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The object of this study was the processes of heat and mass transfer in a column heat exchanger with direct phases contact. The investigated problem is related to the need to determine the estimates of the kinetic characteristics of heat exchange during the concentration of solutions in a contact heat exchanger equipped with dual-flow trays. In particular, it was assumed that the determination of the influence of the gas velocity in the apparatus and the density of liquid irrigation of the plates, as well as the geometry of the plates, on the kinetic coefficients would make it possible to reveal the patterns of heat and mass exchange between the liquid and the air in contact with it in the column apparatus. It was determined that to increase the intensity of mass transfer in the gas and heat transfer in the liquid, it is necessary to increase the gas velocity and irrigation density. Then the gas velocity and irrigation density have approximately the same effect on the intensity of mass transfer in the gas and heat transfer in the liquid. When studying the effect of the geometry of the plate on the kinetic coefficients of heat transfer in the liquid and mass transfer in the gas, it was established that the value of the portion of the free cross section of the plate has a decisive influence on the value of the considered kinetic coefficients.

A generalized equation was built, which makes it possible not only to calculate the enthalpy transfer coefficient during the interaction of sodium chloride solution with air but also to determine the limiting stage of this process.

The results could be used to design a unit for concentrating a hot solution of sodium chloride by evaporating water during blowing with air in a contact heat exchanger. This would make it possible to obtain crystalline sodium chloride using secondary energy resources and other non-traditional sources of thermal energy

Keywords: contact heat exchanger, dual-flow tray, heat exchange, sodium chloride solution, gas emissions, utilization of thermal waste, non-traditional sources of thermal energy, enthalpy transfer coefficient

DETERMINING HEAT AND MASS EXCHANGE KINETIC IN A COLUMN HEAT EXCHANGER WITH DIRECT PHASES CONTACT

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1. Introduction

Energy-intensive branches of industry, such as metallurgy, electricity, chemistry, are the largest source of thermal emissions, in particular, in the form of hot gases. Ideas about using the heat of these gases date back to the period of the first industrial revolution. However, it quickly became clear that due to the relatively low temperature of thermal waste and the presence of corrosive impurities and dust in them, this is not an easy task.

In particular, the use of surface heat exchangers turned out to be economically unjustified due to the high cost, low efficiency of heat exchange, and significant operating costs. Under these conditions, the first devices appeared in which heat exchange was carried out in direct contact between hot gases and liquid. Such equipment is now called heat exchangers with direct phases contact or contact economizers. Their main advantage is the high efficiency of heat exchange due to the absence of any partitions separating the contacting phases [1]. This causes a relatively low consumption and cost of the metal.

The history of the evolution of knowledge about rather complex processes occurring in contact economizers dates

back more than 100 years [2], but interest in the study of contact heat exchange does not decrease. This is connected, firstly, with the search for technologies in which low-potential heat can be used, and the selection of appropriate equipment for these technologies. Secondly, with the complexity of the mechanism of the heat exchange process between gas and liquid, which are in direct contact. With such a contact, two processes simultaneously occur and influence each other: actual heat exchange (dry heat exchange), caused by the difference in liquid and gas temperatures, and mass exchange due to condensation or evaporation of the liquid (wet heat exchange).

Therefore, it is important to investigate the patterns of heat exchange during direct contact between hot gases and liquid, which can enable high efficiency of heat exchange due to the absence of any partitions separating the contacting phases.

2. Literature review and problem statement

Contact heat exchangers (economizers) are similar to cooling towers in terms of design and process mechanism. According to the type of irrigation device, existing contact

heat exchangers can be divided into two main groups. The first of them includes spraying and film devices. The second is equipment with a chord or fluidized nozzle, as well as with the use of plates.

Fill of contact heat exchangers is in many ways similar to packed column devices with continuous phase contact used in the chemical industry. However, detailed design procedures require knowledge of heat and mass transfer patterns during the formation and rise of superheated bubbles, as well as determination of mass and energy balances in the liquid phase and gas distribution system [3]. The results reported in the cited work are in good agreement with the available experimental data, but the application is possible for a direct contact evaporator operating under a semi-periodic mode.

Most often, the fill has the form of Raschig rings, and it is important to study the thermal characteristics of the plates used to concentrate liquids that are in direct contact with hot gas emissions [4]. To evaluate such characteristics, the enthalpy exchange coefficient was chosen since it takes into account heat exchange both due to the temperature difference and due to evaporation. The results, which made it possible to rank the influencing factors on the intensity of heat exchange, are limited by the studied ranges of the structural parameters of the plates.

The desire to increase the intensity of heat exchange causes the need to improve the design of contact devices. The most promising are conventional film fill [5]. However, it should be noted that the traditional designs of contact economizers described above have a number of significant drawbacks from the point of view of the possibility of use for concentrating solutions. This is the low intensity of heat and mass exchange compared to the devices of the chemical industry and the fact that most typical designs of sprinklers cannot work under conditions of sediment formation.

