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Heat transfer rate of direct-contact condensation on baffle trays

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ABSTRACT

The paper deals with the experimentally obtained results of the direct-contact condensation of steam in contact with water in cascade column. Baffle trays were used to obtain contact between water and steam in DN 300 column at atmospheric pressure. Testing of Chernobilski equation showed poor correlation compared to the experimental results so we have found the following form of heat transfer rate

$$Nu = 5.8 \cdot 10^{-6} \cdot Re^{5/3} \cdot Pr^{1/3} \cdot Fr^{-2/3}$$

which could be recommended for heat transfer calculations for this kind of apparatuses.

1. Introduction

Direct-contact condensation has found its application in many industries, and most frequently used apparatuses for this operation are deaerators and barometric condensers. This heat transfer operation is analyzed and described in many engineering books and articles, such as [1,2], etc.

Deaerators (degassers, thermal deaerators) are used to remove air and other dissolved gases from liquid, for example: from boiler feed water prior to its introduction to a boiler or from the feed for evaporation processes. Deaeration (desorption) occurs spontaneously due to the heating of liquid. In most cases liquid is pure water, water solution or water based mixture.

On the other hand barometric condensers are used in various vacuum operations, such as evaporation, drying, vacuum refrigeration, etc. This type of self draining condenser requires a barometric leg, approximately 10 m high, in order to remove water by gravity and overcome friction losses.

In both kinds of apparatuses various types of contact devices are used in order to achieve high level of heat transfer. The simplest contact device is called baffle or segmental or shower tray (cascade, deck, plate). In apparatuses equipped with baffle trays liquid flows downward from top to bottom tray and vapor flows counter-currently upward, like presented in Fig. 1. Vapor contacts liquid while it showers from the tray, penetrating through the liquid curtain. Weir on tray improves the liquid distribution in the shower [3], and at the top of the column there is drop eliminator.

Generally speaking, the column diameter, tray spacing and the number of trays are the most important parameters for trayed column design [5]. 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.

Basically there are two approaches for design of direct-condensation apparatuses [8]:

- empirical approach that can be directly used for estimation of the number of trays (for example procedure given in [6] for barometric condensers with baffle 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).

Although direct-contact condensation is widely studied problem, as presented in [1] or [4], there was only one design procedure found in the open literature for the estimation of the heat transfer rate on baffle trays. Named by the first author Chernobilski [6], it was developed about the 1950s and it is widely cited in literature in Russia and Eastern Europe.

2. Mathematical model of direct-contact condensation heat transfer with closed form solution

The following mathematical model considers the case of direct-condensation heat transfer for one-component vapor–liquid sys-

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Nomenclature

c_L liquid specific heat capacity at constant pressure (J/kg K)
d_e equivalent diameter of the liquid curtain (m)
Fr_w Froude number based on $w_{L,w}$

$$Fr_w = \frac{g \cdot d_e}{w_{L,w}^2}$$

Fr Froude number based on $w_{L,av}$

$$Fr = \frac{g \cdot d_e}{w_{L,av}^2}$$

g acceleration due to gravity (m/s^2)
G_{in} mass flow rate of vapor (steam) at the column inlet (kg/s)
G_{out} mass flow rate of vapor (steam) at the column inlet (kg/s)
h_G specific enthalpy of the vapor (J/kg)
$h_{L,ow}$ height of the water crest over the weir (m)
L mass flow rate of liquid at the certain cross section of column (kg/s)
L_{disp} dispersion of water flow rate
L_e mass flow rate of the water at the end of the liquid curtain (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_s mass flow rate of the water falling over the weir (kg/s)
NTU_L number of transfer units for liquid phase
Nu Nusselt number

$$Nu = \frac{\alpha_L}{\lambda_L \cdot d_e}$$

p_{cond} pressure in the column i.e. pressure of condensation (Pa)
Pr Prandtl number

$$Pr = \frac{\mu_L \cdot c_L}{\lambda_L}$$

Q_t estimated heat power of single tray (W)
Q_{ap} estimated heat power of apparatus (W)
Re Reynolds number

