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Optimization of the process of heating an oil and gas condensate mixture by light naphtha vapor in heat – exchanger condenser 10E04

Optymalizacja procesu podgrzewania mieszaniny kondensatu ropy naftowej i gazu ziemnego parami lekkiej benzyny w wymienniku ciepła skraplacza 10E04

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ABSTRACT: The constant growth in the cost of energy carriers leads to finding the ways to reduce production costs and efficient use of energy resources by optimizing the technological regimes of oil refineries. Based on the current state of the resource base and the country's growing needs for oil and gas resources, the introduction of modern technologies for maintaining and intensifying the production and processing of oil and gas, the priority is to increase the energy efficiency of the fuel and energy complex. In this aspect, scientific research aimed at the efficient use of the heat of process flows in the stage of thermal preparation of local hydrocarbon raw materials for distillation, the development of new, energy and resource-saving technologies for the production of petroleum products, the optimization of the technological regimes of oil refineries in order to increase their thermal efficiency and reduce the cost of manufacture of petroleum products, the synthesis of technological schemes with a rational number of apparatus for the thermal preparation of raw materials for distillation is of great importance. Based on this, a mathematical model has been developed for the process of thermal preparation of an oil and gas condensate mixture for distillation, which makes it possible to determine the energetically optimal heat exchange surface of the apparatus. As a system of equations, the objective function of the optimality criterion is formulated – the specific technological prime cost of the heated oil and gas condensate mixture, which includes changes in the power consumption for its pumping and depreciation deductions for heat exchange equipment depending on parameters the process. On the grounds of investigation of the influence of technological parameters – temperature, composition, and consumption of raw materials in the pipe, as well as steam pressure in the annulus on the efficiency of heat transfer in an industrial heat exchanger, the optimal operating conditions for the heat exchanger 10E-04 at the Bukhara Oil Refinery have been identified: F_{inter} = 219 m^2, Cp = 0.010825 USD/kg, A = 0.010825 USD/kg, C_{st} = 0.02165 USD/kg, and t = 107.3°C, with a specified throughput of G_{rm} = 105 508.3 kg/h.

Key words: oil and gas condensate mixture, atmospheric distillation of the mixture, fuel fraction, naphtha, heating, heat exchanger, condenser, modeling, technological cost, optimization, energy efficiency, refinery installation.

STRESZCZENIE: Stały wzrost kosztów nośników energii prowadzi do konieczności znalezienia sposobów na obniżenie kosztów produkcji i efektywnego wykorzystania zasobów energetycznych poprzez optymalizację reżimów technologicznych rafinerii ropy naftowej. W oparciu o obecny stan bazy surowcowej i rosnące zapotrzebowanie kraju na zasoby ropy naftowej i gazu ziemnego, a także wprowadzenie nowoczesnych technologii utrzymania i intensyfikacji produkcji i przetwarzania ropy naftowej i gazu, priorytetem jest zwiększenie efektywności energetycznej kompleksu paliwowo-energetycznego. W tym aspekcie ogromne znaczenie mają badania naukowe mające na celu efektywne wykorzystanie ciepła przepływów procesowych na etapie termicznego przygotowania lokalnych surowców węglowodorowych do destylacji, opracowanie nowych, energooszczędnych i zasobooszczędnych technologii produkcji produktów naftowych, optymalizację reżimów technologicznych rafinerii ropy naftowej w celu zwiększenia ich wydajności cieplnej i obniżenia kosztów produkcji produktów naftowych, a także syntezę schematów technologicznych z racjonalną liczbą aparatów do termicznego przygotowania surowców do destylacji. Na tej podstawie opracowano model matematyczny procesu termicznego przygotowania mieszaniny kondensatu ropy naftowej i gazu do destylacji, który umożliwia określenie optymalnej energetycznie powierzchni wymiany ciepła aparatu. Jako układ równań sformułowano funkcję celu kryterium optymalności – specyficzny technologiczny koszt główny podgrzanej mieszaniny kondensatu ropy naftowej i gazu ziemnego, który obejmuje zmiany zużycia energii do jej pompowania i odpisy amortyzacyjne na urządzenia do wymiany ciepła

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w zależności od parametrów procesu. Na podstawie badań wpływu parametrów technologicznych – temperatury, składu i zużycia surowców w rurze, a także ciśnienia pary w pierścieniu na efektywność wymiany ciepła w przemysłowym wymienniku ciepła, zidentyfikowano optymalne warunki pracy wymiennika ciepła 10E-04 w rafinerii ropy naftowej w Bucharze: F_{inter} = 219 m^2, Cp = 0.010825 USD/kg, A = 0,010825 USD/kg, C_{st} = 0,02165 USD/kg i t = 107,3°C, przy określonej przepustowości G_{rm} = 105 508,3 kg/h.

Słowa kluczowe: mieszanina kondensatu ropy naftowej i gazu ziemnego, destylacja atmosferyczna mieszaniny, frakcja paliwowa, benzyna ciężka, podgrzewanie, wymiennik ciepła, skraplacz, modelowanie, koszt technologiczny, optymalizacja, efektywność energetyczna, instalacja rafineryjna.

Introduction

Heat exchangers are systems utilized for transferring heat between fluids with different temperatures. These devices find widespread applications in various sectors, including refrigeration, heating, air conditioning systems, power plants, chemical processes, the food industry, automotive radiators, and waste heat recovery units. Heat exchangers can be categorized based on various criteria such as construction, flow arrangement, heat transfer mechanism, and so on (Lahiri et al., 2012; Kharaji, 2021).