Of particular interest, from the point of view of eliminating these shortcomings, are tray devices. This is explained by their higher reliability when working with liquids that form deposits. A number of works report studies of liquid cooling processes in devices equipped with dual-flow trays [6, 7]. An interesting structure is described [8] in authors' certificate. This device is equipped with trays, which the authors call "flat sheet countercurrent plates with large holes (KPLH)". In English terminology, such contact devices are called "dual-flow". Trays with holes with a diameter of 0.12 m and a free cross-section of $0.5 \text{ m}^2/\text{m}^2$ are recommended. However, the optimality of this size of holes and free cross-section in the above works is not sufficiently substantiated.

The design of KPLH plates is simple. They can be mounted and dismantled without dismantling the device body. Anti-corrosion coating is easily applied to them. This design is the most promising for use in a contact heat exchanger for concentrating solutions. As the experience of operating such plates in soda production [9] shows, they work stably even when the tray is overgrown with deposits. This property is important under conditions when the concentration of the solution will be accompanied by the crystallization of salts.

Comparing packed and plate type economizers, one can see that the first of them is much more common. This can be explained by the fact that such equipment is used to heat clean water. Under these conditions, the intensity of heat exchange is inferior to the value of cheapness and low resistance of nozzle devices in comparison with plate-type devices. However, in the case of concentration of the

solution, which is accompanied by the formation of deposits, plate devices have a number of advantages.

Thus, dual-flow trays with a free cross-section of more than 30 % and a hole diameter of 30–120 mm most fully meet the requirements for contact elements for the process of concentration of solutions in direct contact with air.

The mathematical notation of processes in a contact heat exchanger and similar devices, in which interrelated processes of heat and mass exchange occur simultaneously, is traditionally based on the system of equations of the kinetics of these two processes [10]. To solve such a system, data on heat and mass transfer coefficients are required. Mass transfer on dual-flow plates with a free cross-section of 30 % was studied in roots [6, 11]. On the air-water system, it was established that the mass transfer coefficient (K_g) is proportional to the gas velocity (w) to the degree of 0.7–0.95, and the height of the gas-liquid layer on the plate is to the degree of 0.5. With an increase in the free cross-section of dual-flow plates, their specific efficiency increases significantly, which makes them more profitable in terms of energy costs compared to trays with a small free cross-section.

The dependence of mass and heat transfer on operating parameters was most fully investigated in [12], in which equations were built for calculating the corresponding coefficients on trays with large perforations and a free cross-section of more than 30 %. However, it is worth noting that the data on the heat transfer coefficient on dual-flow trays are very limited, and some of them can be found in [12].

But the general problem of the equations found in the literature is their relatively low accuracy, in particular, calculation errors amount to at least 15 %. The low accuracy can be explained, on the one hand, by the instability of the regimes on dual-flow trays, and on the other hand, by methodical problems that arise when determining the mass and heat transfer coefficients in the process under consideration. The peculiarity of evaporative cooling is that heat and mass exchange are closely interdependent and, except in some cases, cannot be studied independently. In practice, researchers believe that mass transfer resistance is concentrated in the gas phase. This approach, on the one hand, has a certain justification in the practice of condensation and evaporation processes, and on the other hand, it is due to the complexity of estimating the heat transfer resistance in a liquid.

A number of methods for calculating contact heat exchangers and cooling towers are based on the joint solution of the equations of "dry" and "wet" heat exchange [6, 13, 14]. This approach opens the way to unification of "dry" and "wet" heat exchange processes, which differ in their physical nature, into a single process of enthalpy transfer between gas and liquid. But the direct use of such equations is complicated by the difficulty of experimentally estimating the temperature and enthalpy at the phase interface. The situation is simplified if we accept the assumption that the main resistance to enthalpy transfer exists on the gas side, while the resistance in the liquid can be neglected. In this case, the enthalpy transfer coefficient becomes equal to the mass transfer coefficient in the gas.

However, there are studies, for example [15], in which it was established that, depending on the process conditions, the resistance to enthalpy transfer from the liquid side is characterized by the value $(1/\alpha_i)[(H^* - H_i)/(t_i - t_i^*)]$ and is 27–46 % of the total resistance, and this proportion decreases with an increase in the density of the irrigation liquid and a decrease in temperature. In work [15], in particular, it was

found that for packed columns the ratio α_l/k_g increases proportionally to $L^{0.43}$. Thus, the resistance to enthalpy transfer from the liquid side is quite large, and the assumption that it can be neglected needs to be verified.

As it follows from the literature [3–15], research aimed at evaluating the kinetic characteristics of heat exchange when combining the processes of “dry” and “wet” heat exchange, which are different in physical nature, into a single process of enthalpy transfer between gas and liquid should be considered justified.

3. The aim and objectives of the study

The purpose of our study was to determine the kinetic of heat and mass exchange between the liquid and the air in contact with it in a column apparatus equipped with dual-flow trays.

To achieve the goal, the following tasks were solved:

- to establish the dependence of the mass transfer coefficient in the gas and the heat transfer coefficient in the liquid on the gas velocity in the apparatus and the irrigation density;
- to establish the dependence of the mass transfer coefficient in the gas and the heat transfer coefficient in the liquid on the geometry of the tray;
- to derive a generalized formula for calculating the enthalpy transfer coefficient.

