Heat Integration Improvement for Benzene Hydrocarbons Extraction from Coke-Oven Gas

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In present work the extraction of benzene hydrocarbons from coke-oven gas on the one column unit is studied. Energy consumption of the investigated plant is defined and disadvantages of existing heat exchangers network are shown. The integrated flexible flowsheet is proposed based on process integration method. It enables to reduce both hot and cold utilities consumption on 2,872 kW with pay back period about 9 months.

1. Introduction

Ukrainian industrial sector consumes 8-12 times more energy for GDP product unit than in developed countries. Japan spends 370 g of equivalent fuel to produce 1 $ of gross domestic product, the same rate for USA is 600 g and for Ukraine is 4.7 kg. It shows that the improvement of energy efficiency is the main objective for Ukrainian economy. Now in Ukraine there are 14 coke oven plants in operation. They are part of metallurgical industry, which is one of the main branches of Ukrainian economy. Besides, the ferrous metallurgy is biggest consumer of energy in Ukrainian industrial sector. In this paper the extraction of benzene hydrocarbons from coke-oven gas on the one column unit is studied by process integration technique.

2. Process description

Extraction of benzene hydrocarbons is the part of complex purification process of coke-oven gas. Coke-oven gas is delivered from desulfurization plant to scrubber where benzene hydrocarbons are extracted by stripping oil. Refined coke-oven gas goes from the top of the scrubber to storages and to power plant. Stripping oil is drawn from the bottom of the scrubber and is fed to heat exchangers. Oil goes through two recuperative heat exchangers and one utility heat exchanger sequentially. Recuperative heat exchangers are heat by top and bottom products of stripping column and utility heat exchanger is heat by steam. Heated stripping oil is delivered to stripping column. Direct steam is also fed to the column. Vapours of benzene hydrocarbons mixture leave the column from the top and is cooled in the recuperative heat exchanger and after in the cooler. Regenerated stripping oil is taken away from the bottom of column and is cooled in the recuperative heat exchanger and in the coolers. Cooled stripping oil is fed to the scrubber again. Principal flowsheet of benzene hydrocarbons extraction plant is shown on Figure 1.

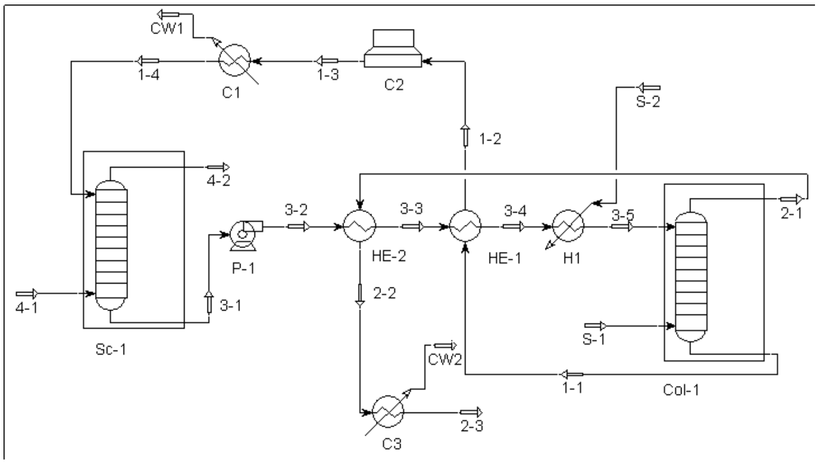

Figure 1: Principal flowsheet of benzene hydrocarbons extraction plant. Sc – scrubber; Col – stripping column; P-1 – pump; HE-1, HE-2 – recuperative heat exchangers; H1 – steam heater; C1, C3 – water cooler; C2 – air cooler; 1-1 – 1-4 – regenerated stripping oil; 2-1 – 2-3 – benzene hydrocarbons mixture; 3-1 – 3-5 – stripping oil from the scrubber; 4-1 – 4-2 – coke-oven gas; CW1, CW2 – cooling water; S-1, S-2 – steam.

