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Direct-contact condensation heat transfer on downcommerless trays for steam–water system

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Abstract

The paper deals with some of the results obtained during experimental research of the direct-contact condensation in cascade column. Downcommerless trays were used to obtain contact between water and steam in DN 300 column and experimental work included 198 runs at atmospheric pressure. Since no equation for the intensity of heat transfer was found in the open literature, original equation was established in the following form

$$\mathrm{NTU_L} = f(F_{\mathrm{LG}})$$

where $\mathrm{NTU_L}$ is number of transfer units for liquid phase and F_{LG} is two phase kinetic energy parameter.

Keywords: Heat transfer; Direct-contact condensation; Downcommerless tray; Number of transfer units; Kinetic energy parameter

1. Introduction

Direct-contact heat exchange has found its application in many industries, and the process of heat transfer between phases in direct contact is analyzed and described in many engineering books and articles, such as [1,2], etc.

Downcommerless trays (also known as trays without downcommers or dual-flow trays) have been used in mass-transfer operations (distillation, absorption and desorption columns) for over sixty years [3,4]. These kinds of trays occupy the whole column cross-section and gas/vapor and liquid flow counter-currently through the same openings.

Downcommerless trays offer several attractions compared to other types of trays: they have lower pressure drop, versions with large openings can be used in fouling services, and they are relatively inexpensive in terms of both material cost and installation cost. On the other hand they have relatively narrow operating range (their turndown ratio tends to be low).

As well as for above mentioned mass-transfer operations, downcommerless trays can be used for direct-contact condensation in columns, so they can be used for (thermal) deaeration. Deaerators role is to remove air and other gases from liquid, for example: from boiler feed water prior to its introduction to a boiler or from the feed for evaporation processes. The process of deaeration (desorption) occurs spontaneously during the heating of liquid. In most cases liquid is pure water, water solution or water based mixture, so heating is usually done with steam introduced at the bottom of the column, while liquid flows from the top to the bottom. Since the phases are in intimate contact, steam condenses on liquid surface. In order to increase the contact surface and the intensity of heat transfer between phases, various contact devices are built into the column, and among other shower trays (decks) are often used.

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Nomenclature

c_L	liquid specific heat capacity at constant pressure, J/(kg K)
F_{LG}	kinetic energy parameter
G_{in}	mass flow rate of vapor (steam) at the outlet, kg/s
L	mass flow rate of liquid at the certain cross-section of column, kg/s
L_{in}	mass flow rate of liquid at the inlet of the column, kg/s
L_{out}	mass flow rate of liquid at the column outlet, kg/s
L_m	arithmetic mean liquid mass flow rate over the tray, kg/s
NTU_L	number of transfer units for liquid phase
p_{cond}	pressure in the column i.e. pressure of condensation, Pa
S_{LG}	transfer surface area, m^2
t_4	temperature of liquid on the lower tray, °C
t_{cond}	temperature of condensation at pressure p_{cond}, °C
t_L	temperature of liquid, °C
$t_{L,in}$	temperature of liquid at the column inlet, °C
$t_{L,out}$	liquid temperature at the column outlet, °C
α_L	heat transfer coefficient in liquid phase, W/(m^2 K)
ΔB	dispersion of water flow rate
ρ_L	density of liquid, kg/m^3
ρ_G	density of vapor (steam), kg/m^3
Θ	correlation ratio
Θ	$=\sqrt{1-\frac{\sum_{i=1}^{n}\left(z_i-z_i^c\right)^2}{\sum_{i=1}^{n}\left(z_i-z_{av}\right)^2}}$
z_i	measured value of the parameter z in ith experimental run
z_i^c	correlated value of the parameter z in ith experimental run
z_{av}	average value of z for complete set of experimental data
z_{av}	$=\frac{\sum_{i=1}^{n}z_i}{n}$
n	number of experimental runs
Δ_{av}	standard deviation
Δ_{av}	$=\sqrt{\frac{\sum_{i=1}^{n}\left(\frac{z_i-z_i^c}{z_i}\right)^2}{n}}$

Generally speaking, the column diameter, tray spacing and the number of trays are the most important parameters for trayed column design [5]. Column diameter and tray spacing depend mostly on vapor and liquid flow rates and total column pressure and relying on recommendations in the open literature these parameters can be estimated with considerable accuracy. The most difficult problem in column design is the estimation of the number of trays which is directly connected with the intensity of heat and/or mass-transfer process between phases in contact.

