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Research article

Process integration of crude oil distillation with technological and economic restrictions

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ABSTRACT

The petrochemical industry is one of the most important industries in the world economy. In the largest oil-producing countries, more than half of GDP is generated by hydrocarbons production and refining. Reduction of oil prices and new environmental regulations are forcing petrochemical companies to improve their energy efficiency.

Improvement of the energy efficiency of Crude oil distillation process at atmospheric vacuum distillation unit (AVDU) with a capacity of 3.3 million ton per year is considered in this paper. The amount of fuel spent for reprocessing of one ton of crude oil has been defined and it is 3.79 kg of natural gas.

This paper shows the ways to achieve the objectives of retrofit in the context of administrative and technical restrictions. The retrofit goal was achieved by the retrofit of the heat exchange network, which allowed reducing gas consumption by 0.98 t/h (natural gas).

The provided case studies show the pathway for efficient retrofit of crude oil distillation and most profitable ways for investment taking into account various administrative and technical constraints. The results of this work allow achieving reduction of energy consumption by 26%.

1. Introduction

World refining industry continues to evolve, but in many countries, its development is hampered due to unreasoned fiscal policy and vague legislation in the area of alternative fuels, which leads to a slowdown in the growth of energy efficiency of industries and low depth of the refinery of the country's (Meshcheryakova, 2015). Refining and petrochemicals are energy intensive industries and energy consumption affects to the finished product cost vastly. Increased energy efficiency in processes at existing refineries is an important element of sustainable development for many oil-producing countries.

In this work during an investigation of Oil Refinery, atmospheric and vacuum distillation processes were examined. With the help of stationary and portable devices, the measurements of process streams parameters were made. There were measured streams temperatures, stream flow rates and compositions of the waste gases of the furnace. The cooling water stream flow rate and temperature of cooling water were measured as well as fuel consumption in the furnaces.

When analyzing the statistical data of the unit for three years, it was determined that the unit consumes mainly thermal energy (Fig. 1).

There are several approaches for exchanger networks design of chemical processes that allow reducing energy consumptions. Process Integration is the most common method to reduce energy consumption in processing industry (Smith, 2016). The main areas are methods of mathematical programming and Pinch Analysis.

Methods of mathematical programming allow solving practically any problem of optimization of energy consumption in industry Pavão et al., 2018 in their work, a meta-heuristic two-level method based on Simulated Annealing and Rocket Fireworks Optimization (SA-RFO), originally developed for single-period HEN synthesis, is adapted to handle multiperiod HEN optimization. Alipour et al., 2018 in their paper presented a multi-follower bilevel programming approach to solve the 24-h decision-making problem faced by a combined heat and power (CHP) based micro-grid (MG). The methodology based on heat integration and mixed integer linear programming to represent process energy requirements with different heat exchange interfaces was proposed by Bütün et al., 2018. Kim et al., 2010 proposed approach consists of unit modeling using thermodynamic principles, mass and energy

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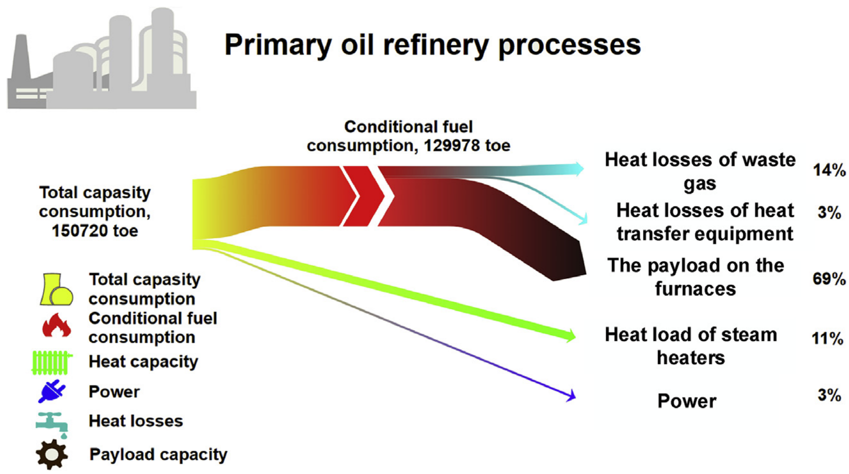

Fig. 1. Energy consumption structure for a typical primary petroleum refining process (for 3 years of operation work).

balances, development of a multi-period Mixed Integer Linear Programming model for the integration of utility systems in an industrial complex, and an economic/environmental analysis of the results. Methods of mathematical programming can be applied to optimize water resources. So Agana et al., 2013 presented the application of an integrated water management strategy at two large Australian manufacturing companies that are contrasting in terms of their respective products. Despite its effectiveness, mathematical methods have their shortcomings, the main one of which is the complexity of checking the results obtained. This greatly complicates the use of mathematical programming in the optimization of real production.

Second direction in the heat integration is Pinch Analysis. Pinch Design method has been developed by Linnhoff and Flower (1978) and is applied to find an optimum energy policy in chemical process industries.

Pinch Analysis due to its simplicity is widely used for targeting of energy consumption, designing the Heat Exchanger Network and identifying Process Integration opportunities. Fodor et al., 2012 modified Total Site Heat Integration methodology and showed possibilities of using the graphical approach. Farhata et al., 2015 presented a new methodology combining Total Site analysis with exergy analysis. Liew et al., 2012 demonstrated the possibility of using the numerical methodology. Liu et al., 2017 and others proposed new absorption-stabilization process with a two-stage condensation section. Compared with the existing process, the proposed process can reduce the cold utility and hot utility by 17.98% and 25.65%, respectively, as well as decrease the total annual operating costs of the heat exchanger network by 17.48%.

Ul'ev and Vasil'ev, 2015 examined the potential for the application of the Pinch method in the coke-chemical industry, including a Total Site approach (Ulyev et al., 2013).

The minimum energy consumption for the overall process of biodiesel synthesis from vegetable oil and methanol was presented Pleşu et al., 2015 using the Pinch Analysis method.

Despite the fact that the Pinch Analysis method has long been known and is widely used in the design of optimal heat exchange networks, the method is constantly evolving, including its graphics component. New graphical technique, based on Pinch Analysis, for the grassroots design of heat exchanger networks, was presented by Gadalla (2015). Pinch Analysis is widely used to assess the potential of using heat pumps (Olsen et al., 2017).

Pinch Analysis is widely used not only for heat integration but also for optimizing power consumption (Fernández-Polanco and Tatsumi, 2016). The use of the Pinch method allows receiving substantial financial profit by minimizing of energy consumption for the account of maximizing heat recovery within the considered energy-technological system.

The flexibility of the Pinch Analysis method makes it possible to use it in financial planning. Roychaudhuri and Bandyopadhyay, 2018 proposed a new algorithm, the minimum opportunity cost targeting algorithm, to address the capital budgeting problems for selecting environmental management projects. This algorithm is based on the principles of Pinch Analysis, a well-established resource conservation methodology and can be directly applied to partially acceptable projects that can be formulated as a linear programming problem.

The literature review shows the universality and wide distribution of the Pinch Analysis method, as well as its effectiveness in various areas of industry, which ensures transparency of the results obtained.