4. The study materials and methods

4.1. The object and hypothesis of the study

The object of our study was the processes of heat and mass exchange in a contact heat exchanger.

A solution of sodium chloride was taken as the liquid. It was assumed that the hydrodynamic conditions of heat exchange and mass exchange in column apparatuses are factors that determine the kinetics of these processes. Hydrodynamic conditions, in turn, depend mainly on the gas velocity in the apparatus and the density of liquid irrigation of the trays, as well as to a certain extent on the geometry of the trays. This made it possible to consider the problem from the point of view of determining the essential kinetic characteristics of evaporation and the limiting stages of the process. It also makes it possible to obtain data on the ratio of heat and mass transfer in gas and heat transfer in liquid under different operating modes and designs of contact heat exchanger trays.

4.2. Structure of the laboratory bench

The experimental part of the work was aimed at identifying the degree of influence of the listed factors on the kinetic coefficients of heat and mass transfer in the process of concentrating the sodium chloride solution in the contact heat exchanger. The study of heat and mass transfer characteristics was carried out on a traditional installation for such work, consisting, in fact, of the investigated apparatus and a set of auxiliary equipment that enables its operation. This equipment included a solution tank, a circulation pump, and a heat exchanger for heating the solution. Air for the work of the bench was pumped with a fan.

The experimental contact heat exchanger had a base to which air supply and solution removal pipelines were con-

nected. Three cylindrical rings with a diameter and height of 0.5 m each were installed on the base one above the other. The trays were installed in the flange connectors of the rings. A cover equipped with a pipeline for air removal and a device for distributing the solution entering the apparatus was attached to the upper tank.

During the experiments, the following parameters were measured at the inlet and outlet of the installation: air and liquid flows, air humidity, liquid and air temperature, and pressure. Samples of the solution were also taken for analysis.

The choice of tray sizes was based on the results of a literature review and the experience of operating contact heat exchangers. A total of 5 types of trays were tested, which differed in the diameter of the holes and the free section. The diameter of the holes in the trays varied on 3 levels and was 50, 65, and 100 mm. The free cross-section also varied at three levels and was $30 \pm 2\%$, 42% , and $53 \pm 2\%$.

4.3. Calculation of kinetic characteristics of heat exchange

The study of the processes taking place in devices equipped with dual-flow trays faces certain difficulties. In practice, it is extremely difficult to accurately measure the values of concentration and temperature separately for liquid and gas inside a working apparatus. Fluid splashes interfere with gas temperature measurements, and air humidity changes even during sampling. Another problem that affects the accuracy of measurements is the averaging of temperatures and compositions of gas and liquid over the tray area.

To overcome these difficulties, a methodology and a computer program were developed to determine the heat and mass transfer coefficients based only on experimental data on the flow rate, composition, and temperature of the contacting streams at the inlet and outlet of the laboratory apparatus. This program contains a mathematical model of a laboratory installation and an algorithm for selecting coefficients in which the results of the experiment and calculation match. The value of the enthalpy transfer and mass transfer coefficients, at which this coincidence occurs, is the result of the calculation.

4.4. Method of separate determination of mass transfer resistance in gas and liquid

In the case of a joint solution of the “dry” and “wet” heat transfer equations, the total rate of heat transfer from liquid to gas, taking into account the heat of vaporization, is:

$$L c_l dt_l = [\alpha_g (t - t_i) + r k_g (x - x_i)] dS, \quad (1)$$

where L is mass flow of liquid, kg/s; r – heat of vaporization, kJ/kg; c_l – heat capacity of liquid, kJ/kg-degree; t_i , t , t_l are the temperatures of the gas-liquid interface, the main volume of gas and liquid, respectively, °C; S – phase contact surface, m²; x_i , x – humidity of the gas in equilibrium with the liquid at the phase interface and in the gas volume, respectively, kg/kg; α_g – heat transfer coefficient in the gas phase, W/m²-degree; k_g – mass transfer coefficient in gas, kg/m²s.

Converting equation (1) on the basis of some assumptions, in particular, that the ratio $\alpha_g/k_g c_g$ (psychrometric coefficient) is equal to unity, it is shown that the rate of heat exchange, which is transferred by steam between the surface of the liquid and the main volume of the gas, is determined by the following equation:

$$G = \frac{dH_g}{dS} = k_g (H_g - H_i), \quad (2)$$

where G is the mass flow of gas, kg/s; H_g , H_i – specific enthalpies of the main volume of gas and gas at the interface, kJ/kg.

The equation that determines the enthalpy transfer coefficient in this case is:

$$\frac{1}{k_H} = \frac{1}{k_g} + \frac{H^* - H_i}{\alpha_l (t_i - t_l)}, \quad (3)$$

where H^* is the specific enthalpy of the gas in equilibrium with the mass of the solution, kJ/kg; α_l is the coefficient of heat transfer in the liquid, kJ/kg·degree.