It can be said that, since the first descriptions of barometric condensers, about a century ago, two approaches for calculation of the direct-condensation apparatuses have been developed:

- empirical approach that can be directly used for estimation of the number of trays (for example procedure given in [6] or in [7] for barometric condensers with shower trays);
- approach based on fundamental equations of heat transfer for cases when the shape and the size of the contact surface is known (drops, jets, bubbles, etc.).

There was no data found in the open literature concerning heat transfer during direct-contact condensation on downcommerless trays, although direct-contact condensation is widely studied problem, as presented in [1] or [8]. On the other hand there are few data about the Murphree tray efficiency of downcommerless trays, gathered primarily on distillation columns. Authors from the former USSR usually presented their work in the form of criteria equations (Sherwood number as a function of Reynolds, Schmidt, Froude and Weber numbers and geometrical parameters); the wide review can be found in [9]. Other approach, like in [10] or [11], was to present Murphree tray efficiency on diagrams that can be used for the trays with similar geometrical parameters and for mixtures similar to those used in experimental work.

The purpose of the research presented here is to establish reliable procedure for the design of the boiler feedwater deaerators with downcommerless trays.

2. Experimental apparatus

Experimental apparatus is presented in Fig. 1. Carbon steel column equipped (pos. 1) with three trays is used as a contact condenser. Column nominal diameter is DN 300 (OD/ID diameter 323.9/309.7 mm). Downcommerless sieve trays (upper tray—pos. 2, middle tray—pos. 3, lower tray—pos. 4) were used as contact devices (each with two glasses mounted on the shell of the column for visual observations). Water was distributed over the upper tray by means of perforated half tube distributor (pos. 5). Bottom of the column (ID 600 mm, height 1000 mm, pos. 11) was used as water reservoir with adjustable water level. During the experimental work water level was held on 500 mm from the lower tray using DN 80 valve (pos. 19). From the bottom of the column water flows into 60 m^3 concrete water basin. Water pump (pos. 6) transports water through

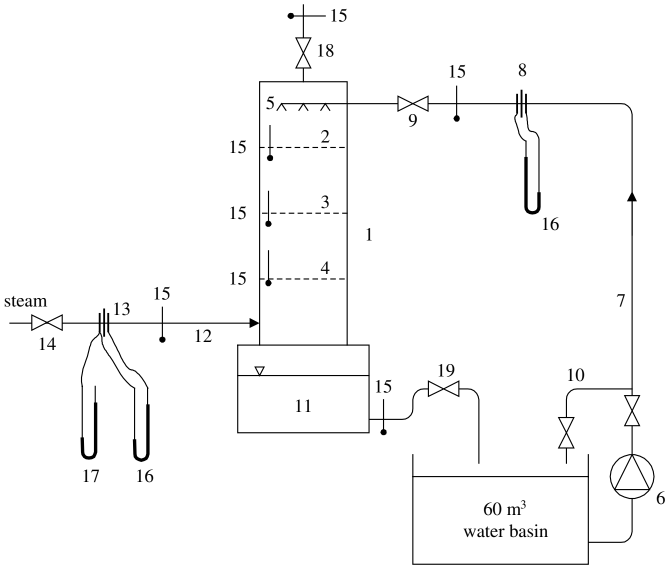

Fig. 1. The experimental installation.

DN 50 pipeline (OD/ID 57/51.2 mm—pos. 7) to the column. Water flow rate is measured and controlled manually using DN 50 valve (pos. 9) and DN 50 bypass line with valve (pos. 10). Steam is taken from steam distributor with DN 150 mm pipeline (OD/ID 159/150 mm—pos. 12), and its flow rate is controlled manually by valve (pos. 14). On the top of the column relief tube DN 20 was mounted together with valve (pos. 18).

Experimental apparatus was placed in district heating heat plant with complete water treatment system, so only demineralized water was used in the experimental work.

The flow rates of water and steam are measured by orifice flow meters in accordance with ISO 5167-1:1991 standard. Differential manometers (pos. 16) and manometer (pos. 17) are with mercury. Temperatures of water and steam at the inlet and outlet of the column are measured using platinum resistant thermometers PT 100 (accuracy 0.1°C—pos. 15).