However, in reality, it is not always possible to implement a completely new heat exchange network. This is especially true for oil refining, where the initial heat exchange network has a high heat transfer area and the space for new heat exchangers is limited. Modern methods of process integration do not offer system solutions that take into account the available heat exchange capacities. For example Boldyryev et al., 2016 tried to implement a Process Integration measures accounting limited process conditions in cement manufacturing, such as streams with solid particles, solid–gas and solid–air heat transfer. This paper proposes an updated methodology of improved heat integration of crude oil refinery with additional variables that generated by the feasibility of unit retrofit. There are some technological, administrative and economic restrictions including limited unit space, equipment purchasing time, start-up and shut-down schedule, target saving etc. The novelty of current research is to find the optimal retrofit pathways taking into account technical, economic and administrative variables and get more realistic and sustainable in terms of future operation.

2. Methods

In this work the solution for the achievement of the objectives, near optimal to the decision. This methodology allows you to maintain the maximum possible number of existing heat exchangers. Heat exchangers network is crude oil distillation was optimized taking into account the next restrictions:

- Limited purchase time for new equipment.
- Separate tender as each type of equipment.
- Reduce energy consumption by at least 10%.
- Minimum changes to existing PFD.
- Limited space for new equipment.
- Retrofit is possible only during planned repairs.

Using the classical method of analysis, we obtained design targets:

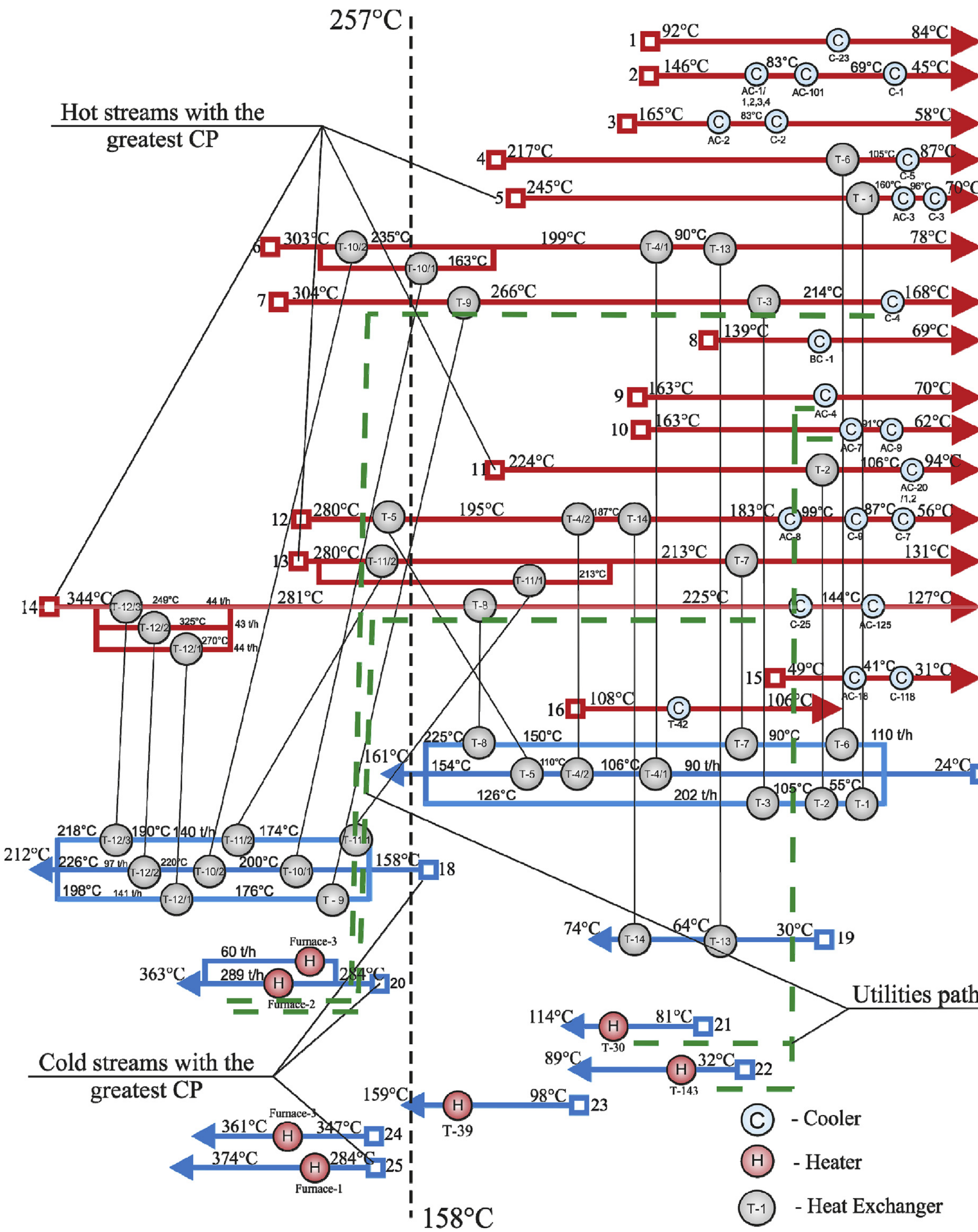

Fig. 2. Retrofit path of HEN.

- Hot utility is 41.9 MW.
- Cold utility is 45.0 MW.
- Heat recovery is 37.0 MW.
- To achieve these targets, it is necessary to install 10056.5 m^2 of the heat exchange surface.
- Capital costs at the same time will be about 2 million dollars.
- To achieve these targets, it will be necessary to completely change the existing heat exchange system.

The investigated process has 19 heat exchangers and the customer is not ready for a large-scale modification of the Heat Exchanger Network (HEN).

The proposed methodology uses utility paths in conjunction with the analysis of the efficiency of existing heat exchangers. Also, the priority is given to technological flows with the highest flow heat capacity (Fig. 2).

At the selected technological flows (Fig. 2), we substituted hot utility with a high-temperature hot stream with a large energy potential and redistributing the heating of the cold streams for cold utilities decreasing. Thus, we achieve the energy goal retaining most existing heat