According to the existing production technology at oil refineries (OR), raw materials are initially heated in three blocks consisting of heat exchangers, then they are heated in a coil furnace, and then the raw materials are physically separated into fractions by distillation (Skoblo et al, 2000; Manovyan, 2001; Akhmetov et al., 2006; Glagoleva and Kapustina, 2006; Akhmetov, 2013). To heat the feedstock in the refinery, hot process streams leaving the distillation column are used – distillates of fuel fractions and fuel oil (Skoblo et al, 2000; Manovyan, 2001; Glagoleva and Kapustina, 2006). As it is known, tubular heat-exchange apparatus is mainly used to heat hydrocarbon feedstock at oil refineries (Skoblo et al., 2000; Akhmetov et al., 2006; Akhmetov, 2013).

The heat exchanger design can be divided into two main categories, thermal and hydraulic design and mechanical design. In thermal and hydraulic design, the focus is on calculating an adequate surface area to transfer a certain amount of heat, pressure dope, pumping power work, etc. (Lahiri et al, 2012).

An analysis of the market for heat exchangers in Russia and the CIS countries (Report of the Academy on Industrial..., 2008; Khudaiberdiev and Khudaiberdiev, 2010; Heat exchange equipment market, 2022) showed that oil refining, petrochemical, and chemical industries use an average of 406 to 1,250 heat exchangers. Almost 90% of the fleet of heat exchangers used by factories are tubular apparatus, the total mass of which is 35–40% of the mass of oil refinery (Skoblo et al, 2000; Ismaylov et al., 2024). However, along with high reliability, these devices are characterized by large weight and size parameters (diameter: 0.63–1.8 m, tube length: 5–10.6 m, weight – up to 35–40 t) and low thermal efficiency (heat transfer coefficient: 50–100 W/m$^2 \cdot$ K) (Akhmetov et al., 2006; Khudaiberdiev, 2019a). In this regard, the main ways to reduce the consumption of thermal energy are the intensification of heat and mass transfer (HMT) processes, the reduction of heat loss to the environment by radiation, cooling water and air, as well as the maximum use of heat from low-potential flows (Babkin et al., 2014; Khudaiberdiev, 2019a; Khurmamatov et al., 2023a).

It was found that the use of the heat of waste oil fractions in tubular heat exchangers makes it possible to intensify the heating of crude oil feedstock from 10.4 to 45.9%, depending on their phase state and process parameters. In addition, the use of hydrocarbon vapors during oil refining contributes to an increase in the technological efficiency of an oil refinery from 2.5–5% (Khudaiberdiev, 2019a).

Consequently, carrying out further research aimed at identifying the optimal limits of the technological regime of an oil refinery installation, developing new energy-saving methods for heating oil feedstock, effectively organizing heat-and-mass transfer (HMT) processes in apparatuses, as well as clarifying the methods for calculating the main processes and oil refining apparatuses acquire important scientific and practical significance. In this aspect, the use of modern methods for calculating heat exchange in refinery apparatuses, based on the principles of system analysis and mathematical modeling of processes, contributes to the development of efficient and compact designs of apparatuses that provide significant savings in energy, metal, and operating costs (Khudaiberdiev, 2019a).

Materials and methods

In most heat transfer problems, hot and cold fluids are divided by a solid wall. In this case, the mechanism of heat transfer from the hot fluid to the cold fluid can be categorized into three steps (Ezgi, 2017):

- heat transfer from the hot fluid to the wall by convection;
- heat transfer through the wall by conduction;
- heat transfer from the wall to the cold fluid by convection.

In this paper, a method for optimal calculation and design of the process of heating local hydrocarbon feedstock (oil, gas condensate, and their mixtures) with light fraction vapors in tubular heat exchangers of refineries is proposed, taking into