To identify the mechanism of heat exchange, its limiting stages, and directions of intensification, it is necessary to be able to calculate the components of heat transfer resistance. From the point of view of the theory of double-film heat and mass transfer, it is, firstly, the resistance to heat transfer from the main volume of the liquid through the laminar film to the phase interface. Secondly, it is the transfer of the mass of water in the form of steam and physical heat from the interface of the phases through the laminar film into the main gas volume. It can be seen from equation (3) that the first of the resistances of interest is determined by the expression $(1/\alpha_l)[(H^* - H_i)/(t_l - t_i)]$, and the second by the ratio $1/k_g$.

When devising the research methodology, it was assumed that the mass transfer coefficient in the gas phase k_g and the heat transfer coefficient in the liquid phase α_l do not depend on temperature. Under such conditions, equation (3) turns into a linear dependence of the value $1/k_H$ on the ratio $(H^* - H_i)/(t_l - t_i)$ with coefficients $a_0 = 1/k_g$ and $a_1 = 1/\alpha_l$. Accepting these assumptions and conducting several experiments with the same air and solution flows, but at different temperatures, it is possible to calculate these coefficients, and if so, then separately the values of k_g and α_l .

Implementation of the above methodology required some additional calculations. It was a certain difficulty to determine the values of H_i and t_i because their direct measurement is impossible and can be estimated very approximately. A good approximation can be obtained by considering that the difference $t_l - t_i$ has a value of several degrees or even a fraction of a degree, and therefore it can be assumed with sufficient accuracy that:

$$\frac{H^* - H_i}{t_l - t_i} \approx \left(\frac{dH}{dt} \right)_{t=t_i}. \quad (4)$$

Another difficulty is the lack of literature data on the enthalpy of air H^* in equilibrium with sodium chloride solution. To determine this value per 1 kg of its dry fraction in equilibrium with NaCl solution, the well-known formula is used:

$$H^* = c_{ca} t + x H_v^*, \quad (5)$$

where c_{ca} is the specific heat capacity of dry air (considered constant and equal to 1.01 kJ/kg·K); t – solution temperature, °C; H_v^* is the specific enthalpy of water vapor in equilibrium with the solution, kJ/kg.

The boiling point of the solution is higher than the boiling point of pure water. The enthalpy of such a pair was calculated using the formula:

$$H_v^* = H_{pv} + c_v \Delta t, \quad (6)$$

where H_{pv} is the specific enthalpy of saturated water vapor at a pressure equal to the partial pressure of the vapor above the solution, kJ/kg; c_v is the specific heat capacity of water vapor (in the studied temperature range of 10...80 °C it is taken as constant: $c_v = 1.76$ kJ/kg·K); Δt is the difference between the boiling point of the solution and the solvent, °C.

To calculate the value of Δt , we used known data on the dependence of the normal boiling temperature of the solution on the concentration of sodium chloride in it and the pressure of saturated water vapor. As a result, combining these data with Babot's rule for a concentration of NaCl in a solution of 25–26 % (it was this solution that was used in most of the experiments) resulted in the following equation for the dependence of Δt on the temperature of the solution:

$$\Delta t = (273,15 + t) \left[1 - \frac{1}{1 + 5,353 \cdot 10^{-5} (273,15 + t)} \right]. \quad (7)$$

The system of equations (5) to (7) was used to calculate the enthalpy of air in equilibrium with a NaCl solution close to saturation, and the latter to obtain point estimates of the dH/dt derivative. Statistical treatment of the results of these calculations made it possible to derive the following equation

$$dH/dt = 0.7 + 0.0144t + 0.00036t^2. \quad (8)$$

The resulting dependence allows the use of calculated indicators during the mathematical processing of results.

4. 5. Processing of experiment results

The primary results of the experiments were processed in three stages. At the first stage (pre-processing), a number of auxiliary values and factor values necessary for calculating the heat and mass transfer coefficients were calculated based on the data tables obtained in the experiments. The second stage involved finding the coefficients of mass transfer and heat transfer. At the third stage, the type of equations connecting these kinetic characteristics with the factors affecting them was selected.

To build mathematical models describing the dependence of kinetic coefficients on the input variables of the process, methods of experiment planning and regression analysis were used.

When studying the influence of the gas velocity in the apparatus and the irrigation density on the mass transfer coefficient in the gas and the heat transfer coefficient in the liquid, the following ranges of variation were chosen:

- air speed (w): (0.25–3.5) m/s,
- irrigation density (I): (0.4–5) kg/m²s.

The initial variables (heat and mass transfer coefficients) are taken as:

- mass transfer coefficient in gas (k_g , kg/m²s),
- coefficient of heat transfer in the liquid (α , W/m² degrees).

In the experiments of this series, trays with holes with a diameter of 65 mm and a free cross-section of 0.42 were installed in the laboratory apparatus.

To estimate the coefficients of the regression equations, the plan of the full factorial experiment 2² was used. To estimate the variance of the experimental errors, parallel experiments were conducted in the center of the plan in the number of $n=5$.

When studying the influence of the fraction of the free section of the tray (ψ) and the diameter of the holes (d) in it on the kinetic coefficients of heat and mass transfer, experiments were conducted with five standard sizes of trays, which were characterized by different geometric characteristics. The first group of sizes: $d=0.05$ m, $\psi=0.55$ and 0.31 m²/m². The second group: $d=0.1$ m, $\psi = 0.28$ and 0.52 m²/m². The last standard size $d=0.065$ m, $\psi=0.42$ m²/m². In all experiments, the air velocity was set at 0.7 m/s in the full cross section of the device, and the irrigation density was 1.5 kg/s·m². The temperature of the initial solution was changed in each experiment.