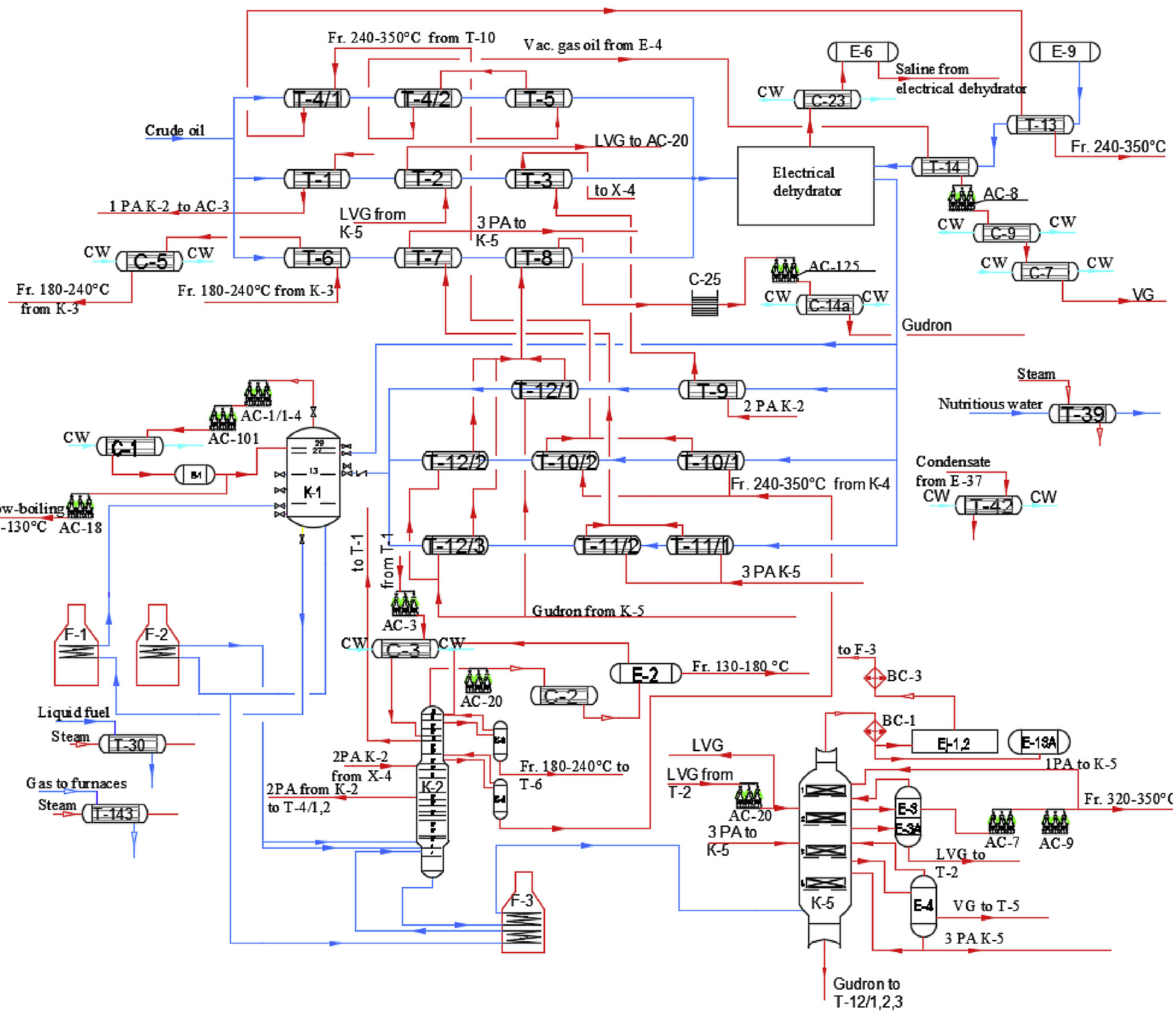

Fig. 3. Flowsheet of the Crude oil distillation unit. K – distillation column; C – cooler; F – furnace; T – Heat exchanger; AC – Air Cooler; E – Tank; BC–barometric condenser; Ej – ejector.

exchangers. Due to technological restrictions, the use heat potential of vapors from the top of the distillation columns is limited.

3. Results and discussion

3.1. Process description

Crude oil enters the AVDU-4 and is pumped through heat exchangers: T-1, T-2, T-3, T-4/1, T-4/2, T-5, T-6, T-7, T-8, where it is heated by the heat of the Pump Around (PA) of the atmospheric and vacuum columns, Fr. 180–240 °C, vacuum gas oil, Gurdon, and Fr. 140–160°C (Fig. 3).

Desalted and dehydrated oil from an electrical dehydrator is pumped through heat exchangers: T-9; T-12/1; T-10/1,2; T-12/2; T-11/1,2; T-12/3, where it is heated by the heat of II PA K-2, fraction 240–350 °C, III of the PA K-5 and Gurdon to a temperature is 212 °C (Fig. 3).

During the inspection of the AVDU-4, flows that are involved in heat exchange have been determined. These are flows that are heated or cooled by pumping through heat exchange equipment, namely: ovens, refrigerators and recuperative heat exchangers. The temperatures of the streams that are involved in the heat exchange with the help of portable pyrometers were measured.

Using the material balance of the unit, the mass flow rates were determined. The mass heat capacity of the flows, as well as the mass evaporation heat for hydrocarbon flows, was determined on the basis of the computer model. To refine the thermal balance a computer model of the process has been created using the UniSim design software (Seider et al., 2003). Basic information about the computer model is presented in Appendix 1.

Since on the of AVDU-4 mass expenditures are measured only for products leaving the installation, internal flows such as Pump Around (PA) and Hot Jets are determined on the basis of the temperature measurements and the heat balance of the process using the UniSim design software.

For example, 1 Pump Around of column K-2 (stream No. 5) was determined based on the measured initial and final flow temperatures, which were 245 °C and 70 °C, respectively, the flow rate was calculated automatically in the computer model. The conditions for the calculation were: initial and final flow temperature, quality of the products obtained, the mass flow of unstable gasoline from the top K-2.

The flows that pass through the kilns (flows 20, 24 and 25) were

Table 1
Parameters of process flows for the Crude oil distillation unit (Fig. 1).

№	Stream	Type	T_S, °C	T_T, °C	G, t/hour	CP, kW/K	ΔH, kW
1.	Wastewater from dehydrators	Hot	92	84	16.3	18.84	151
2.	Vapors from the top of column K-1 (condensation) + condensate cooling	Hot	146	45	67.2	487.8	9106
3.	Vapors from the top of column K-2 (condensation) + Condensate cooling (Fr. 130–180°C)	Hot	165	58	65.0	632.2	11,415
4.	Fr. 180–240°C from K-3	Hot	217	87	57.4	38.9	5057
5.	1 Pump Around K-2	Hot	245	70	65.4	44.7	7814
6.	Fr. 240–350°C from K-4	Hot	303	78	59.0	41.7	9387
7.	2 Pump Around K-2	Hot	304	168	66.3	49.9	6792
8.	Vapors from the top of column K-5 (condensation)+ Condensate cooling	Hot	139	69	8.2	472.5	1076
9.	Fr. 320–350°C	Hot	163	70	9.4	5.8	537
10.	1 Pump Around K-5 (Fr. 320–350°C)	Hot	163	62	21.7	13.3	1341
11.	2 Pump Around K-5	Hot	224	94	79.1	51.5	6698
12.	Heavy vacuum gas oil	Hot	280	56	43.4	28.3	6344
13.	3 Pump Around K-5	Hot	280	131	72.1	49.7	7409
14.	The tar from K-5	Hot	344	127	131.0	90.9	19,727
15.	Gasoline fraction from the unit	Hot	49	31	67.6	40.5	729
16.	The steam condensate from E-37	Hot	108	106	6.2	7.2	14
17.	Crude oil	Cold	24	161	401.7	239.5	32,815
18.	Desalted oil	Cold	158	212	378.3	260.9	14,090
19.	Wash water to electrical dehydrator	Cold	30	74	9.1	10.5	464
20.	Supply stream to column K-2+ Evaporation	Cold	284	363	349.3	291.8	28,317
21.	Liquid fuel to furnaces	Cold	81	114	2.5	1.4	47
22.	Gas to furnaces	Cold	32	89	7.2	4.5	258
23.	Nutritious water	Cold	98	159	12.6	14.9	908
24.	Fuel oil from the bottom of K-2	Cold	347	361	228.6	189.9	2659
25.	Hot jet K-1 + Evaporation in Hot jet K-1	Cold	284	374	201.8	332.9	18,660

calculated taking into account the fuel burned in the furnaces and the efficiency of the furnaces, these data were determined as a result of the inspection of the furnaces. The composition of the waste gases and the heat losses from the furnace walls were measured. The obtained results allowed to determine the load on evaporation in the streams. After that, the results obtained were checked in the computer model using the UniSim design software.