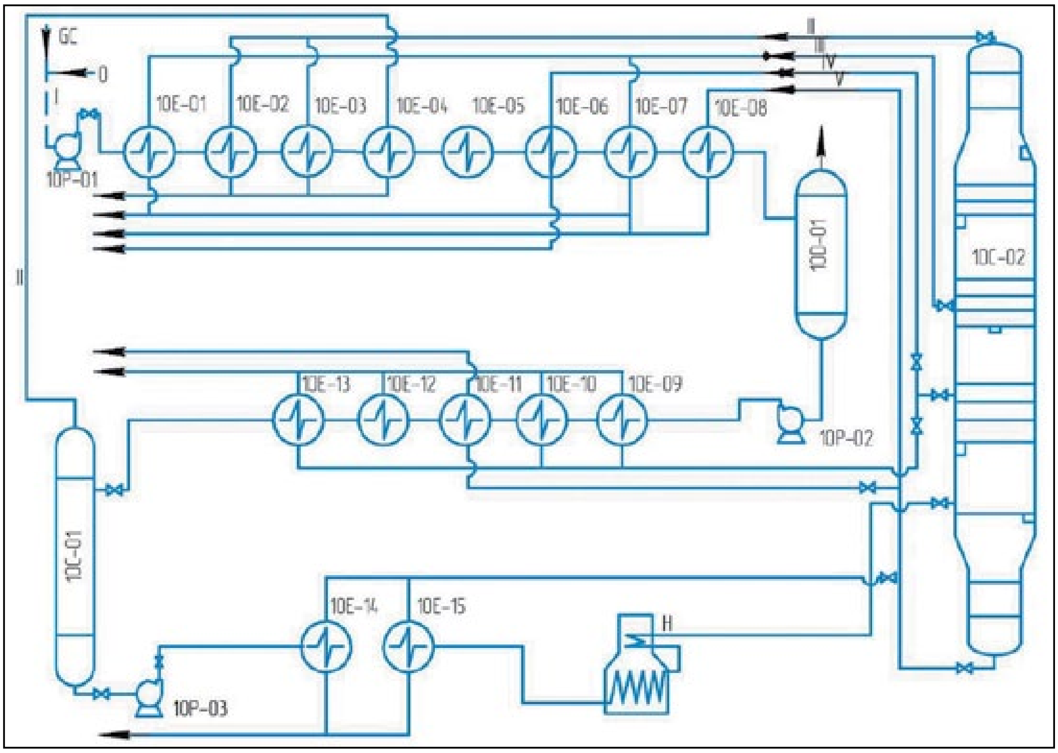

Figure 1. Technological scheme of thermal preparation of oil and gas condensate mixture in the atmospheric distillation unit of the Bukhara Oil Refinery; N – oil, GC – gas condensate, D01 – condensate evaporator, C01 – pre-fractionation column, C02 – atmospheric distillation column, P01, 02, 03 – raw material pumps, H – coil heater, I – raw material line, II – combined naphtha line, III – kerosene line, IV – kerosene line, V – gas-oil line

Rysunek 1. Schemat technologiczny termicznego przygotowania mieszaniny ropy naftowej i kondensatu gazowego w instalacji destylacji atmosferycznej rafinerii ropy naftowej w Bucharze; N – ropa naftowa, GC – kondensat gazowy, D01 – parownik kondensatu, C01 – kolumna wstępnej frakcjonacji, CO2 – kolumna destylacji atmosferycznej, P01, 02, 03 – pompy surowca, H – podgrzewacz wężownicy, I – linia surowcowa, II – linia benzyny ciężkiej, III – linia naftowa, IV – linia naftowa, V – linia gazu i ropy

account temperature changes in the properties of raw materials and hydrocarbon coolant. Such refinement of the process parameters contributes to the development of energy-optimal designs of heat exchangers. There is also a significant reduction in the technological cost of heated feedstock due to a decrease in the consumption of heat and electricity for the process, as well as the synthesis of rational schemes for a block of heat exchangers for heating the raw materials of an oil refinery (Khudaiberdiev, 2019a; Khudaiberdiev and Rakhimzhanova, 2022; Khurmamatov and Auesbaev, 2023).

The thermal preparation of oil and gas condensate mixtures at the refinery is carried out in three stages (Figure 1), in sequentially connected blocks of heat exchangers of the atmospheric distillation unit. Hot process streams (fuel fraction distillates, circulating refluxes, bottom residue) are used as heat carriers, supplied to the tube or inter-tube space of the apparatus.

The first stage of preheating the feedstock consists of a system of eight sequentially connected heat exchangers. Dehydrated and desalinated oil from the ELOU block tank, as well as gas condensate from the tanks, are fed through pumps and a mixer to the raw material pump intake and then pumped through the heat exchangers: 10E-01 (heat carrier – kerosene in the inter-tube space), 10E-02 (heat carrier – naphtha vapors from columns S01 and S02 in the tube space), 10E-03 (heat carrier – top column vapors from column S02 in the tube space), 10E-04 (heat carrier – top column vapors from column S01 in the tube space), 10E-05 (heat carrier – kerosene in the tube space), 10E-06 (heat carrier – kerosene in the tube space), 10E-07 (heat carrier – circulating kerosene in the tube space), and 10E-08 (heat carrier – fuel oil in the tube space).

The mixture heated in the heat exchangers of Stage I is fed into the condensate evaporator D01, where it is separated into gaseous and liquid hydrocarbon phases.

The gaseous phase from the top of the evaporator is directed straight to the atmospheric distillation column S02. The liquid hydrocarbon phase (bottom product) from the bottom of the evaporator is pumped through the apparatus of Stage II of preheating, which consists of a system of five sequentially connected heat exchangers: E09 (heat carrier – combined gas oil in the tube space), E10 (heat carrier – light gas oil in the tube space), E11 (heat carrier – fuel oil in the inter-tube space), E12 (heat carrier – heavy gas oil in the inter-tube space), E13 (heat carrier – circulating stream of heavy gas oil in the inter-tube space).

The feedstock heated in the heat exchangers of Stage II is fed into the lower part of the pre-fractionation column S01.

The liquid phase from the bottom of the pre-fractionation column S01 (bottom product) is pumped through two sequentially connected heat exchangers of Stage III preheating the condensate:

- E14 (heat carrier – circulating stream of heavy gas oil in the inter-tube space);
- E15 (heat carrier – fuel oil in the inter-tube space).

From there, the stream is directed to the coil furnace F01 for final heating.

This calculation and experimental research were carried out in accordance with the process regulations for atmospheric distillation of a mixture of oil and gas condensate at the Bukhara Refinery (Khudaiberdiev, 2019b; Khurmamatov et al., 2023b). The heated hydrocarbon feedstock is a working mixture of oil (30%) and gas condensate (70%). According to the regulations of the oil refinery, the technical parameters of the industrial tubular heat exchanger-condenser of the 10E-04 are determined, namely: the values of the technological modes of the process of heating the oil and gas condensate mixture (concentration of the mixture, flow rate of the mixture and heating coolant – naphtha, condensation temperature of vapors t_{con}, initial t_{in} and heated mixture in the apparatus t_{out}, incoming and outgoing temperature), limit values of the main properties of the mixture and heat-transfer (density, viscosity). Technological limitations to calculations were determined by the requirements of the enterprise standards for distillates of fuel fractions of total naphtha, straight-run kerosene, and straight-run diesel fuel produced by the refinery.