The input variables were normalized according to the formula:

$$x_i = \frac{2x_m - (x_{i_{\max}} + x_{i_{\min}})}{x_{i_{\max}} - x_{i_{\min}}}, \quad i = 1, 2, \dots, N, \quad (9)$$

where x_i is the value of the i th input variable in normalized form, x_{in} is the value of the i th input variable in natural form, $x_{i_{\min}}$ is the lower limit of the range of variation of the i^{th} input variable, $x_{i_{\max}}$ is the upper limit of the range of variation of the i th input variable, N – the number of experiments according to the plan of a full factorial experiment ($N=4$).

Using formula (9), the absolute values of the input variables were translated into the dimensionless range $[-1; +1]$.

The following notations are used for the input variables when calculating the coefficients of the regression equations:

- $w \rightarrow x_1$;
- $l \rightarrow x_2$;
- $\psi \rightarrow x_3$;
- $d \rightarrow x_4$.

The general structure of mathematical models is as follows:

$$k = b_0 \prod_{i=1}^k x_i^{b_i}, \quad (10)$$

where k is the kinetic coefficient (in this case, the mass transfer coefficient in gas, k_g , kg/m²s or α , W/m²degree); b_i is the coefficient to be determined, k is the number of input variables (for each series of experiments $k=2$).

Construction of regression equations was preceded by linearization (10) by logarithmization:

$$\ln k = \ln b_0 + \sum_{i=1}^k b_i \ln x_i, \quad (11)$$

Estimation of the variance of model inadequacy was carried out using the formula:

$$s_D^2 = \frac{S_R}{\Phi_1}, \quad (12)$$

where S_R is the sum of the squares of deviations of the calculated values of the kinetic coefficients ($\ln k=y$) according to the regression equation (y_{je}) from the experimental values (y_{je}):

$$S_R = \sum_{j=1}^N (y_{je} - y_{jc})^2, \quad j = 1, N. \quad (13)$$

Φ_1 – number of degrees of freedom:

$$\Phi_1 = nk - n. \quad (14)$$

The estimation of the variance of the errors of the experiments was determined by the formula:

$$s^2 = \frac{S_E}{\Phi_2}, \quad (15)$$

where:

$$S_E = \sum_{j=1}^N \left(y_{je} - \frac{1}{N} \sum_{j=1}^N y_{je} \right)^2, \quad (16)$$

Φ_2 – number of degrees of freedom:

$$\Phi_2 = n - 1. \quad (17)$$

The significance of the coefficients was evaluated using the Student's test, the adequacy test was carried out by testing the hypothesis about the equality of the estimates of the variance of the inadequacy and experimental errors.

5. Results of determination of kinetic patterns of heat and mass transfer

5.1. Determining the dependence of kinetic coefficients on the gas velocity in the apparatus and the irrigation density

After linearization (11), the input variables took the following form:

$$X_1 = \ln w; X_2 = \ln l. \quad (18)$$

Table 1 gives the results of the experiments.

Table 1

Plan and results of experiments to determine the kinetic coefficients k_T and α

Experiment No.	Experiment plan in natural and normalized form				Kinetic coefficients	
	X ₁		X ₂		k_g	α
1	1.252763	+1	1.609438	+1	11.09	84.1
2	1.252763	+1	-0.91629	-1	1.30	23.1
3	-1.38629	-1	1.609438	+1	1.03	17.3
4	-1.38629	-1	-0.91629	-1	0.12	6.0
5	-0.06677	0	0.346574	0	1.21	21.3
6	-0.06677	0	0.346574	0	1.14	22.2
7	-0.06677	0	0.346574	0	1.17	22.3
8	-0.06677	0	0.346574	0	1.12	21.7
9	-0.06677	0	0.346574	0	1.17	22.9

The following regression equations of type (11) in normalized form are obtained:

$$\ln k_g = 1.1584 + 1.19X_1 + 1.073X_2, \quad (19)$$

$$\ln \alpha = 21.672 + 0.732X_1 + 0.588X_2. \quad (20)$$

The corresponding equations of the form (10) in their natural form are:

$$k_g = 0.916w^{0.902}l^{0.85}, \quad (21)$$

$$\alpha = 19.138w^{0.555}l^{0.466}. \quad (22)$$

Table 2 gives the results of statistical data processing to verify the adequacy of models (19), (20).

Table 2

Results of static data processing

Kinetic coefficients	Estimated data of variance estimates					
	S_R	ϕ_1	s_r^2	S_E	ϕ_2	s_e^2
k_g	3.67E-05	1	3.67E-05	0.00345	4	0.000863
α	0.0156	1	0.0156	0.00305	4	0.000763

Fig. 1, 2 show the response surfaces described by equations (21), (22).