The parameters of a process flow for the flow sheet of crude oil distillation are given in Table 1.

The composite curves (Fig. 4) at a specified recuperation value provide a possibility to determine the minimum temperature drop between the hot and cold composite curves (ΔT_{min}), which corresponds to vertical heat exchange in a recuperative system.

Based on the analysis of the process flowsheets (Fig. 3) and the flow data (Table 1), the networks of heat-exchange systems were constructed

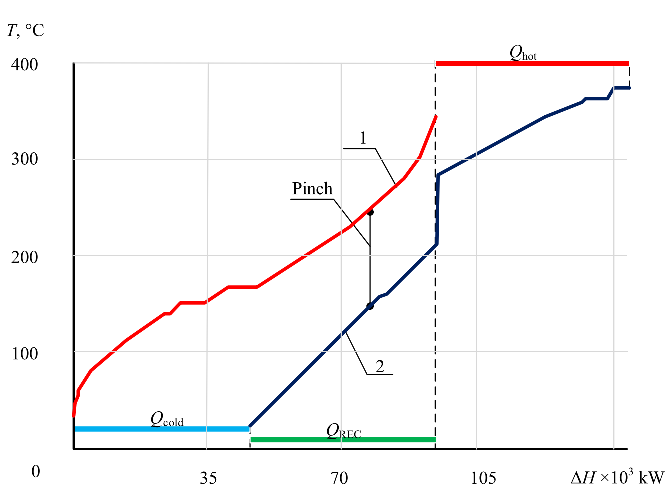

Fig. 4. Composite curves of existing process. 1 and 2 – hot and cold composite curves, Q_{Hmin} and Q_{Cmin} are the capacity of hot and cold outer-energy carriers, Q_{REC} is the recuperation capacity. Hot utility is 50.5 MW, cold utility is 45.0 MW, heat recovery is 47.6 MW. Hot Pinch temperature is T_{hot} = 257°C, cold Pinch temperature is T_{cold} = 158°C, minimum temperature difference is 99°C.

(Fig. 5). The networks display process flows, including outer energy carriers and heat-exchange operations between them.

The smallest distance between the composite curves along the ordinate axis is shown by the Pinch location. The heat balance for existing recuperative heat exchangers is presented in Table 2.

3.2. Retrofit

To implement the project of retrofit, the Pinch Analysis method, which had demonstrated its efficiency in the studies performed earlier in organic chemical (Ulyev et al., 2016), inorganic chemical (Boldyryev and Varbanov, 2015) and petrochemical industries (Tovazhnyanskii et al., 2009) was selected. An advantage of this method is the possibility of finding the minimum discounted cost of the project, which is governed by economic and thermodynamic laws (Smith, 2016).

The selection of an optimal retrofit project is performed by attaining a value of ΔT_{min} at which the reduced costs will be minimal. This value is attained by a compromise between the reduced energy and capital costs. Some reduced cost dependences on the minimum temperature drop were plotted in Fig. 6 using the HILECT software (Boldyryev et al., 2017). For the considered process to be optimally integrated, it was evaluated the principal capital and specific cost values, which significantly affect the cost of a project.

The most economically important values that materially affect the present value of the project must be determined in order to economically optimal integration process (Odejobi et al., 2015).

The cost of the hot utilities used in the process is assumed to be 140 USD per 1 kW year given the fact that in the year 8400 working hours.

Necessary capital investment and payback period can be assessed before the project using the price of heat transfer equipment, provided by the manufacturer. The capital cost of the heat exchanger can be defined by Eq (1):

$$\text{Capital cost} = A_T + B_T \cdot (S)^C \quad (1)$$

where: A_T = 40000 USD – the cost of installing a heat exchanger; B_T – rate equivalent to the cost of 1 m^2 surface area of heat transfer, for heat exchangers B_T = 2000; S – effective surface area of heat exchangers; C– factor reflecting the non-linear dependence on the value of the cost of the heat exchanger to the heat transfer surface, C = 0,87 (Klemeš et al.,