Stream data were analyzed and process streams were selected for heat integration improvement. Two hot and one cold streams were identified and included to heat integration. Thermophysical properties and stream data are collected in Table 1.

Table 1. Stream data.

№	Stream	Type	TS, °C	TT, °C	W, kg/s	CP, kW/°C	dH, kW	α, kW / (m^2·°C)
1	Refrigerated stripping oil	hot	139	30	71.67	109.83	11971	0.91
2.1	Benzene hydrocarbons condensation	hot	90	90	2.36	r[1]=1797 kJ/kg	4243	8.14
2.2	Benzene hydrocarbons vapor cooling	hot	139	90	2.36	5.14	252	0.50
2.3	Benzene hydrocarbons liquid cooling	hot	90	30	2.36	6.62	397	1.07
3	Stripping oil from the scrubber	cold	30	140	74.03	120.37	13240	0.91

[1] – phase transition heat, kJ/kg.

3. Heat integration

3.1 Analysis

Composite curves and grid diagram are the main pinch analysis tools used in present work. They sufficiently well described in literature by Linnhoff et al. (1982), Smith (2005), Kemp (2007), Klemes et al. (2010) and others. The use of pinch method for existing industrial process usually have some problems which concern to properties of components, old equipment type, process automation, process limitations and others. The first step of this case study is the analysis of energy consumption of existing heat exchangers network for benzene hydrocarbons extraction. For existing heat exchanger network the ΔT_{min} = 30 °C. It gives the possibility to build the composite curves and to define the target values of hot and cold utilities. Composite curves for ΔT_{min} = 30 °C are shown on Figure 2, cold utility consumption is 7,617 kW, and of hot utility is 3,998 kW.

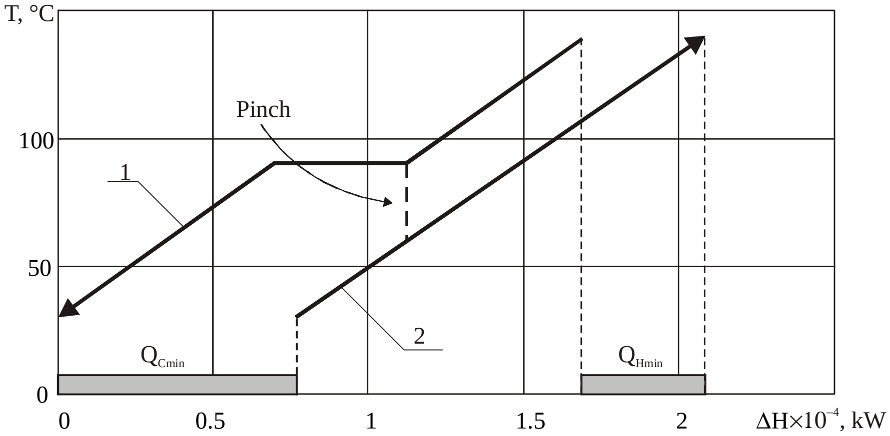

Figure 2: Composite curves for ΔT_{min}=30°C. 1 – hot composite curve; 2 – cold composite curve; Q_{Hmin} – hot utility consumption; Q_{Cmin} – cold utility consumption; T – temperature; ΔH – enthalpy difference.

This amount of utilities differs from that in a real process. Hot utility consumption of existing process is 4,462 kW and cold utility is 8,085 kW. It comes out of pinch crossing and use of cold utility above the pinch point. These disadvantages of existing heat exchangers network are well illustrated on grid diagram (Figure 3). The similar trend is typical for all Ukrainian industrial processes, which were designed for low prices of utilities. Now these plants cannot produce the competitive product. Selection of optimal design of heat exchangers network lets to reduce utility consumption and pollution by flue gases, as was shown by Varbanov et el. (2005) and later by Klemes et el. (2007). The selection of best design for heat exchangers network is the cost optimization problem, as shown by Linnhoff (1990).