Temperatures were measured on each tray using PT 100 temperature probe mounted on the tray as presented in Fig. 2. Probes were mounted inside the shell which had a

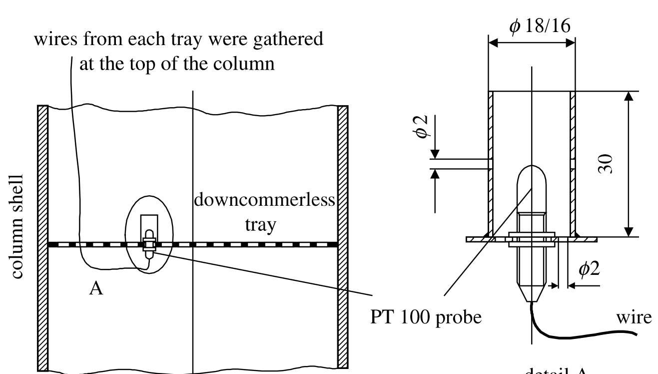

Fig. 2. Position of the PT 100 thermometers on tray.

role to accumulate certain amount of liquid around the probe. The dimensions of the shell and the diameter of holes in shell (required for the water exchange around probe) were established after series of trials on water–air system.

For each experimental run the following parameters were measured:

- water flow rate at the inlet of the column L_{in};
- steam flow rate at the inlet of the column G_{in};
- water temperature at the inlet $t_{L,in}$;
- water temperature at the outlet $t_{L,out}$;
- temperatures on each tray (t_4 will be of interest in the analysis of experimental results).

Three downcommerless trays placed in the column (ID 309.7 mm) are equal, made of cupper sheet (thickness 1 mm). Orifices (holes) on the trays are drilled, having diameter of 8 mm. The holes are placed in the corners of equilateral triangles at distances between centers (pitch) of 12 mm. Total number of holes is 539 and the ratio of hole area to column cross-section area is 36.0%.

2.1. The behavior of column during experimental work

For each experimental run measurements were performed after reaching the stationary conditions (controlled by the measurements of temperatures, pressure and flow rates). There were 213 experimental runs. Water flow rates at the column inlet varied from 3.0 m^3/h to 13.6 m^3/h (39 $m^3/(m^2\,h)$ to 180 $m^3/(m^2\,h)$) and the steam flow rate was in range 203–1070 kg/h. The column worked at atmospheric pressure (100.1–101.8 kPa). Temperature of inlet water was in range 20–30 °C, and the temperature at the bottom of the column (water outlet temperature) was 39–98 °C, while the steam at the inlet was slightly superheated (102–117 °C).

Experimental work showed that for water flow rates less than 40 $m^3/(m^2\,h)$, working regimes were unpurposeful, because the height (accumulation) of liquid phase on tray was very small. In these cases steam easily passed through the liquid phase and no froth was formed, so steam passed through the column and through relief tube.

When downcommerless trays are used as contact device in columns the shape and the size of contact surface is not exactly known. Generally, transfer between phases in columns with downcommerless trays, occurs in two zones:

- two-phase bubbling zone (froth) on tray, where the gas phase is dispersed, and
- in the "settling zone" (space between tray and upper surface froth) where the liquid phase is dispersed and falls down through the openings on tray.

During the experimental work it was noticed that in the case of small steam flow rates heat transfer occurred only on water jets formed below the lower tray. When the steam

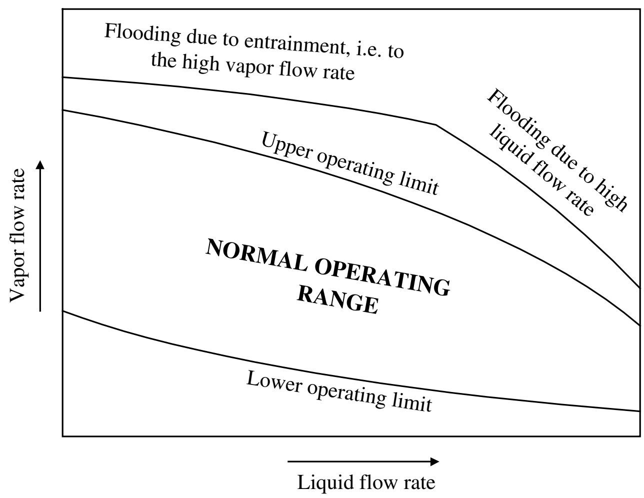

Fig. 3. General form of the working diagram for downcommerless trays.

flow rates were increased, bubbles were formed in liquid layer on tray, so heat was transferred both on jets and on froth on the tray.