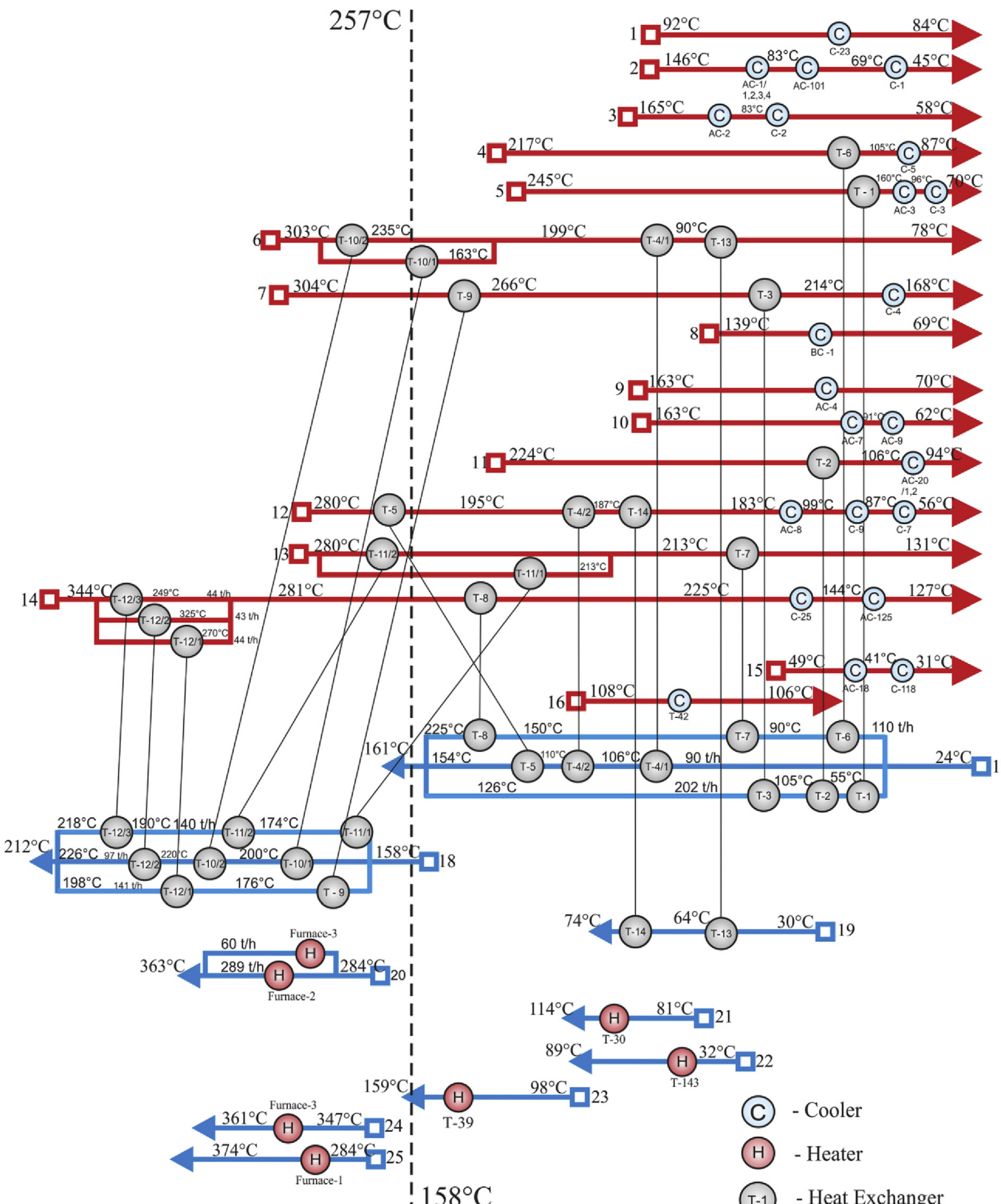

Fig. 5. Grid diagram of existing process: H is a heater, C is a cooler.

2015).

The construction of cost curves for the considered process (Fig. 6) has enabled the estimation of $\Delta T_{\min.opt}$, which was 6 °C. The construction of composite curves for the system of flows in each engineering process with allowance for found $\Delta T_{\min.opt}$ (Fig. 7) provides a possibility to estimate the target energy values for the reconstruction project.

The targets for the reconstruction project will be: heat recovery is 56.6 MW; the useful capacity for heating the streams is 41.9 MW; the useful capacity of cooling the streams is 37.0 MW; hot temperature of Pinch is 290°C; the cold temperature of Pinch is 284 °C.

However, as it was explained earlier it is not always possible to design and implement a completely new heat exchange network. In order to assess the potential use of existing heat exchangers, it is necessary to analyze their operating parameters (Table 3).

Analysing the existing heat exchangers, we can conclude that many of them work not in the optimal mode, which leads to a low value of heat transfer.

However, the heat transfer coefficients, which are indicated in the equipment passports and the analysis carried out during the inspection of the installation, indicate that many of the existing heat exchangers can work much more efficiently.

Composite Curves shows that the heating of cold streams occurs

Table 2
Heat balance for existing recuperative heat exchangers.

№ HE	Hot streams		Cold streams		Heat loss, kW
	№ Strems	Heat load on the hot side, kW	№ Strems	Heat load on the cold side, kW	
T-6	4	4357	18	4329	28
T-1	5	3796	18	3734	62
T-10/2	6	1419	19	1338	80
T-10/1	6	2921	19	2810	111
T-4/1	6	4548	18	4401	147
T-13	6	501	20	458	43
T-9	7	1898	19	1751	147
T-3	7	2597	18	2529	67
T-2	12	6079	18	6023	57
T-5	13	2543	18	2361	182
T-4/2	13	239	18	215	25
T-14	13	120	20	105	14
T-11/2	14	1666	19	1545	121
T-11/1	14	1666	19	1545	121
T-7	14	4078	18	3936	142
T-12/3	15	2901	19	2704	197
T-12/2	15	567	19	461	106
T-12/1	15	2259	19	2140	120
T-8	15	5091	18	4919	171

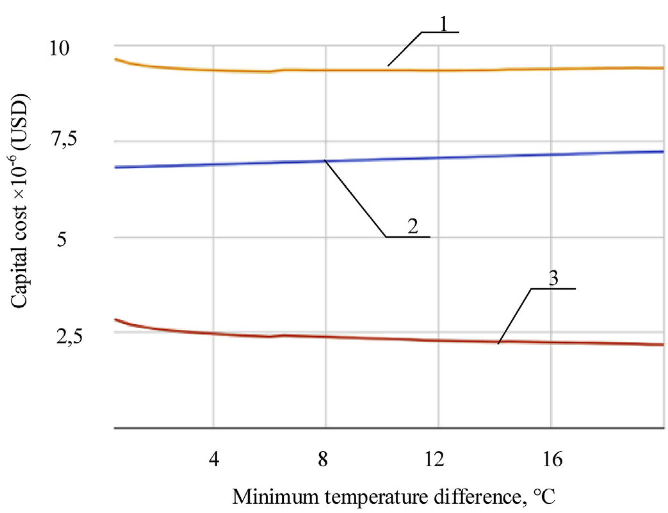

Fig. 6. Cost curves of Crude oil distillation process (Fig. 1). 1 – Total costs; 2 – Energy costs; 3 – Investments.

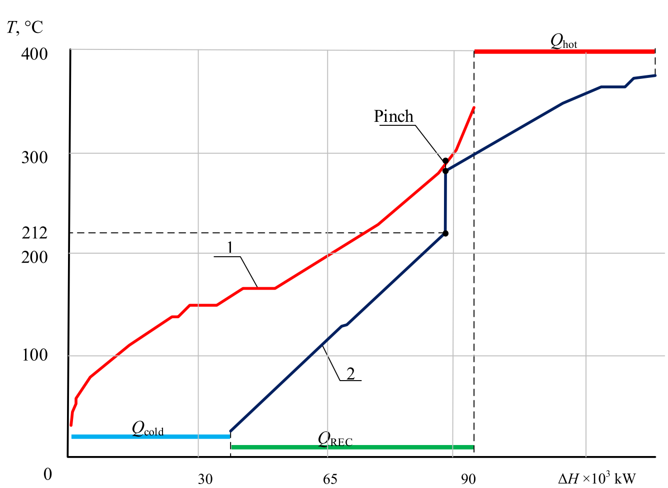

Fig. 7. Composite curves after reconstruction for found ΔT_{opt}. 1 and 2 – hot and cold composite curves, Q_{Hmin} and Q_{Cmin} are the capacity of hot and cold outer-energy carriers, Q_{REC} is the recuperation capacity.

Table 3
Characteristics of existing heat exchangers.

№ Heat Exchanger	Q, kW	Heat transfer area, м^{2}	ΔT_{LN}, °C	Heat transfer coefficient, K·kW/(m^{2}°C)
T-6	4329.1	480	102.28	0.09
T-1	3734.0	480	161.50	0.05
T-10/2	1353.8	480	55.59	0.05
T-10/1	2842.9	480	32.39	0.18
T-4/1	4400.7	480	78.73	0.12
T-13	3,58.1	320	35.88	0.03
T-9	1771.1	480	117.72	0.03
T-3	2529.5	240	123.85	0.09
T-2	6022.6	826	80.26	0.09
T-5	2361.3	480	104.16	0.05
T-4/2	214.7	480	82.98	0.01
T-14	1,05.3	320	115.97	0.01
T-11/2	1563.1	480	60.99	0.05
T-11/1	1563.1	480	77.73	0.04
T-7	3935.6	480	51.21	0.16
T-12/3	2735.5	826	88.30	0.04
T-12/2	406.1	826	111.37	0.01
T-12/1	2164.7	826	118.10	0.02
T-8	4919.5	826	65.04	0.09

smoothly to 212 °C, after which a sudden increase in the temperature of the cold streams to 284 °C is due to the process of oil separation in the K-1 column.