$$Re = \frac{w_{L,av} \cdot d_e \cdot \rho_L}{\mu_L}$$

S_{LG} heat transfer surface (surface of liquid–vapour contact) (m^2)
t_{cond} temperature of condensation at pressure p_{cond} (°C)
t_L temperature of liquid (°C)
$t_{L,e}$ temperature of liquid at the end of the liquid curtain (°C)
$t_{L,in}$ temperature of liquid at the column inlet (°C)
$t_{L,out}$ liquid temperature at the column outlet (°C)
$t_{L,s}$ temperature of liquid at the beginning of the liquid curtain (°C)
T_h height of the heat transfer surface (tray spacing) (m)
$w_{L,av}$ free fall average velocity (kg/s)
$w_{L,w}$ free fall starting velocity (velocity of the water falling over the weir) (kg/s)
W weir width (m)
α_L heat transfer coefficient in liquid phase (W/(m^2 K))
δ_L depth of liquid curtain (m)
ρ_L density of liquid (kg/m^3)
λ_L thermal conductivity of liquid (W/(m K))
μ_L dynamic viscosity of liquid (Pa s)
η coefficient of the contraction of free fall
Θ correlation ratio

$$\Theta = \sqrt{1 - \frac{\sum_{i=1}^{n}(z_i - z_i^c)^2}{\sum_{i=1}^{n}(z_i - z_{av})^2}}$$

Δ_{av} standard deviation

$$\Delta_{av} = \sqrt{\frac{\sum_{i=1}^{n}\left(\frac{z_i - z_i^c}{z_i}\right)^2}{n}}$$

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

tem in counter-current column. Inlet vapor is usually saturated or slightly superheated, and during its contact with cold liquid, both latent and sensible heat transfer takes place, but the transfer of sensible heat between phases is usually neglected.

Assuming that overcoming of flow resistances in apparatus takes neglectable part of fluid energy (in practice less than 1%) operation is isobaric (pressure in apparatus is equal to the pressure of the inlet steam p_{cond}, Pa, which defines the temperature of condensation t_{cond}, °C), and the total energy balance can be reduced to the heat balance equation. Further more heat losses through the shell of apparatus and the influence of desorption of dissolved gases (CO_2, air) on heat transfer are usually neglected.

Heat balance for differential section of the liquid phase (Fig. 2) is

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

In order to obtain the closed form solution the increment of liquid flow rate due to condensation is usually treated as negligible ($dL = 0$), so the liquid flow rate is constant ($L = \text{const}$). Therefore, Eq. (1) is simplified to the form that can be easily integrated:

$$\int_{t_{L,s}}^{t_{L,e}} \frac{dt_L}{t_{cond} - t_L} = \int_0^{S_{LG}} \frac{\alpha_L}{L \cdot c_L} \cdot dS_{LG} \quad (2)$$

For engineering purposes, it is convenient to assume that fluid thermo-physical properties can be treated as constant, and 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. $\alpha_L = \text{const}$). The closed form solution is usually expressed introducing the number of transfer units for liquid phase:

$$\ln \frac{t_{cond} - t_{L,s}}{t_{cond} - t_{L,e}} = \frac{\alpha_L \cdot S_{LG}}{L \cdot c_L} = NTU_L \quad (3)$$

