mean temperature difference for each sub-volume.

Accurate prediction of U_{vi} through the direct contact heat exchange process requires knowledge of the temperature profile and consequently the temperature driving force. This quantity, however, varies due to backmixing and the non-uniform two-phase bubbles' size and distribution in the column. This can be solved by using the log-mean temperature

difference, which can be measured accurately in a practical exchanger. A first assumption of the log-mean temperature difference for our case might be written as:

$$(\Delta T_{lm})_i = \frac{(T_{co} - T_{ci})_i}{\ln \left(\frac{T_{di} - T_{ci}}{T_{di} - T_{co}} \right)_i} \quad (15)$$

Equations (13-15) can be combined to yield:

$$U_{vi} = \frac{\dot{m}_c c_{pc}}{A \Delta Z_i} \ln \left[\frac{(T_{di} - T_{co})_i + \left(\frac{\dot{m}_{di}}{\dot{m}_c} \right) \frac{h_{fg}}{c_{pc}}}{(T_{di} - T_{co})_i} \right] \quad (16)$$

Figures (2-4) show the variation of the volumetric heat transfer coefficient along the direct contact condenser height at a constant continuous mass flow rate and initial temperature of the dispersed phase. There is clearly a steady decrease in the volumetric heat transfer coefficient along the column height. This is consistent with previous experimental studies of one-component direct contact condensation in a packed column [22]. This reduction in volumetric heat transfer coefficient can be accounted for by the increase in the heat transfer resistance that builds up due to the gradual condensation of the bubbles as they move along the column. The condensate layer remains confined within the mother bubble during the course of the direct contact condensation, which leads to an increased heat transfer resistance. However, the presence, and increasing influence, of non-condensable gas could be another reason for this decreasing in U_v [22]. Furthermore, Figs. (2-4) indicate that the volumetric heat transfer coefficient increases when the dispersed phase mass flow rate is increased, as well as with when the continuous phase mass flow rate is increased.

The dependency of the volumetric heat transfer coefficient on the mass flow rate ratio for different initial dispersed phase temperatures is shown in Fig. 5. It is obvious that there is a linear effect from the mass flow low rate ratio on U_v for all the initial dispersed phase temperatures considered here, over the range of ratios explored in the experiments. An

increase in U_v values is shown to correspond with increased the continuous mass flow rate ratio. Initially at low mass flow rate ratio (achieved through low dispersed phase mass flow rate), the amount of thermal energy provided by vapour is insufficient to raise the continuous phase temperature to a level which significantly reduces the temperature driving force. A higher mass flow rate ratio reduces the temperature driving force by enhancing the condensation process, consequently enhancing the volumetric heat transfer coefficient. The impact of continuous phase mass flow rate, however, is implicitly shown in Fig. 5, which is consistent with a justification mentioned above.

Figures 6 and Fig. 7 are presented to show the effect of the initial dispersed phase temperature on the volumetric heat transfer coefficient for two different mass flow rate ratios. It is clear that the initial temperature does not have a significant effect on U_v . This supports the assumption that latent heat dominates in each direct contact condenser sub-volume. Further studies, in which the vapour has a higher degree of superheat, may display a more profound effect of initial temperature on U_v .

Fig. 2. Experimental volumetric heat transfer coefficient along the direct contact condenser for $T_{di} = 40^{\circ}\text{C}$ for 5 different dispersed phase flow rates

Fig. 3. Experimental volumetric heat transfer coefficient along the direct contact condenser for $T_{di} = 43.5^{\circ}\text{C}$ for 5 different dispersed phase flow rates

Fig. 4. Experimental volumetric heat transfer coefficient along the direct contact condenser

for $T_{di} = 47.5^{\circ}\text{C}$ for 5 different dispersed phase flow rates

(a)

(b)

(c)

Fig. 5. Experimental volumetric heat transfer coefficient versus mass flow rate ratio for three different dispersed phase initial temperature