Mostly, the existing heat exchangers can be used to heat the cold flow to 212 °C, where the logarithmic temperature difference is high (Fig. 7).

In order to increase the efficiency of heat transfer in existing heat exchangers, it is necessary to exclude heat transfer through Pinch by designing a new heat exchangers network. (Fig. 8).

The temperature of the crude oil before the front of the electric dehydrators was reduced to 130° for optimum desalination process (Fig. 8).

3.3. Results

The developed heat exchange network made it possible to use 18 existing heat exchangers, which significantly reduced costs during the reconstruction of the Crude oil distillation unit.

The comparison of the optimal goals for the reconstruction project and the achieved goals in the reconstruction project is given in Table 4.

The use classical methods of processes integration are confronted with restrictions during the retrofit of real productions. Restrictions can be both administrative and technical. The proposed methodology makes it possible to achieve the goals that blazes to the optimum while minimizing changes in the existing heat exchange system. The disadvantage of this approach is that it is not possible to fully achieve optimal goals.

This case study shows the way of reconstruction for continuous production, in which the initial heat exchange network has a significant surface. The design of a completely new (optimal) heat exchange scheme for continuous production leads to significant costs in connection with the increase in the terms of planned repair, which significantly affects the economic efficiency of reconstruction projects. Proposed let it allow to carry out the reconstruction of production in terms of planned repair (30–45 days) or during the work of the unit. In fact, the proposed path is the only way to approximate the efficiency of existing plants, many of which were designed in the 60s - 80s of the last century to the existing energy consumption standards.

4. Conclusion

Process integration of crude oil distillation with technological and economic restrictions was carried out in this work.

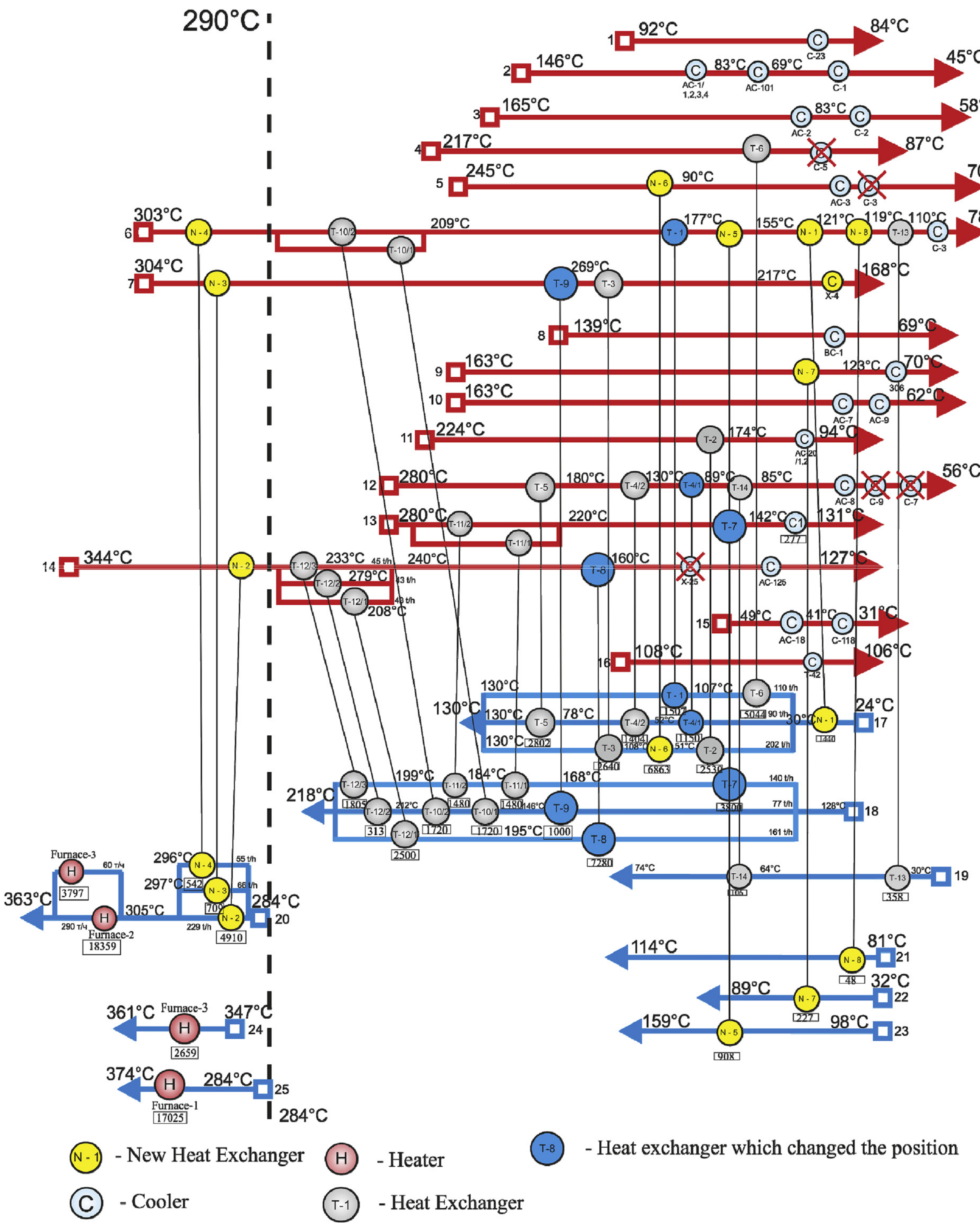

Fig. 8. Network diagram of the retrofit project. New heat exchangers are highlighted in yellow; Heat exchangers that have changed the position are highlighted in blue. (For interpretation of the references to colour in this figure legend, the reader is referred to the Web version of this article.)

Table 4
Comparative characteristics of energy values for the reconstruction project.

	Target vales	Achieved values
Hot utilities, MW	41.9	43.5
Cold utilities, MW	37.0	33.7
Recuperation, MW	56.6	59.9
New heat transfer area, m^2	10,056.5	3171.3
Cost of new heat exchangers, USD	209,495,6	115,956,9

The application of Pinch Analysis together with an analysis of the potential use of existing heat exchangers for the integration of Crude distillation oil processes has allowed decreasing the consumption of hot and cold utilities by 26% and 25%, respectively. These values are close to optimum. At the same time, capital costs are 1.8 times less than the implementation of a completely new (optimal process). This is due to much lower costs for new heat exchangers. The discounted payback period for the considered project is 4 years. The obtained engineering and economic results indicate that the presented ways of reconstructions are economically feasible.

The provided case studies show the pathway for efficient retrofit of the petrochemical industry and most profitable ways for investment. The results of this work can be used in petrochemical industry and other industries for efficient energy use, CO_2 mitigation and sustainability improvement of industrial regions.