As it is known, the thermal preparation of feedstock for distillation and its rectification are energy-consuming processes at refineries. Therefore, with the constant growth of rates for oil and energy resources, large oil refineries cannot always meet modern requirements for the efficient use of electrical and thermal energy. This circumstance indicates the need to improve most technological processes, including the process of thermal preparation (heating) of feedstock for distillation, and to optimize the operation mode of the unit of heat exchange equipment in the oil refinery.

When solving the problem of optimizing the process of heating hydrocarbon feedstock, either the rational boundaries

Table 1. Technical and technological characteristics of heat exchangers E01-E08 at Bukhara Oil Refinery

Tabela 1. Charakterystyka techniczna i technologiczna wymienników ciepła E01-E08 w rafinerii ropy naftowej w Bucharze

Specification	Units	10E-01	10E-02	10E-03	10E-04	10E-05	10E-06	10E-07	10E-08
Total length of apparatus	[m]	5.927	6.600	7.893	7.674	7.430	7.561	7.620	7.631
Shell diameter and thickness	[m]	0.625/0.012	1.2/0.013	1.329/0.013	1.021/0.013	0.674/0.01	0.855/0.012	9.29/0.012	0.907/0.013
Number of baffles	[units]	8	9	10	9	8	8	11	10
Diameter of heat transfer tubes	[m]	0.02/0.025	0.02/0.025	0.02/0.025	0.02/0.025	0.02/0.025	0.02/0.025	0.02/0.025	0.02/0.025
Number of heat transfer tubes	[units]	288	580	1106	644	268	454	524	514
Length of heat transfer tubes	[m]	4.8	4.8	6	6	6	6	6	6
Pressure of mixture	[bar]	16.2	15.6	15.1	13.6	11.9	11	10.4	10
Initial temperature of mixture (t_1)	[°C]	20	25.6	49.1	96.1	111.7	119.7	126.3	142.3
Final temperature of mixture (t_2)	[°C]	25.6	49.1	96.1	111.7	119.7	126.3	142.3	150
Heat carrier		kerosene	general naphtha	general naphtha	general naphtha	kerosene	gas oil	kerosene	fuel oil
Oil and gas condensate mixture flow area		inter-tube	tube	tube	tube	tube	tube	tube	tube
Heat carrier pressure	[bar]	10.4	1.3	1.5	1.5	11	10.6	7.3	11.5
Initial temperature of heat carrier (t_3)	[°C]	125.56	148.72	164.95	136.6	138.83	159.44	170.67	257.65
Final temperature of heat carrier (t_4)	[°C]	103.22	146.01	151.71	114.3	125.56	136.85	167.04	185.32

of its technological parameters or the minimum required heat exchange surface is determined, which provides a given performance (thermal power) G_{rm} of the apparatus.

To identify the optimal boundaries of the mode of heating oil feedstock with the heat of fuel fractions, the technological cost of the heated feedstock C_t, which indirectly evaluates the apparatus's operating economy, was chosen as an optimality criterion. The technological cost of production includes the cost of raw materials, heat carriers, heat and electricity, salaries of maintenance personnel, and other costs (Boyarinov and Kafarov, 1975; Bochkarev, 2014; Karimi et al., 2021; Khudaiberdiev et al., 2022; Khurmamatov et al., 2023c).

Since oil is not subjected to technological processing during heating, its cost C_{rm} does not depend on the mode of operation of the heat exchanger. It should also be taken into account that the hot flows of fractions and fuel oil leaving the distillation column of the installation are subject to cooling to the temperature of their storage (Skoblo et al, 2000; Manovyan, 2001; Akhmetov et al., 2006; Glagoleva and Kapustina, 2006; Akhmetov, 2013). Based on this, in order to increase the thermal efficiency of an oil refinery, these hot streams are used for sequential multi-stage heating of raw materials in heat exchangers. Therefore, the costs associated with the use of heat from hot streams do not affect the technological cost of the heated mixture C_t in the apparatuses. In addition, the salary of technical personnel for maintenance of heat exchangers is fixed and it does not depend on the intensity of their operation. Taking into account the above circumstances, the costs associated with the purchase of raw materials, heat carriers, and the salary of technical personnel are not included in the expression of the optimality criterion of the process under study. Because of this, the objective function of the chosen optimality criterion for the process of heating oil feedstock in heat exchangers is formulated (Boyarinov and Kafarov, 1975; Khudaiberdiev, 2019a; Karimi et al., 2021; Khudaiberdiev et al., 2022).

$$C_t = C_{rm}G_{rm} + C_{hm}G_{hm} + C_e(N + N_d) + A_{he}F_{he} + A_p(N + N_d) \quad (1)$$

where:

C_{rm}, C_{hm}, C_e – represent the respective costs of raw materials, heating medium, and electricity,

G_{rm}, G_{hm} – denote the consumption rates of raw materials and the heating medium,

N, N_d – stand for the power of the pump for transferring oil and distillate fractions,

F_{he} – represents the heat transfer surface area of the heat exchangers,

A_{he}, A_p – are the depreciation charges for technological equipment and pumps.