Two more observations are of significance.

According to literature it was expected that non-condensed gases and a little amount of steam would flow continuously trough the relief tube, at relatively constant flow rate and temperature. This was not happening during the experimental work. In stead, the mixture of gas and steam was for a few seconds flowing out of the column, but in another few seconds ambient air was flowing into the column. The reason for such phenomenon is the instability of two-phase flow regimes on downcommerless trays.

The operation of column was very stable until the temperature of water at the outlet ($t_{L,out}$) approaches temperature of condensation (t_{cond}). In such cases (usually when $t_{cond} - t_{L,out} < 5$ °C) even the small variations in flow rates of water and steam caused instabilities which lead to the flooding of the column. Working regime with water flow rate of 85 $m^3/(m^2\,h)$ is empirically established as border regime for two types of flooding. For smaller water flow rates steam was breaking through the water layer on all three trays to the top of the column and also though the relief tube. In case of greater water flow rates, water was accumulated first at the bottom tray and then at upper trays. The first case is flooding of column due to high steam flow rate and former case is flooding due to high water flow rate. This phenomenon is opposite to one described in [4], so the diagram presented in Fig. 3 can be assumed as general for downcommerless trays.

3. Mathematical model of direct-contact condensation heat transfer

The following mathematical model considers the case of direct-condensation heat transfer for one-component vapor–liquid system in counter-current column (liquid flows downwards and vapor upwards). Inlet vapor is usually saturated or slightly superheated. When brought in contact with cold liquid, both latent and sensible heat transfer occurs. The amount of sensible heat transferred between phases is usually small, compared to the amount of latent heat and further on it will be neglected.

When fluid flows through the apparatus, part of the fluid energy is spent to overcome flow resistances. Considering that this part of the energy is negligible (in practice less than 1% compared to the total fluid energy), for engineering calculations two conclusions can be drawn:

- the change of total energy is equal to the change in enthalpy of fluid, which means that the total energy balance is reduced to the heat balance equation;
- pressure in apparatus is constant and equal to the pressure of the steam at the inlet (p_{cond} Pa) which means that temperature of condensation (t_{cond} °C) is also known.

Further simplification is that for engineering calculations it is useful to consider fluid thermo-physical properties constant, and to neglect heat losses through the shell of apparatus and the influence of the desorption of dissolved gases (CO_2, air) on heat transfer.

Heat balance for differential section of the liquid phase (Fig. 4) is

$$L \cdot c_L \cdot dt_L + dL \cdot c_L \cdot t_L = \alpha_L \cdot (t_{cond} - t_L) \cdot dS_{LG} \tag{1}$$

The increment of liquid flow rate due to condensation is usually treated as negligible compared to the whole liquid flow rate, which means that $dL = 0$ and liquid flow rate is constant ($L = \text{const.}$). Therefore Eq. (1) is simplified to the form that can be easily integrated

$$\int_{t_{L,out}}^{t_{L,in}} \frac{dt_L}{t_{cond} - t_L} = \int_0^{S_{LG}} \frac{\alpha_L}{L \cdot c_L} \cdot dS_{LG} \tag{2}$$

For engineering purposes it is convenient to assume that the intensity of heat transfer does not change much along the heat transfer surface, so the calculations can be carried out with the mean value of heat transfer coefficient (i.e. $\alpha_L = \text{const.}$), and after integration the number of transfer units for liquid phase is

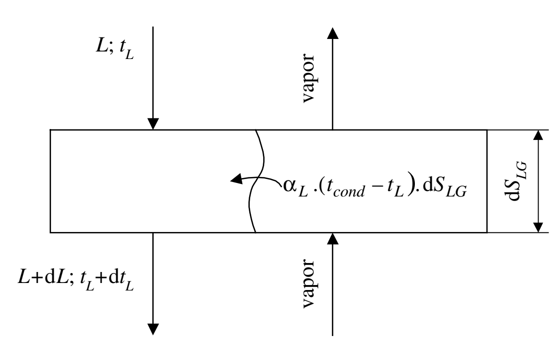

Fig. 4. Differential section of apparatus.