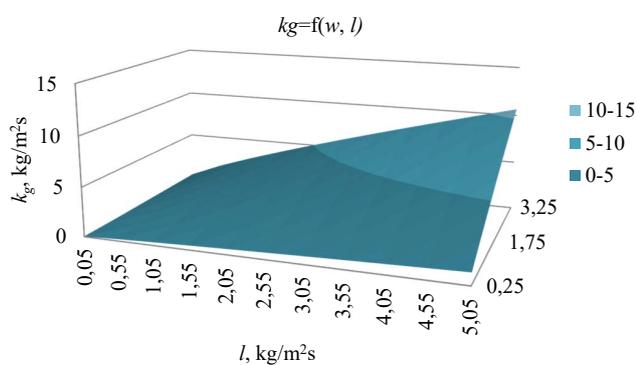


Fig. 1. Response surface $k_g = f(w, l)$

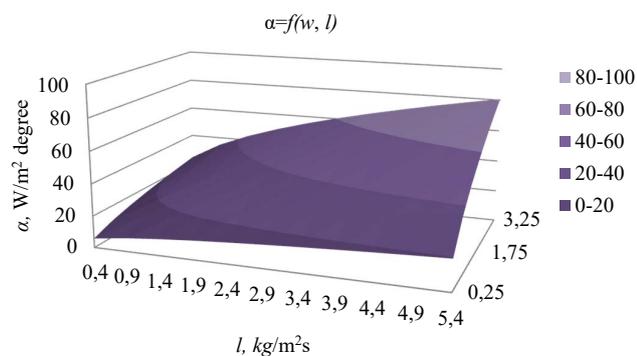


Fig. 2. Response surface $\alpha = f(w, l)$

Fig. 3, 4 show the projections of the intersection of the response surfaces (21), (22) by the planes $k_g = \text{const}$ and $\alpha = \text{const}$, respectively, at the levels corresponding to their average values according to experimental data (Table 1), i. e., $k_g = 1.162 \text{ kg/m}^2\text{s}$, $\alpha = 22.08 \text{ W/m}^2 \text{ degree}$.

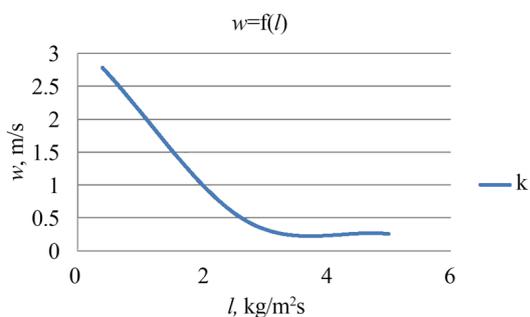


Fig. 3. Projection of the intersection of the response surface $k_g = f(w, l)$ by the plane $k_g = 1.162 \text{ kg/m}^2\text{s}$

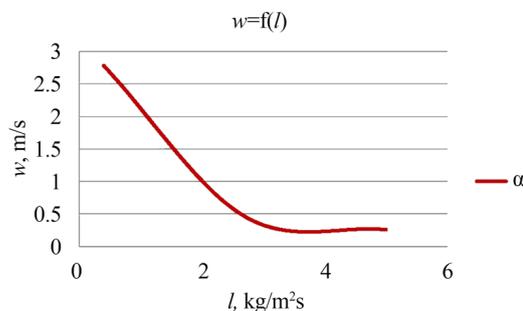


Fig. 4. Projection of the intersection of the response surface $\alpha = f(w, l)$ with the plane $\alpha = 22.08 \text{ W/m}^2 \text{ degree}$

Fig. 5 shows the combined projections of the intersection of the response surfaces (21), (22) and the directions of the selection of the sets w and l that satisfy, for example, the conditions $k_g > 1.162 \text{ kg/m}^2\text{s}$ and $\alpha < 22.08 \text{ W/m}^2 \text{ degree}$ (shown by arrows of the corresponding color).

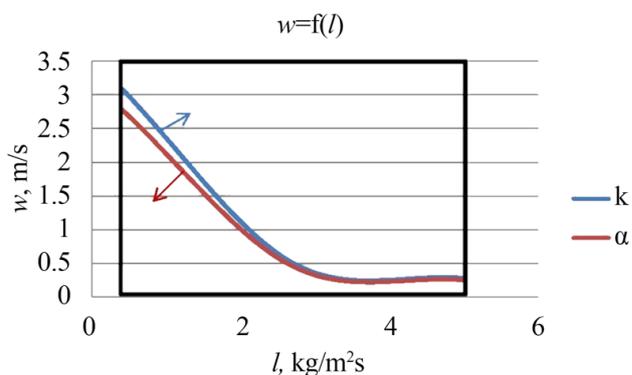


Fig. 5. Combined projections of the intersection of the response surfaces (21), (22) and the directions of the selection of the sets w and l (on the example, the conditions $k_g > 1.162 \text{ kg/m}^2\text{s}$ and $\alpha < 22.08 \text{ W/m}^2 \text{ degree}$ are fulfilled)

Fig. 1, 2 make it possible to determine the nature of the response surfaces and the direction of movement towards the optimum by controlled change of the values of the input variables by methods of experimental optimization, and Fig. 3, 4 make it possible to choose the ranges of values of the input variables that provide the specified requirements for the values of the kinetic coefficients.