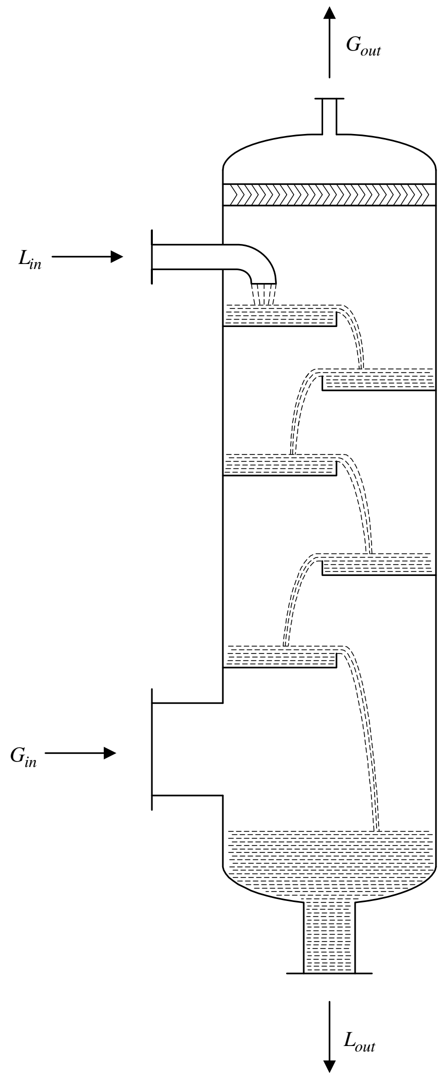

Fig. 1. Condenser with baffle trays.

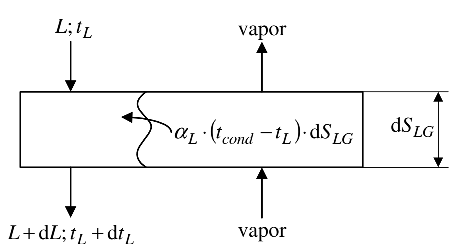

Fig. 2. Differential section of the apparatus.

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

3. Literature data on heat transfer on baffle trays

The design procedure of Chernobilski [6] is based on calculation of the number of transfer units for each tray using:

$$\mathrm{NTU_L} = 0.0668 \cdot Fr_{\mathrm{w}}^{0.2} \cdot \left(\frac{T_{\mathrm{h}}}{d_{\mathrm{e}}}\right)^{0.7} \tag{4}$$

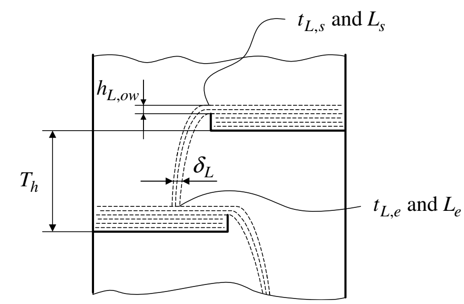

Fig. 3. Basic geometrical parameters, flow rates and temperatures.

Basic geometrical parameters for Chernobilski procedure are presented in Fig. 3. The velocity of the water falling over the weir (free fall starting velocity) is

$$w_{\mathrm{L,w}} = \frac{L_{\mathrm{s}}}{\rho_{\mathrm{L}} \cdot W \cdot h_{\mathrm{L,ow}}} \tag{5}$$

where the height of the water crest over the weir is

$$h_{\mathrm{L,w}} = \left(\frac{3}{2} \cdot \frac{L_{\mathrm{s}}}{\eta \cdot W \cdot \rho_{\mathrm{L}} \cdot \sqrt{2 \cdot g}}\right)^{2/3} \tag{6}$$

and $\eta = 0.63$ is the liquid free fall contraction coefficient.

Equivalent diameter is calculated using

$$d_{\mathrm{e}} = \frac{2 \cdot W \cdot \delta_{\mathrm{L}}}{W + \delta_{\mathrm{L}}} \tag{7}$$

where the width of the water curtain is

$$\delta_{\mathrm{L}} = \frac{L_{\mathrm{s}}}{\rho_{\mathrm{L}} \cdot W \cdot w_{\mathrm{L,av}}} \tag{8}$$

and the average velocity of water falling between trays is

$$w_{\mathrm{L,av}} = \frac{w_{\mathrm{L,w}} + \sqrt{w_{\mathrm{L,w}}^2 + 2 \cdot g \cdot T_{\mathrm{h}}}}{2} \tag{9}$$

In Billet [7] gave the procedure for estimation of column height based on a diagram "derived from experience gained in counter-current condensers with cascades. On an average, industrial-scale direct-contact condensers are designed with five plates in a cascade". It must be noted that this procedure should be used with great caution, and only for rough estimations of a column height, since the description of the limitations of the procedure are not given in [7].