The expenses on raw materials and the heating medium do not influence the technological cost of heated crude oil in heat exchangers. Due to the aforementioned circumstances, the costs associated with the purchase of crude oil, heating mediums, and salaries of technical personnel are not included in the expression of the optimality criterion of the investigated process:

$$C_t = C_e(N + N_d) + A_{he}F_{he} + A_p(N + N_d) \quad (2)$$

The cost of feedstock, which is not processed during heating, depends on its quality and not on the heat exchanger operation. Distillate fractions and bottom residue from the atmospheric column are cooled for storage. To boost refinery efficiency, these hot streams are used for multi-stage feedstock heating. Therefore, costs related to these streams do not affect the technological cost of the heated feedstock. Consequently, costs for purchasing feedstock and heat carriers are excluded from the optimality criterion.

The objective function of the optimality criterion (5.39) can thus be expressed as:

$$C_t = C_eN + A_{he}F_{he} + A_pN \quad (3)$$

As it is known, tubular heat exchange apparatuses of three preliminary heating blocks in the oil refining plant have different designs and productivity. For identifying the optimal composition of heat exchanger blocks in the oil refining plant and developing its energy-saving technological scheme, it is expedient to adopt the specific technological cost of heated raw material ($C_{stc} = C_t/G_{rm}$) as a criterion of optimality. In this case, (3) can be expressed as:

$$C_{stc} = 1/G_{rm}\,[C_eN + A_{he}F_{he} + A_pN] \quad (4)$$

A comparative assessment of the impact of the cost item on the technological cost of the heated mixture is carried out by analyzing the equations for calculating the parameters included in the objective function expression (4).

The power of the pump N [kW] required to overcome the hydraulic resistance in the fluid pumping system (Pavlov et al., 2006):

$$N = (G_{rm} \cdot \Delta P)/(1000\,\rho\eta_p) \quad (5)$$

where:

ΔP – hydraulic resistance of the flow pumping path [Pa],

ρ – flux density [kg/m^3],

η_p – pump efficiency coefficient.

The pressure loss to overcome the forces of internal friction in the tubes of the heat exchanger is determined by the formula (Pavlov et al., 2006):

$$\Delta P = 0.5\upsilon^2\rho\,(\lambda n \cdot 1/d_{equ} + \textstyle\sum\varphi_i) \quad (6)$$

where:

$\upsilon = 4\,G_{rm}/(\pi d_i^2\rho)$ – speed of raw material flow in the apparatus tubes [m/s],

$\lambda = f(Re)$ – coefficient of friction, determined depending on the mode of flow in the tubes according to the *Re* number,

$Re = (\upsilon d_i \rho)/\mu$ – Reynolds number,

μ – dynamic coefficient of viscosity of the feedstock [Pa·s],

$\sum\varphi_i$ – is the total coefficient of local resistance in the apparatus and pipelines (Pavlov et al., 2006).

The heat transfer surface *F* of heat exchangers, taking into account their productivity in terms of raw materials G_{rm}, is determined by the expression (Khudaiberdiev, 2019a; Çengel and Ghajar, 2020; Kharaji, 2021):

$$F = Q/(K\Delta t_d) = G_{rm}\,(c_{in}t_{in} - c_{out}t_{out})/(K\Delta t_d) \quad (7)$$

where:

$Q = G_{rm}(c_{in}t_{in} - c_{out}t_{out})$ – thermal load of the apparatus [W],

c_{int}, c_{out} – heat capacity of the raw material at the temperatures of its inlet to the apparatus tin and at its outlet t_{out} [J/(kg·°C)],

K – the overall heat transfer coefficient in the apparatus [W/(m²·°C)],

Δt_d – useful temperature difference between the raw material and the coolant [°C].

The condensation of vapors occurs with the counterflow direction of the heat carriers: fuel fraction vapors condense in the shell side of the shell-and-tube heat exchanger, on the outer surface of the heat transfer tubes, while the oil and gas condensate mixture flows inside the tubes.

The heat transfer coefficient from experimental data is determined by the formula. The calculated value of the heat transfer coefficient [W/(m²·K)] is found by the known formula (Chen and Ren, 2008; Kharaji, 2021):

$$K = 1/(1/\alpha_1 + \delta/\lambda + 1/\alpha_2) \quad (8)$$

where:

α_1, α_2 – are the heat transfer coefficients on the side of the condensing vapor and the heated mixture, respectively [W/(m²·K)],

$\delta_w = 0.0025$ m is the thickness of the tube wall,

$\lambda_w = 46.5$ W/(m·K) is the thermal conductivity of the tube wall material (Whitaker, 2013).

The heat transfer coefficient from the condensing vapor to the wall of the horizontal tube is determined by the equation (Barulin et al., 2009):

$$\alpha_1 = 0{,}72 \cdot \varepsilon_r[\lambda^3 \cdot \rho^2 \cdot r \cdot g/\mu \cdot (t_{con} - t_w) \cdot d_{out}] \quad (9)$$

where:

$\varepsilon_r \le 1$ is the row coefficient,

λ, ρ, and μ are the thermal conductivity [W/(m·K)], density [kg/m³], and viscosity [Pa·s] of the general naphtha at $t_{a.n} = 125.25$°C (its average temperature),

$r = (354.1 - 0.03768 \cdot t)/\rho_{15}^{15}$ is the latent heat of vaporization of the general naphtha [kJ/kg] determined by the Craig equation,

$t_{con} = 136.6$°C is the initial temperature of the general naphtha,

d_{out} – is the outer diameter of the tubes [m].