5. 2. Determining the dependence of kinetic coefficients on the geometry of the tray

After linearization (11), the input variables took the following form:

$$X_3 = \ln \psi ; X_4 = \ln d.$$

Table 2 gives the results of experiments on determining the kinetic coefficients k_T and α for different values of the fraction of the free cross section of the tray and the diameter of the holes in the tray.

The calculation of coefficients in the regression equations and the statistical analysis of their significance showed that the diameter of the holes in the tray does not exert a statistically significant effect on any of the output variables. Therefore, the equations $k_g = f(\psi, d)$ $\alpha = f(\psi, d)$ take a one-dimensional form:

$$k_g = 0.298\psi^{-1.67}, \tag{23}$$

$$\alpha = 16,038\psi^{-0.396}. \tag{24}$$

Table 2
Plan and results of experiments to determine the kinetic coefficients k_g and α

No. of entry	ψ , m ² /m ²	d , m	w , m/s	l , kg/m ^{2s}	k_g	α
1	0.52	0.1	0.7	1.5	0.86	20.2
2	0.55	0.05	0.7	1.5	0.83	20.9
3	0.28	0.1	0.7	1.5	2.46	26.9
4	0.31	0.05	0.7	1.5	2.15	25.2
5	0.42	0.065	0.7	1.5	1.212	20.3
6	0.42	0.065	0.7	1.5	1.141	23.8
7	0.42	0.065	0.7	1.5	1.174	22.3
8	0.42	0.065	0.7	1.5	1.166	21.7
9	0.42	0.065	0.7	1.5	1.121	22.9

The corresponding graphical dependences and coefficients of determination are shown in Fig. 6, 7.

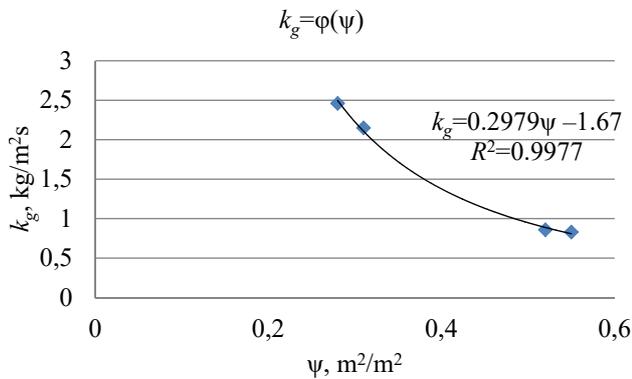


Fig. 6. Dependence $k_g = \phi(\psi)$

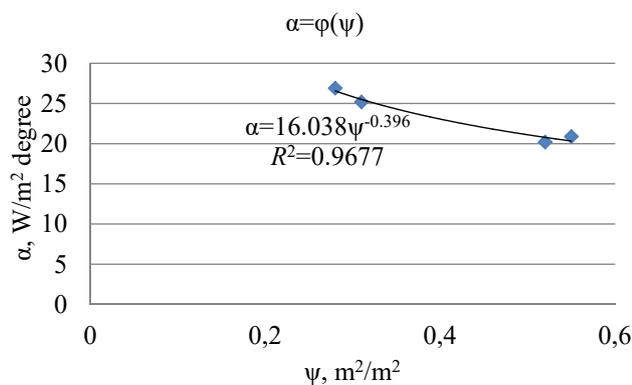


Fig. 7. Dependence $\alpha = \phi(\psi)$

From Fig. 6, 7 it follows that the chosen structure of the equation is correct, and equations (3), (24) adequately reflect the law of change of kinetic coefficients.

5. 3. Construction of a generalized formula for calculating the enthalpy transfer coefficient

The resulting kinetic equations make it possible to construct a generalized formula for calculating the enthalpy transfer coefficient k_H in the process of concentrating a so-

dium chloride solution in a contact heat exchanger equipped with dual-flow trays with large diameter holes. For this, formula (3) was used, solved explicitly with respect to k_H :

$$k_H = \frac{1}{\frac{1}{k_g} + \frac{1}{\alpha} \cdot \frac{dH}{dt}}. \tag{25}$$

Taking into account (21) and (22) and the accuracy of estimating the coefficients of the equations, the following equations are derived for determining the coefficients of mass transfer in a gas and heat transfer in a liquid:

$$k_g = 0.202w^{0.8}l^{0.86}\psi^{-1.8}, \tag{26}$$

$$\alpha = 13.2w^{0.48}l^{0.47}\psi^{-0.47}. \tag{27}$$

The resulting equations (8), (26), and (27) allow us to use the generalized formula (25) to calculate the enthalpy transfer coefficient k_H in the process of concentrating the sodium chloride solution in the contact heat exchanger.