$$\ln\frac{t_{\mathrm{cond}}-t_{\mathrm{L,in}}}{t_{\mathrm{cond}}-t_{\mathrm{L,out}}}=\frac{\alpha_{\mathrm{L}}\cdot S_{\mathrm{LG}}}{L\cdot c_{\mathrm{L}}}=\mathrm{NTU_L} \tag{3}$$

In general the number of transfer units can be used as a measure of the intensity of heat transfer and depends on the manner of formation, shape and the size of the contact surface, phase flow rates and their thermo-physical properties.

4. Analysis of the experimentally obtained results

Necessary condition for the analysis of the experimental results is that stationary working regime in apparatus is established, which means that the values of process parameters (flow rates, temperatures, pressure, etc.) are unchangeable over the sufficient period of time. During experimental runs, the stationarity of working regime was controlled by the measurement of the process parameters and by the following analyses.

For stationary working conditions equations of material and heat balances, for complete condensation of steam, are

$$L_{\mathrm{in}}+G=L_{\mathrm{out}} \tag{4}$$

$$L_{\mathrm{in}}\cdot c_{\mathrm{L}}\cdot t_{\mathrm{L,in}}+G\cdot h_{\mathrm{G}}=L_{\mathrm{out}}\cdot c_{\mathrm{L}}\cdot t_{\mathrm{L,out}} \tag{5}$$

There are two water flow rates at the outlet of the column that can be calculated from Eqs. (4) and (5)

$$L_{\mathrm{out}}\ (4)=L_{\mathrm{in}}+G \tag{6}$$

$$L_{\mathrm{out}}\ (5)=\frac{L_{\mathrm{in}}\cdot c_{\mathrm{L}}\cdot t_{\mathrm{L,in}}+G\cdot h_{\mathrm{G}}}{c_{\mathrm{L}}\cdot h_{\mathrm{L,out}}} \tag{7}$$

Arithmetic mean value of water flow rate at the outlet is

$$L_{\mathrm{out}}=\frac{L_{\mathrm{out}}\ (4)+L_{\mathrm{out}}\ (5)}{2} \tag{8}$$

and dispersion of water flow rate can be calculated

$$\Delta B=\frac{\sqrt{\left[L_{\mathrm{out}}\ (4)-L_{\mathrm{out}}\right]^2+\left[L_{\mathrm{out}}\ (5)-L_{\mathrm{out}}\right]^2}}{L_{\mathrm{out}}} \tag{9}$$

In engineering practice it is commonly assumed that dispersion of 5% in balance is acceptable. For 213 working regimes resulted from experimental work the dispersion was less that 5%. Since 15 regimes were with significant entrainment or with flooding of column, they were omitted from further analysis. So 198 working regimes are of significance (186 of them was with dispersion less than 3%).

Measurements showed that complete heat transfer occurred only on the bottom tray. Only in cases of flooding of bottom tray, the middle tray started to work. Such regimes were, due to flooding very unstable so the experiment was stopped in such cases.

As stated before the number of transfer units can be used for determining the intensity of heat transfer and depends on both fluid flow rates and properties, as well as geometrical parameters of trays. Since one-component system was used in relatively small diapason of temperatures and at atmospheric pressure, there was no possibility to investigate the influence of fluid properties on the number of transfer units. Also, the influence of tray geometrical parameters could not be introduced in analysis because experiments were performed on only one tray.

According to mathematical model the number of transfer units can be calculated using Eq. (3), which rearranged for the bottom tray, has the following form

$$\mathrm{NTU_L}=\ln\frac{t_{\mathrm{cond}}-t_4}{t_{\mathrm{cond}}-t_{\mathrm{L,out}}} \tag{10}$$

Having on mind that the number of authors concluded that the kinetic energy parameter is the key parameter for fluidodynamic, but also mass-transfer design of various types of columns, the influence of this parameter on the intensity of heat transfer was analyses. For this purpose kinetic energy parameter is defined as

$$F_{\mathrm{LG}}=\frac{L_{\mathrm{m}}}{G_{\mathrm{in}}}\cdot\sqrt{\frac{\rho_{\mathrm{G}}}{\rho_{\mathrm{L}}}} \tag{11}$$