The temperature of the inner tube wall t_w is taken to be 5°C lower than the temperature of the general naphtha: $t_w = t - 5 = 131.6$°C.

The calculation of the heat transfer coefficient α_2 from the tube wall to the heated mixture in the condenser is performed in the following sequence.

The average velocity of the oil and gas condensate mixture flow in the tubes υ (m/s) is determined by the expression:

$$\upsilon = z \cdot 4G/(3600\pi \cdot n \cdot d_{in}^2) \quad (10)$$

The flow regime of the liquid is determined by the Reynolds number *Re*:

$$Re = (\vartheta \cdot d_{in} \cdot \rho)/\mu \quad (11)$$

where: ρ and μ are the density [kg/m³] and viscosity [Pa·s] of the heated mixture at its average temperature.

For a more accurate calculation of the density of hydrocarbon raw materials over a wide range of temperatures (up to 300°C), the equation of A.K. Manovyan is applied, which guarantees an accuracy of no less than 97% (Manovyan, 2001):

$$\rho_4^t = 1000\rho_4^{20} - \frac{0.58}{\rho_4^{20}}(t-20) - \frac{\left[t - 1200\left(\rho_4^{20} - 0.68\right)\right]}{1000}\cdot(t-20) \quad (12)$$

The heat capacity of oil feedstock *c* [kJ/kg·°C], taking into account its temperature *T* and relative density ρ_{20} is determined by the formula (Manovyan, 2001):

$$c = 1.5072 + \frac{T-223}{100}\cdot\left(1.7182 - 1.5072\rho_4^{20}\right) \quad (13)$$

Depending on the flow regime, the criterion equation for calculating the Nusselt number *Nu* is chosen (Kharaji, 2021):

- for laminar flow ($Re < 2320$)

$$Nu = 0.17 \cdot Re^{0.33} \cdot Pr^{0.43} \cdot Gr^{0.1} \cdot \left[\frac{Pr}{Pr_w}\right]^{0.25} \quad (14.1)$$

- for turbulent flow ($Re > 10\,000$)

$$Nu = 0.21 \cdot Re^{0.8} \cdot Pr^{0.43} \cdot \left[\frac{Pr}{Pr_w}\right]^{0.25} \quad (14.2)$$

Then, according to the value of *Nu*, the heat transfer coefficient α_2 [W/m²·K] from the heat transfer tube wall to the heated mixture flow is calculated (Kharaji, 2021):

$$\alpha_2 = Nu \cdot \lambda / d_{eq} \quad (15)$$

where:

λ – is the thermal conductivity of the heated (oil and gas condensate) mixture [W/m·K],

d_{eq} – is the equivalent diameter of the flow cross-section [m].

All this is reflected in the results of calculating the coefficients of heat output and heat transfer in heat exchangers using the well-known method (Skoblo et al., 2000; Pavlov et al., 2006; Çengel and Ghajar, 2020; Ismailov, 2023; Ismailov et al., 2024), where the indicators of physical, chemical and thermal physical properties of raw materials and heating agent are taken at average values of their temperatures. The main difference between our method of calculating heat transfer coefficients in heat exchangers and the known calculation methods is the consideration of the continuous change in the indicators of the properties of heat transfer media from the temperature, which increases the accuracy of the heat transfer coefficient calculation results up to 20–30% (Khudaiberdiev, 2019a; Çengel and Ghajar, 2020).

It was revealed that the amount of depreciation deductions A_a is a variable, depending on the duration T of intensive operation of heat exchangers. Basic parameter of heat exchange equipment – power [kW], which affects the efficiency of the hydraulic system as a whole (Khudaiberdiev, 2019a; Çengel and Ghajar, 2020; Khudaiberdiev et al., 2022):

$$A_a = (EC_a)/24\,TF_a \quad (16)$$

where:

$E = 0.15$ – standard coefficient of efficiency of capital investments in the industry,

C_a – cost of the apparatus [USD],

$F_a = \pi \cdot d \cdot n \cdot 1$ – total sum of the outer surfaces of the inner tubes of the heat exchanger.

Similarly, the expressions for depreciation charges for A_p pumps are:

$$A_p = (EC_p)/24\,TN_a \quad (17)$$

where:

C_p – cost of the pump [sum],

N_a – actual power of the pump [kW].

Thereby, the objective function of the optimality criterion for heating the oil and gas condensate mixture by hydrocarbon heat carrier in a horizontal tubular apparatus is formulated as a system of equations (4, 12, 13, 5, 6, 7, 16, 17):

$$C_{stc} = 1/G_{rm}\,[C_e N + A_{he} F_{he} + A_p N] \quad (4)$$

$$\rho_4^t = 1000\rho_4^{20} - \frac{0.58}{\rho_4^{20}}(t-20) - \frac{\left[t - 1200\left(\rho_4^{20} - 0.68\right)\right]}{1000}\cdot(t-20) \quad (12)$$

$$c = 1.5072 + \frac{T-223}{100}\cdot\left(1.7182 - 1.5072\rho_4^{20}\right) \quad (13)$$

$$N = (G_{rm}\Delta P)/(1000\,\rho\eta_{pump}) \quad (5)$$

$$\Delta P = 0.5\upsilon^2\rho(\lambda_{nl}/d_{equ} + \textstyle\sum\varphi_i) \quad (6)$$

$$F = G_{rm}(c_{out}t_{out} - c_{in}t_{in})/(K_{con} \cdot \Delta t_{mid}) \quad (7)$$

$$A_a = (EC_a)/24\;TF_a \quad (16)$$

$$A_p = (EC_p)/24\;T_p N_a \quad (17)$$

Limitations in the field of study of the objective function (4) are established by the temperature of the heated mixture at the output of the heat exchanger t_{lim} ($t_{out} \cdot 220°C$) (Tekhnologicheskiy reglament..., 2009).