Nomenclature

ΔH	change of enthalpy, kW
CP	flowrate heat capacity, kW
Q_{Cmin}	requirement for cold utility, kW
Q_{Hmin}	requirement for hot utility, kW
Q_{REC}	heat recovery, kW
T_S	supply temperature, °C
T_T	target temperature, °C
G	flowrate, t/h

Appendix. №1

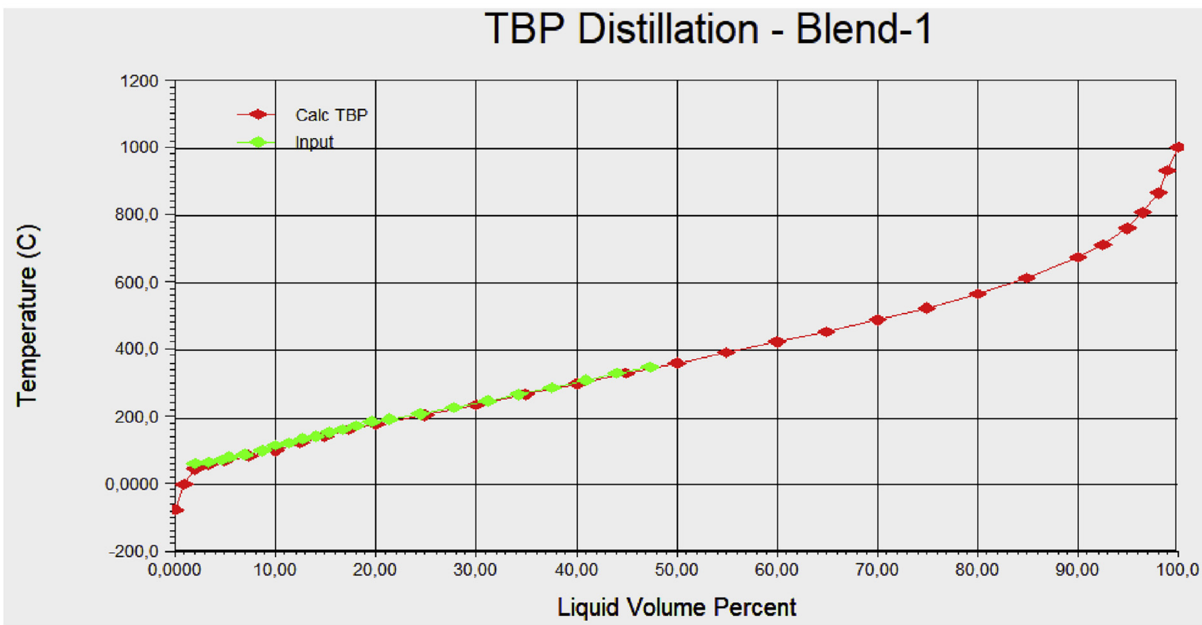

Fig. 1. Crude oil TBP Distillation Blend.

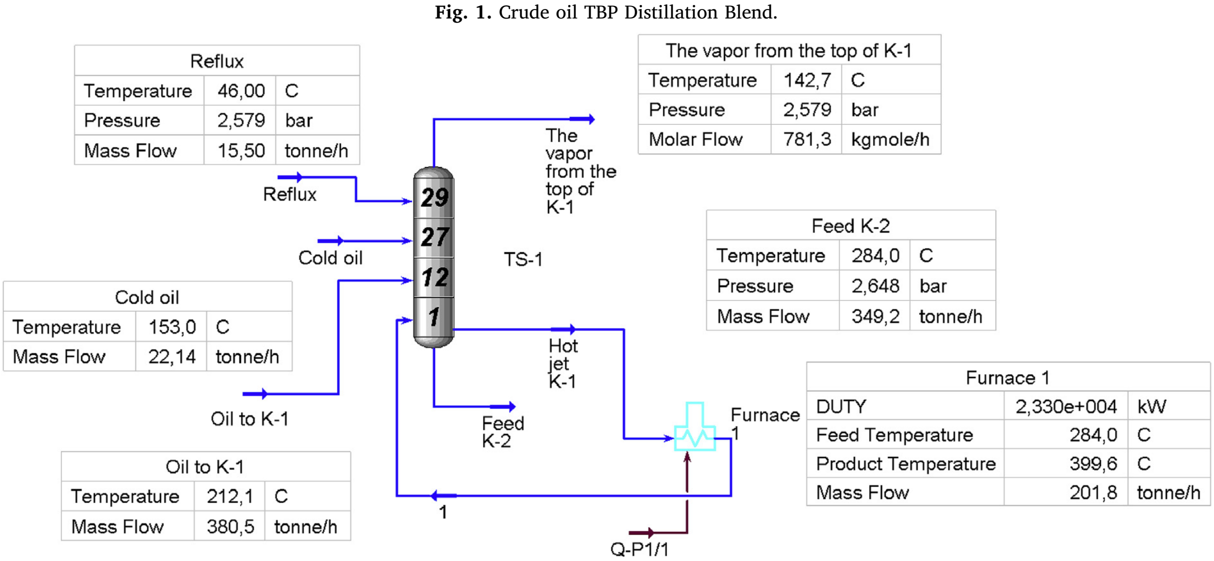

Fig. 2. The configuration of the first atmospheric distillation column (K-1).