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352 **Fig. 6.** Effect of mass flow rate ratio and dispersed phase initial temperature on the volumetric
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358 **Fig. 7.** Effect of the initial temperature of the dispersed phase on the volumetric heat transfer
359 coefficient for two different mass flow rate ratios

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4.2. Model validation

The analytical model was developed to predict the variation of the volumetric heat transfer coefficient with height in the direct contact condenser at different mass flow rate ratios and initial temperatures. As shown experimentally above, the heat transfer coefficient increases with increasing mass flow rate ratio and decreases upon moving up the column. Fig. 8 represents the model validation by fitting with present experimental data for three different dispersed phase initial temperatures and five different mass flow rate ratios. In general, a good agreement can be seen between the analytical predictions and the experimental measurements. The maximum divergence can be seen at the vapour inlet the bottom of the direct contact column. This could be due to the unstable zone above the sparger. In this region a range of bubble shapes and sizes could emerge due to the effect of injection pressure and initial condensation. The heat transfer in this area should be unstable and faster than that higher up the column. This may render some of the assumptions outlined above invalid in this small region. The high or maximum temperature difference and the faster bubbles velocity in addition to minimum heat transfer resistance which mainly control by a condensate layer formed due to the condensation progress could be affected the heat transfer. Interestingly, the divergence between the experimental results and analytical prediction seems to be smaller at lower initial temperatures of the dispersed phase. This supports the hypothesis regarding the intensive condensation at a large initial temperature difference.

Above this initial mixing zone, the agreement tends to be excellent between the present analytical model and the experimental results. It can be concluded that the present model is suitable to predict the volumetric heat transfer coefficient for a steady state direct contact condenser.

(a)

(b)

(c)

Fig. 8. Theoretical volumetric heat transfer coefficient validation through comparison of the results from the experiments with the analytical prediction in equation (10).

5. Conclusions:

Experimental measurements and an analytical prediction of the variation of the volumetric heat transfer coefficient with height in a three-phase direct contact condenser has been presented. From the experimental study, the following conclusions can be drawn:

- The volumetric heat transfer coefficient decreases gradually along the height of the direct contact column.
- The variation of the volumetric heat transfer coefficient with column height is shown to be insignificantly influenced by the mass flow rate ratio.
- The volumetric heat transfer coefficient increased with increasing the continuous phase mass flow rate.
- No significant effect of the dispersed phase initial temperature on the volumetric heat transfer coefficient is shown within the present experiments temperature range.
- A good agreement is obtained between the present analytical model and the experimental results, especially at a distance far from the vapour injection point.

Nomenclatures

A	cross section area(m^2)
a	two-phase radius (m)
C _p	Heat capacity, $kJ/kg \cdot ^\circ C$
D	two-phase bubble diameter (m)
H	total direct contact column height (m)
h_d	surface heat transfer coefficient ($kW/m^3^\circ C$)
h_{fg}	Latent heat of condensation, kJ/kg
k_c	continuous phase thermal conductivity ($kW/m^\circ C$)
kv	function appears in Eq.(8b)

419	$\dot{m}$	mass flow rate (kg/min)
420	n_o	initial two-phase bubbles number per cubic meter
421	Q	heat transfer rate, (kW)
422	R	mass flow rate ratio
423	T	temperature,(°C)
424	ΔT	temperature difference,(°C)
425	ΔT_{lm}	log-mean temperature difference (°C)
426	U_r	relative velocity (m/s)
427	U_v	volumetric heat transfer coefficient, $kJ/m^3.s.C^o$
428	Z	position (m)
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430	Greek symbols	
431	β	function appears in Eq.(5)
432	ϵ	thermal diffusivity (m^2/s)
433	ρ_v	vapour phase density (kg/m^3)
434	ϕ	holdup
435	Subscripts:	
436	c	continuous phase
437	d	dispersed phase
438	di	initial value
439	ci	initial condition of the continuous phase
440	ci	initial condition of the dispersed phase
441	i	section (1, 2,...)
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