4. Experimental apparatus

Experimental apparatus is presented in Fig. 4. Carbon steel column equipped (pos. 1) with three trays is used as a contact condenser. Column nominal diameter is DN 300 (OD/ID 323.9/309.7 mm), and it is equipped with three baffle trays (top tray – pos. 2, middle tray – pos. 3, bottom tray – pos. 4). Three baffle trays placed in the column are equal, made of copper sheet (thickness 1 mm) with 40 mm weir height and $W = 289$ mm. Tray height between top and middle and middle and bottom tray is 400 mm. Water was distributed over the top 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 bottom 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 DN 50 pipeline (OD/ID 57/51.2 mm – pos. 7) to the column. Water

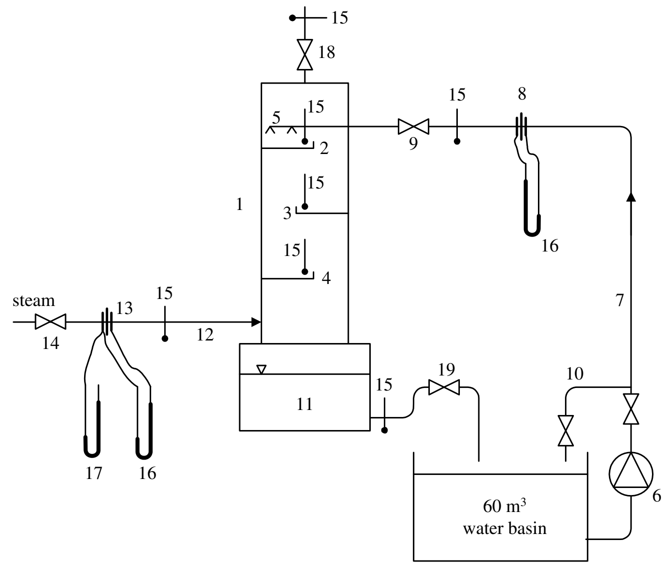

Fig. 4. Experimental apparatus.

flow rate is measured and controlled manually using DN 50 valve (pos. 9) and DN 50 bypass line with valve (pos. 10). Steam is supplied through the DN 150 mm pipeline (OD/ID 159/150 mm – pos. 12), and its flow rate is manually controlled using valve (pos. 14). On the top of the column relief tube DN 15 was mounted together with valve (pos. 18).

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

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 $t_{\mathrm{L,in}}$;
- water temperature at the outlet $t_{\mathrm{L,out}}$;
- temperatures on each tray ($t_{\mathrm{L,s}}$ and $t_{\mathrm{L,e}}$ for each tray).

The flow rates of water (pos. 8) and steam (pos. 13) are measured by orifice flow meters in accordance with [12]. Differential pressure was measured using differential mercury manometer (pos. 16) and mercury manometer (pos. 17) was used for static pressure measurements. 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). Water temperatures on each tray were measured using PT 100 temperature probe submerged in water on tray.

For each experimental run measurements were performed after reaching the stationary conditions (controlled by the measurements of temperatures, pressure and flow rates). Water flow rates at the column inlet varied from 2.4 $\mathrm{m^3/h}$ to 9.8 $\mathrm{m^3/h}$ (flow rate per unit of cross sectional area was 32–130 $\mathrm{m^3/(m^2\,h)}$) and the steam flow rate at the column inlet was up to 915 kg/h. The column worked at atmospheric pressure (99.7–102.0 kPa). Temperature of inlet water was in range 12–37 °C, and the temperature at the bottom of the column (water outlet temperature) was up to 95 °C, while the steam at the inlet was slightly superheated (up to 110 °C).

Experimental work showed that for water flow rate per unit of cross sectional area less than 32 $\mathrm{m^3/(m^2\,h)}$, working regimes were unpurposeful, because the water curtain was not formed over the whole weir width. In such regimes steam easily avoid the penetration of the water, and passed through the column and through relief tube, with very little heat exchanged. In regimes considered in further analysis the complete amount of steam was condensed and there was practically no flow of steam through the relief tube.

5. Analysis of the experimentally obtained results

Necessary condition for the analysis of the experimental results is that stationary working regime in apparatus is established, i.e. that the dispersion of heat balace is in acceptable range. In engineering practice it is commonly assumed that dispersion of 3–7% in heat balance is acceptable [9,10], and in some cases, like in [11], the acceptable dispersion is up to 10%.