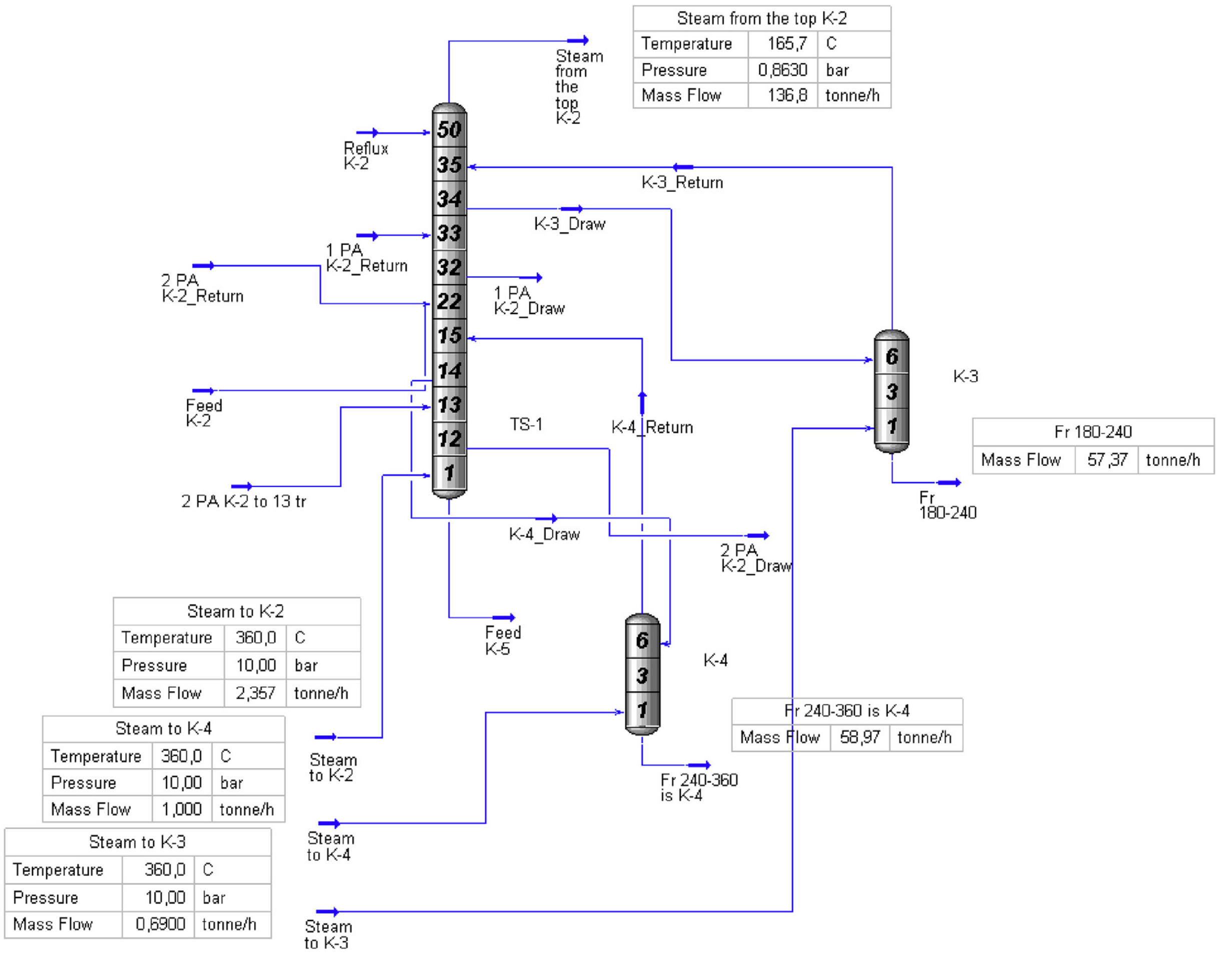

Fig. 3. The configuration of the second atmospheric distillation column (K-2).

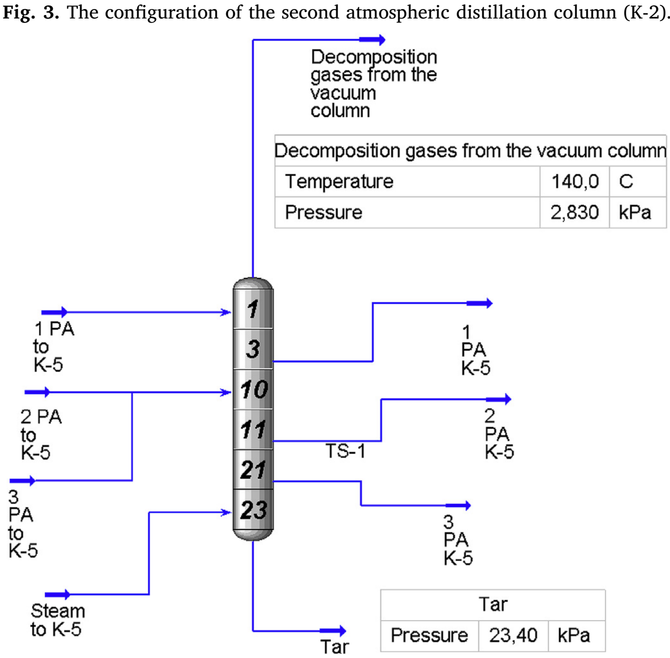

Fig. 4. The configuration of the vacuum column (K-5).

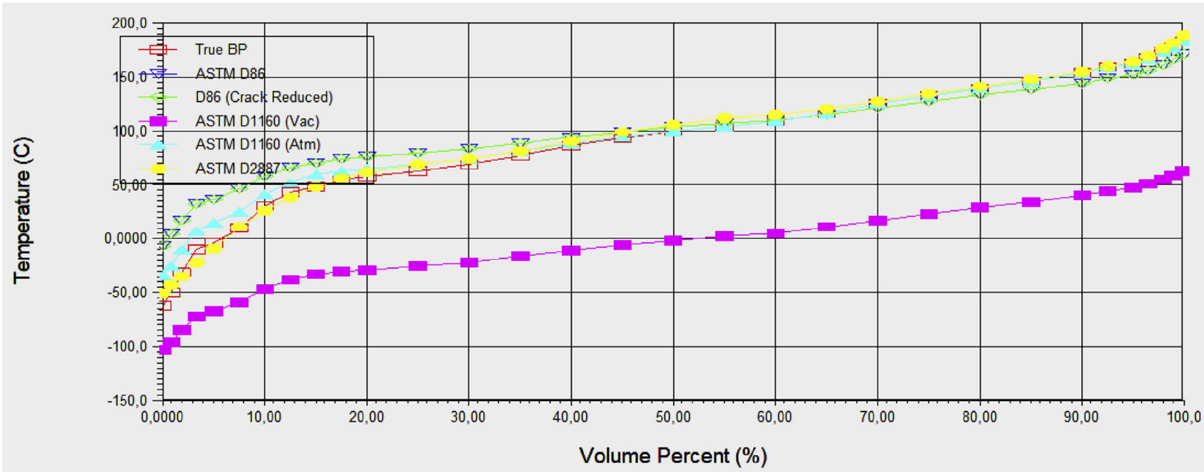

Fig. 5. The quality of gasoline from K-1 and K-2.

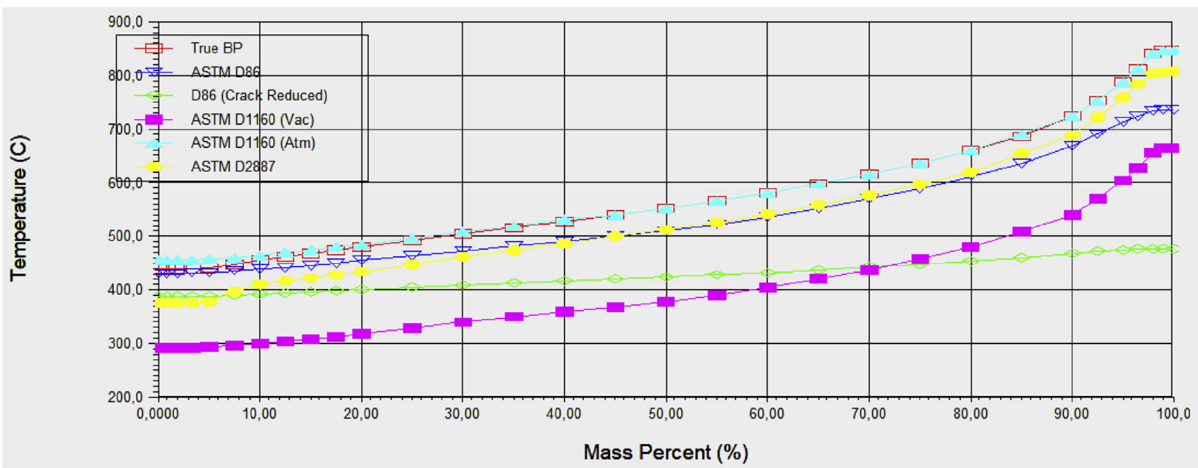

Fig. 6. The quality of Tar from K-5.

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