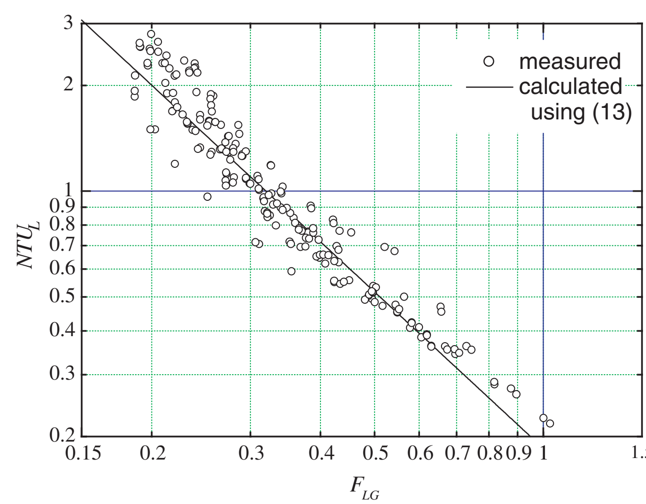

Fig. 5. Experimentally obtained and calculated values of $\mathrm{NTU_L}$ vs. F_{LG}.

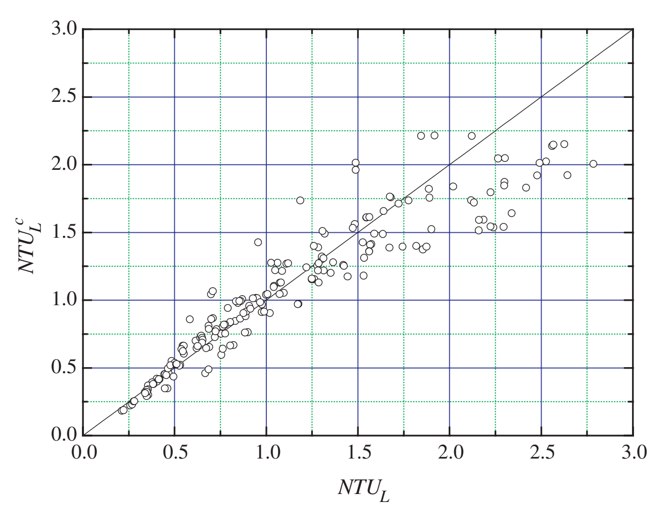

Fig. 6. Correlation field for Eq. (13).

where G_{in} is the inlet flow rate of steam and L_{m} is arithmetic mean water flow rate over tray

$$L_{\text{m}} = \frac{L_{\text{in}} + L_{\text{out}}}{2} \tag{12}$$

Regression analysis resulted in following equation:

$$\text{NTU}_{\text{L}}^{\text{c}} = 0.185 \cdot F_{\text{LG}}^{-1.48} \tag{13}$$

with statistical parameters $\Theta = 0.925$ and $\Delta_{\text{av}} = 15.9\%$, so it can be concluded that Eq. (13), in diapason of performed measurements, can be successfully used.

Measured and calculated values of the number of transfer units are presented in Fig. 5, and correlation field of (13) in Fig. 6.

5. Conclusion

Subject of this article is a research of the intensity of heat transfer during direct-contact condensation on downcommerless sieve trays. In order to establish a reliable procedure for deaerator column design experimental research was performed on DN 300 (ID 309.7 mm) column with three identical trays (the perforated area being 36.0% of the tray area). Experimental runs (198) were found to be competent for regression analysis.

The number of transfer units for liquid phase (NTU_{L}) depends is mainly on kinetic energy parameter (F_{LG}). Regression analyses gave Eq. (13) with acceptable statistical parameters for practical engineering purposes i.e. for dimensioning of deaerators with downcommerless trays.

Concerning that Eq. (13) is the result of experimental work with only one two-phase system, one-component (steam–water) and only one tray, there was no possibility to investigate the influence of fluid properties and tray geometry on the number of transfer units. To obtain more general equation, which could be acceptable for all cases (without obstacles listed before), a large scale of additional experimental work should be done. For this reason, hereby presented results give guidance for further research on parameters that characterize heat transfer during direct-contact condensation in columns with downcommerless trays.

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