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 ($G_{\mathrm{out}} = 0$), are

$$L_{\mathrm{in}} + G_{\mathrm{in}} = L_{\mathrm{out}} \tag{10}$$

$$L_{\mathrm{in}} \cdot c_{\mathrm{L}} \cdot t_{\mathrm{L,in}} + G_{\mathrm{in}} \cdot h_{\mathrm{G}} = L_{\mathrm{out}} \cdot c_{\mathrm{L}} \cdot t_{\mathrm{L,out}} \tag{11}$$

There are two water flow rates at the outlet of the column that can be calculated from Eqs. (10) and (11)

$$L_{\mathrm{out}(10)} = L_{\mathrm{in}} + G_{\mathrm{in}} \tag{12}$$

$$L_{\mathrm{out}(11)} = \frac{L_{\mathrm{in}} \cdot c_{\mathrm{L}} \cdot t_{\mathrm{L,in}} + G_{\mathrm{in}} \cdot h_{\mathrm{G}}}{c_{\mathrm{L}} \cdot h_{\mathrm{L,out}}} \tag{13}$$

Arithmetic mean value of water flow rate at the outlet is

$$L_{\mathrm{out}} = \frac{L_{\mathrm{out}(10)} + L_{\mathrm{out}(11)}}{2} \tag{14}$$

and dispersion of water flow rate can be calculated

$$L_{\mathrm{disp}} = \frac{\sqrt{[L_{\mathrm{out}(10)} - L_{\mathrm{out}}]^2 + [L_{\mathrm{out}(11)} - L_{\mathrm{out}}]^2}}{L_{\mathrm{out}}} \tag{15}$$

For 235 measured working regimes the dispersion was less that 5%, and for another 37 working regimes the balance dispersion was between 5% and 7%.

Measurements showed that the heat transfer occurred mostly on the bottom tray, but the certain amount of heat was transfer on the other two trays. The rough estimation of the heat power of each tray was done using:

$$Q_{\mathrm{t}} \approx \frac{L_{\mathrm{s}} + L_{\mathrm{e}}}{2} \cdot c_{\mathrm{L}} \cdot (t_{\mathrm{L,e}} - t_{\mathrm{L,s}}) \tag{16}$$

and for whole apparatus

$$Q_{\mathrm{ap}} \approx \frac{L_{\mathrm{in}} + L_{\mathrm{out}}}{2} \cdot c_{\mathrm{L}} \cdot (t_{\mathrm{L,out}} - t_{\mathrm{L,in}}) \tag{17}$$

If the heat transferred at any tray was greater than 5% of the whole amount of heat transferred in apparatus the working regime on tray was accepted for further analysis ($Q_{\mathrm{t}}/Q_{\mathrm{ap}} \geqslant 5\%$).

Two hundred and seventy two working regimes that fulfilled described criteria are considered as significant for further analysis:

- 240 regimes were gathered on bottom tray;
- 9 regimes were gathered on middle tray;
- 23 regimes were gathered on top tray.

and the measured number of transfer units was in range $\mathrm{NTU_L} = 0.0324 - 2.18$.

The procedure of Chernobilski [6] was tested against the experimentally obtained data and the statistical parameters are as follows $\Theta = 0$ and $\Delta_{\mathrm{av}} = 90.4\%$, so it can be concluded that this procedure can not be considered as reliable. Further more, it is

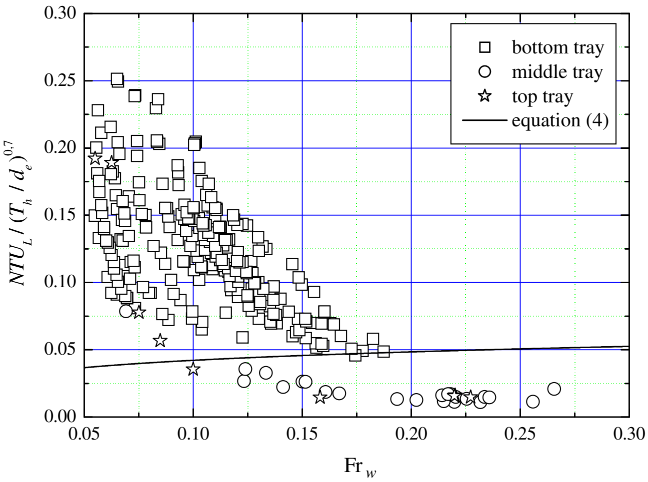

Fig. 5. Experimentally obtained values and Eq. (4) proposed by Chernobilski [6].