The solution of the above system of equations (4), (5), (6), (12), (13), (16), (17) is reduced to identifying the optimal boundaries of the technological regime for heating the oil and gas condensate mixture in tubular heat exchangers operating in the mode of efficient use of the heat of the cooled fraction flows.

Results and discussions

To select the optimal number of heat exchange apparatuses for the first stage of hydrocarbon feedstock preheating, which is a system of 8 (eight) heat exchangers connected in series, a systematic analysis of the efficiency of this industrial equipment of the enterprise has been performed.

The statics of the process of heating the oil and gas condensate mixture, consisting of 30% oil and 70% gas condensate, by the heat of condensing vapors of the total naphtha fraction leaving the raw material pre-fractionation columns and the atmospheric distillation column of the Bukhara Oil Refinery primary distillation unit was studied. The industrial heat exchanger-condenser 10E-04 used for heating the operating mixture has the following design parameters (Tekhnologicheskiy reglament..., 2009): d = 20/25 mm, l = 6 m and n = 644 pcs, and its heat transfer surface in terms of d_{out} = 0.025 is $F_a = \pi \cdot l \cdot n \cdot d$ = 304 m^2. The heating of the mixture in the tubes of this apparatus was studied at the following regulated values of the technological parameters of the process (Tekhnologicheskiy reglament..., 2009; Khudaiberdiev and Rakhimzhanova, 2021): G_{rm} = 105 508 kg/h, ρ_{20} = 768 kg/m^3, t_{in} = 96.1°C, t_{out} = 111.7°C and naphtha in the annular space of the apparatus t_{con} = 136.6°C. The calculated value of the heat transfer coefficient from the pipe wall to the mixture is α_2 = 878 W/(m^2 · K), and the value of the overall heat transfer coefficient in the apparatus is K = 270 W/(m^2 · K).

The cost of the apparatus, according to factory data, is C_a = 40 492,175 USD. The installed power of the pump for pumping the oil and gas condensate mixture through the eight sequentially connected heat exchangers of the block is $N_a = 250$ kW, and the cost of the pump is $C_p = 19\,634.8$ USD. The depreciation deductions are calculated as follows:

For the heat exchanger 10E-04:

$$A_a = (EC_a)/24\ TF_a = 0.14 \cdot 40\,492.175/(24 \cdot 340 \cdot 273) = = 0.0025\ \text{USD/m}^2$$

For the pump:

$$A_p = (EC_p)/24\ T_pN_a = 0.14 \cdot 19\,634.8/(24 \cdot 340 \cdot 250) = = 0.00135\ \text{USD/kW}$$

Later, the objective function of the optimality criterion for the process of heating the oil and gas condensate mixture (4), (5), (6), (12), (13), (16), (17) in this heat exchanger-condenser 10E-04 was studied in relation to the operating conditions of the atmospheric oil distillation unit at the Bukhara Oil Refinery (Tekhnologicheskiy reglament..., 2009). The calculation of the objective function (4) was carried out at the above values of the design and technological parameters of the process – G_{rm}, G_d, t_{in}, t_{out}, and t_{con} are presented in Table 2.

Based on the results of calculations, curves were plotted for changes in unit costs [USD/kg] (Figure 2) – depreciation deductions for a heat exchanger (A) (curve 2) and energy costs required to carry out the process C_p (curve 1) versus the temperature of heating the oil and gas condensate mixture tout in the range of its temperature increase 96.1–111.7°C at a constant temperature of condensation of vapors of total naphtha $t_{con} = 136.6$°C.

Figure 2 shows that the behaviour in the components of the unit cost of heating the oil and gas condensate mixture in the heat exchanger is the cost of pumping the mixture by a pump through the tubes and process pipes of the apparatus, including depreciation of pumping equipment $C_p = C_eN + A_pN$, as well as depreciation deductions for the heat exchanger $A = A_aF$, depend on the heating temperature of the mixture.

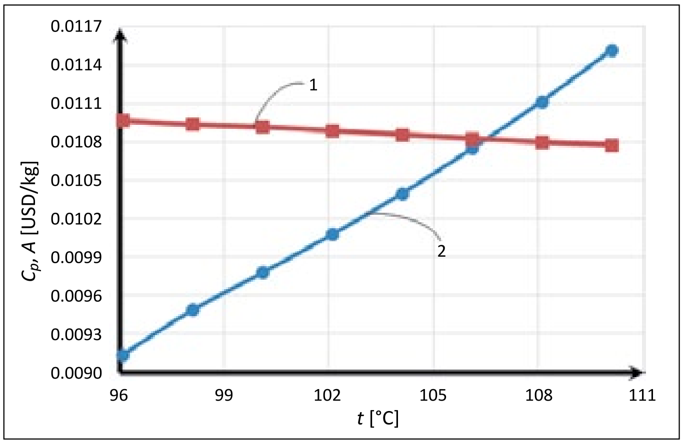

Figure 2. Variation in energy costs and depreciation deductions during the heating of oil and gas condensate mixture with heavy naphtha vapors in the heat exchanger 10E-04 at a throughput of $G_{rm} = 105\,508.3$ kg/h and $G_d = 21\,960$ kg/h;
curve 1 – energy costs required to carry out the process (C_p),
curve 2 – depreciation deductions for a heat exchanger (A)