3.2 Synthesis

The function of total reduced cost of heat exchangers network was used for definition of optimal ΔT_{min} for heat exchangers network. It typical function used in the pinch analysis

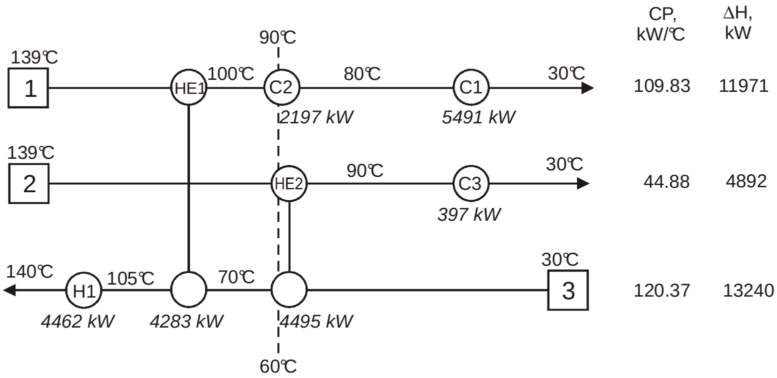

Figure 3: Grid diagram of existing process. HE1, HE2 – recuperative heat exchangers; C1 – C3 – coolers; H1 – heater; CP – stream heat capacity; ΔH – enthalpy difference.

and well described by Linnhoff and Ahmad (1990). The following cost parameters are used. Cost of installation of one heat exchanger or repiping of existing one equal to 10,000 USD. Cost of 1 m^2 of heat transfer area is 500 m^2. Nonlinear cost factor of heat transfer is 0.87. Cost of hot utility is 200 USD/kWy; cost of cold utility 20 USD/kWy. The interest rate assumed 10% and payback time 5 years. The ΔT_{min} is varied from 1 to 80 °C. Cost curves presented on Figure 4 show that minimum of the objective function under variable is 10 °C. This value of ΔT_{min} is optimal for heat exchangers network.

Composite curves built for optimal ΔT_{min} = 10 °C show the target values of utilities consumption. Hot utility target value is 1,590 kW and cold utility target is 5,210 kW. Pinch point localization is on the temperatures 90 °C and 80 °C respectively. These curves for optimal ΔT_{min} are presented on Figure 5.

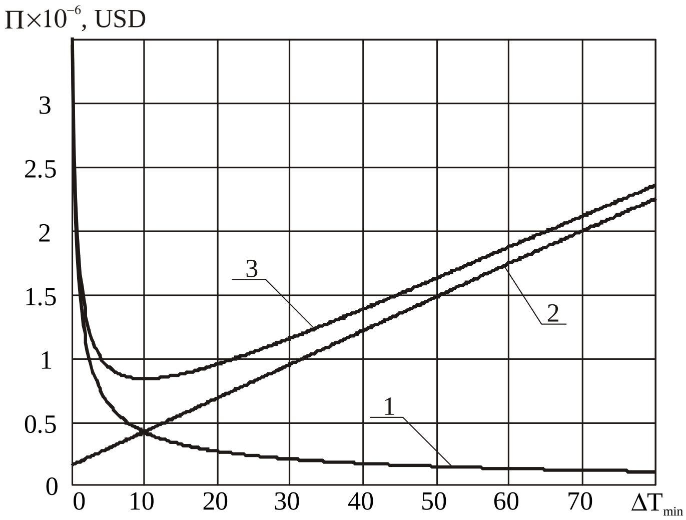

Figure 4: Cost curves: 1 – reduced capital costs; 2 – operation costs; 3 – reduced total costs; ∏ - costs, USD; ΔT_{min} – minimal temperature difference on heat exchangers, °C.