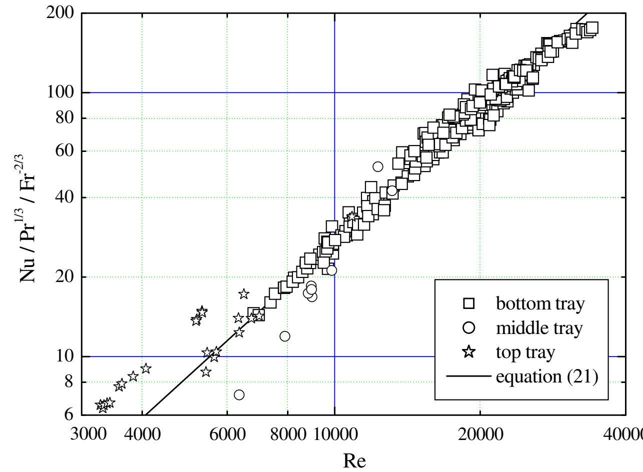

Fig. 6. Experimentally obtained values and Eq. (21).

obvious that the measurements showed that the increase in Fr_w leads to decrease in NTU_L, which is opposite to the character of Eq. (4). Measured and calculated values using the Chernobilski's procedure are presented in Fig. 5.

In order to obtain more reliable design procedure we have tried to find an appropriate correlation using dimensionless numbers to form the equation of the following form:

$$Nu = f(Fr, Re, Pr) \tag{18}$$

Heat transfer coefficient can be obtained from (3):

$$\alpha_L = \frac{NTU_L \cdot c_L}{S_{LG}} \cdot \frac{L_s + L_e}{2} \tag{19}$$

and we have assumed that the heat transfer surface is the surface of water curtain

$$S_{LG} = 2 \cdot T_h \cdot W \tag{20}$$

The heat transfer surface was $S_{LG} = 0.231\ m^2$ for top and middle tray with $T_h = 400$ mm and $S_{LG} = 0.312\ m^2$ for bottom tray with $T_h = 540$ mm. The measured heat transfer coefficient was in range $\alpha_L = 400\text{–}34800\ W/(m^2\,K)$.

Regression analysis resulted with the following equation:

$$Nu = 5.8 \cdot 10^{-6} \cdot Re^{5/3} \cdot Pr^{1/3} \cdot Fr^{-2/3} \tag{21}$$

with statistical parameters $\Theta = 0.983$ and $\Delta_{av} = 13.3\%$, so it can be concluded that the Eq. (21), in diapason of performed measurements, can be successfully used. It must be noted that the dimensionless numbers obtained during measurements were in range $Re = 3280\text{–}34100$, $Fr = 0.0237\text{–}0.135$, $Pr = 2.43\text{–}8.52$ and $Nu = 55.8\text{–}2000$. Measured values of the number of transfer units and the ones calculated using Eq. (21) are presented in Fig. 6.

6. Conclusion

Subject of this article is a research of the intensity of one-component heat transfer during direct-contact condensation on baffle trays. In order to establish a reliable procedure for deaerator column design experimental research was performed on DN 300 (ID 309.7 mm) column equipped with three identical trays. Two hundred and seventy two working regimes were found to be relevant for regression analysis.

Since the testing of design procedure of Chernobilski [6] against experimental data showed poor correlation the new form of equation was established. It was found that the Eq. (21) has acceptable statistical parameters for practical engineering purposes i.e. for design of deaerators with baffle trays.

Concerning that Eq. (21) is the result of experimental work with only one two-phase system, one-component (steam-water) and only one tray design, at this point we were not able to investigate the influence of fluid properties and tray geometry on the number of transfer units. 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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