6. Discussion of results of determining the kinetic patterns of heat and mass transfer in a column heat exchanger with direct phase contact

The results given in equations (21), (22) and shown in Fig. 1, 2, indicate that to increase the intensity of mass transfer in the gas and heat transfer in the liquid, it is necessary to increase the value of the gas velocity and the irrigation density. This means that in order to find the optimal area when moving along the response surface, it is worth choosing steps in such a direction that enables an increase in the values of the input variables. Experimental optimization methods can be used for this. In addition, as can be seen from (21), (22), gas velocity and irrigation density have approximately the same effect on the intensity of mass transfer in the gas and heat transfer in the liquid, which at first glance seems paradoxical. It is usually believed that the mass transfer in a gas is determined by the gas velocity. Irrigation density makes a minor contribution to the change of this indicator. For a cross-current dish, this appears to be true. But when using countercurrent trays, where liquid and gas pass through the same holes, an increase in the density of irrigation leads to a decrease in the cross-section of gas passage. It also leads to an increase in its speed in the holes and a corresponding increase in mass transfer in the gas.

When studying the combined effect on both output variables, it is worth considering that they are not in conflict with each other. This is evidenced by the signs of the coefficients in equations (21) and (22). Therefore, to enable the specified levels of intensity of mass transfer in the gas and heat transfer in the liquid, the selection of gas velocity and irrigation density values should be carried out in the same directions. This can be graphically represented as in Fig. 5, but with arrows pointing up from the corresponding projection, which corresponds to the condition $k_g > 1.162$ kg/m²s and $\alpha > 22.08$ W/m²degree.

As follows from equations (23), (24), and Fig. 6, 7, the value of the fraction of the free section of the tray has a decisive influence on the value of the considered kinetic coefficients. This is evidenced by the high value of the coefficients of determination for the dependences $k_g = \phi(\psi)$ and $\alpha = \phi(\psi)$: $R^2 = 0.9977$ and $R^2 = 0.9677$, respectively. This, however, could

be expected, since the fraction of the free cross-section of the plate determines the gas velocity in the holes and, therefore, the intensity of transport processes in the gas phase.

It should be noted that formula (25) makes it possible to identify the limiting stage of the water evaporation process. Thus, the $1/k_g$ term of this equation represents the heat transfer resistance in the form of enthalpy of water vaporization, and $(1/\alpha)(dH/dt)$ is the heat transfer resistance from the liquid volume to the interface. Comparing the values of these resistances gives a quantitative idea of the directions of process intensification. It is also important to note that the equations for the kinetic coefficients were obtained taking into account the constructed generalized formula for calculating the enthalpy transfer coefficient k_H in the process of concentrating the sodium chloride solution. At the same time, it was taken into account that the process takes place in a contact heat exchanger equipped with dual-flow trays with large diameter holes. As a result, these equations make it possible to obtain qualitative estimates based on the analysis of equations (26), (27). In particular, thanks to this, it was determined that an increase in gas velocity to the greatest extent reduces the resistance to enthalpy transfer in both liquid and gas. But the power of this influence on heat exchange is much less than on mass exchange. In turn, increasing the temperature of the solution reduces resistance, mainly heat transfer. This information could be used in technical and economic calculations to optimize the design and operation of the contact heat exchanger.

The practical aspect of our study is that the work was carried out as part of the design of a unit for concentrating a hot solution of sodium chloride by evaporating water while blowing air in a contact heat exchanger.

The results of the study are limited by the specified ranges of variation in the input variables, within which the obtained equations are adequate. Exceeding the limits of these ranges could lead to obtaining other values of the intensity of mass transfer in the gas and heat transfer in the liquid. This allows us to talk about the promising area of future research, which may involve conducting further experiments in accordance with the plans determined by the directions and steps of movement along the response surface in the direction of the gradients. But this is associated with certain costs directly related to the need to conduct an active experiment.

7. Conclusions

1. Adequate regression equations describing the effect of gas velocity and irrigation density on the kinetic co-

efficients – gas mass transfer coefficient and liquid heat transfer coefficient – were derived. It was determined that to increase the intensity of mass transfer in the gas and heat transfer in the liquid, it is necessary to increase the gas velocity and irrigation density. It was established that gas velocity and irrigation density have approximately the same effect on the intensity of mass transfer in the gas and heat transfer in the liquid.

2. When studying the effect of the geometry of the tray on the kinetic coefficients of heat transfer in the liquid and mass transfer in the gas, it was established that the value of the fraction of the free cross section of the tray has a decisive influence on the value of the considered kinetic coefficients. The resulting high values of the coefficients of determination ($R^2=0.9977$ for the mass transfer coefficient in the gas and $R^2=0.9677$ for the heat transfer coefficient in the liquid) testify to the adequacy of our equations, which take the form of power-law dependences.

3. A generalized equation was built, which makes it possible not only to calculate the enthalpy transfer coefficient during the interaction of a sodium chloride solution with air but also to determine the limiting stage of this process. This equation could be used as part of the mathematical support for the operation of the plant for obtaining crystalline sodium chloride using secondary energy resources and other non-traditional sources of thermal energy.

Conflicts of interest

The authors declare that they have no conflicts of interest in relation to the current study, including financial, personal, authorship, or any other, that could affect the study and the results reported in this paper.

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Data availability

All data are available in the main text of the manuscript.

Use of artificial intelligence

The authors confirm that they did not use artificial intelligence technologies when creating the current work.

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