Rysunek 2. Zmiana kosztów energii i odpisów amortyzacyjnych podczas podgrzewania mieszaniny kondensatu ropy naftowej i gazu ziemnego z ciężkimi oparami benzyny w wymienniku ciepła 10E-04 przy przepustowości $G_{rm} = 105\,508{,}3$ kg/h i $G_d = 21\,960$ kg/h; krzywa 1 – koszty energii niezbędnej do realizacji procesu (C_p), krzywa 2 – odpisy amortyzacyjne dla wymiennika ciepła (A)

The analysis of the values of the components of the cost of the heated working mixture C_{stc} is reduced to the following: according to the calculations, the number of depreciation deductions for the heat exchanger $A = A_aF$ (curve 2), depending on the mode of its operation, intensively increases along the sloping curve from 106.66 to 134.39 sums/kg. To pump a given flow rate of the mixture $G_{rm} = 105\,508.3$ kg/h through the tubes of the apparatus, $N = 8.1$–8.08 kW of power will be required.

Table 2. The computed optimal criterion results for heating a mixture consisting of 30% oil and 70% gas condensate using heavy naphtha vapors in heat exchanger 10E-04 determined at operational conditions with a throughput of $G_{rm} = 105\,508.3$ kg/h and $G_d = 21\,960$ kg/h, alongside a condensation temperature of 136.6°C

Tabela 2. Obliczone wyniki optymalnego kryterium dla ogrzewania mieszaniny składającej się z 30% ropy i 70% kondensatu gazowego przy użyciu ciężkich oparów benzyny w wymienniku ciepła 10E-04 określone w warunkach roboczych przy przepustowości $G_{rm} = 105\,508{,}3$ kg/h i $G_d = 21\,960$ kg/h, przy temperaturze kondensacji 136,6°C

t_m [°C]	Δt_m [°C]	K [W/(m^2 · °C)]	F [m^2]	$A_{he}F_{he}$ [$/kg]	$C_eN_p + A_pN_p$ [$/kg]	C_{stc} [$/kg]
103.9	32.7	266.350	184.21	0.00914	0.01097	0.0239
104.9	31.7	267.050	191.26	0.00949	0.01094	0.0240
105.9	30.7	267.769	196.97	0.00978	0.01092	0.0241
106.9	29.7	268.480	203.05	0.01008	0.01089	0.0242
107.9	28.7	269.210	209.56	0.01040	0.01086	0.0243
108.9	27.7	269.950	216.54	0.01075	0.01083	0.0244
109.9	26.7	270.000	224.03	0.01112	0.01080	0.0246
110.9	25.7	271.400	232.09	0.01152	0.01078	0.0247

Taking into account the cost of electricity $C_e = 0.00343$ USD/kW (data for 2021) for refineries and depreciation of pumping equipment, the total costs for pumping a given flow rate of the mixture $C_p = N(C_e + A_p)$ falls in the range from 0.01097 to 0.01078 USD/kg (curve 1). We consider that the point of intersection of curves 1 and 2, where the value of the heat exchange surface of the apparatus $F_{inter} = 219$ m^2, $C_p = 0.010825$ USD/kg, $A = 0.010825$ USD/kg, $C_{st} = 0.02165$ USD/kg and temperature of the heated mixture at the outlet of the apparatus $t_{out} = 107.23$°C, characterizes the optimal operating conditions of the 10E-04 heat exchanger at its given productivity $G_{rm} = 105\ 508.3$ kg/h in the conditions of the Bukhara Oil Refinery.

The comparable cost of C_p and A is explained by the high price of imported pumps used at refineries (Grigoryev et al., 1999; Khudaiberdiev, 2012, 2019a; Khudaiberdiev et al., 2012).

Conclusion

Thus, through the investigation of the objective function of the optimality criterion for heating the oil and gas condensate mixture in a shell-and-tube apparatus, the optimal values of the technological parameters for the process in the industrial heat exchanger 10E-04 at the Bukhara Oil Refinery were identified: $F = 219$ m^2, $C_p = 0.010825$ USD/kg, $A = 0.010825$ USD/kg, $C_{stc} = 0.02165$ USD/kg, and $t_{opt} = 107.23$°C.

As shown, in this performance mode, the reserve of the heat transfer surface of the heat exchanger 10E-04 is $\Delta F = [(304 - 219)/304]100\% = 28\%$ (12) or 85 m^2. This indicates insufficient utilization of the thermal capacity of the apparatus, as well as the possibility of further increasing the flow rate of the heated mixture in the apparatus.

The proposed methodology for optimizing the heating of crude oil and gas condensate includes a comprehensive analysis of the energy efficiency of heat exchangers across three stages of thermal preparation before primary distillation. It encompasses evaluating current efficiency, optimizing operational modes, developing modernization recommendations, conducting a comprehensive performance assessment, and considering specific operational conditions. This holistic approach aims to enhance heat exchanger efficiency, reduce energy consumption, and improve overall process performance.

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