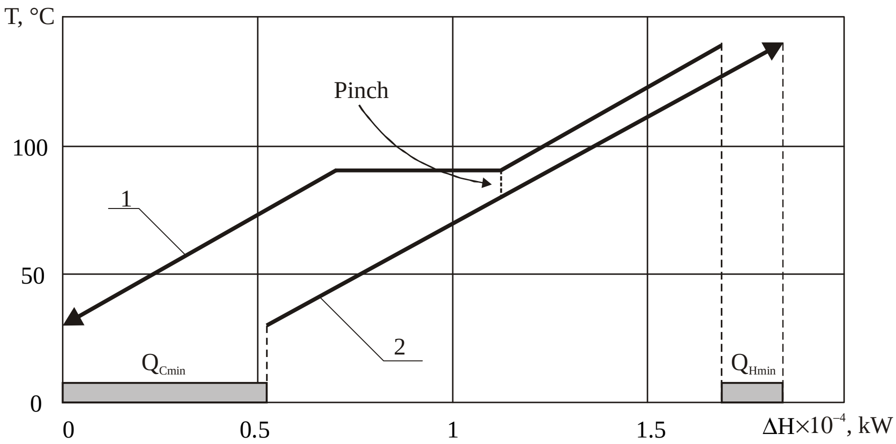

Figure 5: Composite curves for optimal ΔT_{min}=10 °C. 1 – hot composite curve; 2 – cold composite curve; Q_{Hmin} – hot utility consumption; Q_{Cmin} – cold utility consumption; T – temperature; ΔH – enthalpy difference.

3.3 Retrofit design

Composite curves give all information for heat exchangers network design. Grid diagram tool is used and retrofit project presented on Figue 6. There are 4 recuperative heat exchangers, 2 coolers and 1 heater. We can see some loops on this diagram but to break them is not expediently because of big duties of heat exchangers and different phase state of stream 2 (see Table 1 and Figure 6).

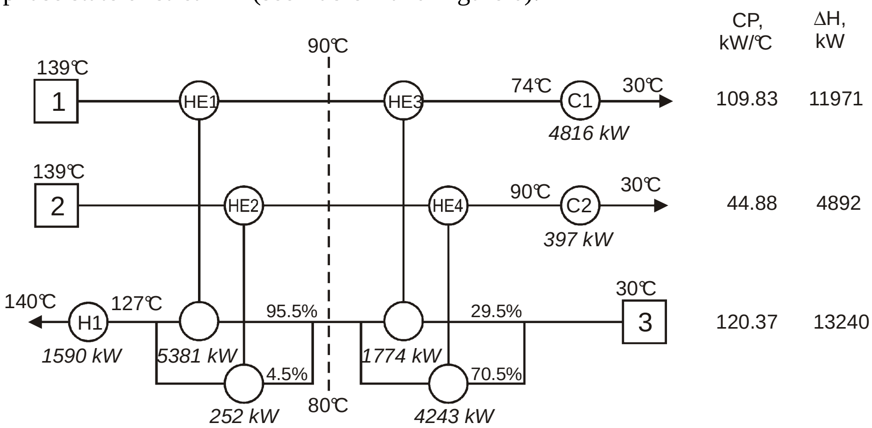

Figure 6: Grid diagram of integrated process. HE1 – HE4 – recuperative heat exchangers; C1, C2 – coolers; H1 – heater; CP stream heat capacity; ΔH – enthalpy difference.

4. Results and discussion

The optimal design of heat exchangers network for benzene hydrocarbons extraction lets to reduce both hot and cold utilities consumption on 2,872 kW. It saves the 631,840

USD/y. Reduced operating cost for optimal ΔT_{min} = 10 °C is 422,220 USD, capital cost for heat exchangers network realization is 412,930 USD and total reduced cost is 835,150 USD. The simple payback period is 8.5 month.

5. Conclusion

This case study shows the big energy saving potential for benzene hydrocarbons extraction from coke-oven gas. But benzene hydrocarbons extraction is a small part of coke-oven plant. The observation of whole site lets to achieve much better results than retrofit of one unit. Decreasing of utility consumption of coke oven plant is reducing the coke-oven gas usage. Reduction of big amount of coke-oven gas enables to use it for power generation. The last point can help the Ukrainian economy to save considerable funds and to reduce CO_2 and other harmful gaseous emissions.

6. Acknowledgements

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