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Integration Processes of Benzene-toluene-xylene Fractionation, Hydrogenation, Hydrodesulphurization and Hydrothermoprocessing on Installation of Benzene Unit

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The heat exchanger network (HEN) of the unit of benzene production at petrochemical plant was inspected and the obtained data were analyzed for possible plant retrofit targeting the minimal energy consumption and increasing of plant efficiency. The benzene-toluene-xylene fractionation, hydrogenation, hydrodesulphurization and hydrothermo processing units of the plant were analysed and the data of existing HEN flowsheets were extracted. The minimum temperature difference for the retrofit was determined by analyzing the cost parameters of the proposed modifications as well as the energy cost. Using Pinch Analysis methodology the Composite Curves and Grid Diagram were obtained for the integrated processes of these plant units. The new flowsheet of the HEN of the regarded units was developed; and the possible application of heat transfer equipment was analyzed and proposed.

1. Introduction

Refinery processes are one of the key components of modern industry, but at the same time they are the most power-consuming. The global rising world demand in energy for oil and petrochemical industries requires the energy efficient plant designs with the retrofitting of the existing HENs and application of effective equipment. The main part of the technological units in the countries formed after collapse of the Soviet Union was put into operation in 50ies – 70ies of the last century and more than 80 % of the equipment and flowsheets are now ineffective and consume a lot of energy as well as produce a lot of hazardous emissions. This led to the fact that the average crude oil processing yield of the oil refining plant now for some enterprises is about 70 %, when in Europe and US this parameter is usually above 90 %. The industry efficiency should be increased targeting less energy consumption and high process efficiency, what assumes its modernization and technological upgrading (Stepanov et al., 1989). The retrofit of petrochemical enterprise requires a lot of investment, high level of operating and installation costs and aimed to provide the acceptable level of crude oil processing yield by means of known techniques. The use of the methods of Pinch Analysis allows determining the rational opportunities for energy and cost minimization by improving the use of thermal energy (Kemp, 2007). The retrofit of HEN usually is performed in two steps: the first is to identify the Pinch point and to adjust properly the heat transfer loads between the heat exchangers to maximize the heat recovery between processes; the second is to propose the structural modification of the HEN focusing Pinch, which include the heat exchangers application and their installation, re-piping of the network. This approach provides various retrofit options with different degree of heat recovery, energy saving and additional capital investment (Klemeš, 2013). The petrochemical plants usually contain several process units connected to each other and to the common utility system. In the present work the enterprise OOO “Sibur-Kstovo” (Nizhny Novgorod, Russian Federation) is under consideration and its four interconnected units (benzene-toluene-xylene fractionation, hydrogenation, hydrodesulphurization and hydrothermoprocessing) involved in benzene production are analysed using the Pinch Integration.

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2. The process description

The present work observes four inter-connected units of petrochemical plant: benzene-toluene-xylene fractionation unit, hydrogenation unit, hydrodesulphurization unit and hydrothermoprocessing unit. Using Process Integration it is possible to integrate their HEN to the common network to provide the less energy consumption of these units. The unit of pyrocondensate processing consists of the following process blocks (Ulyev et al., 2014): (1) Extraction of benzene-toluene-xylene fraction from pyrocondensate in column K-301; (2) Hydrogenation, hydrodesulphurization and hydrothermoprocessing in the reactors R-301, R-302, R-303/1 R-303/2; (3) Stabilization and separation of hydrodealkylate in columns K-305, K-306, K-307 with separation of commercial benzene; (4) Compression of hydrogen-containing gas; (5) Cleaning, drying of hydrogen-containing gas and concentration of hydrogen; (6) Compressing the low pressure methane M-303. The flowsheet of the existing processes is demonstrated in Figure 1.

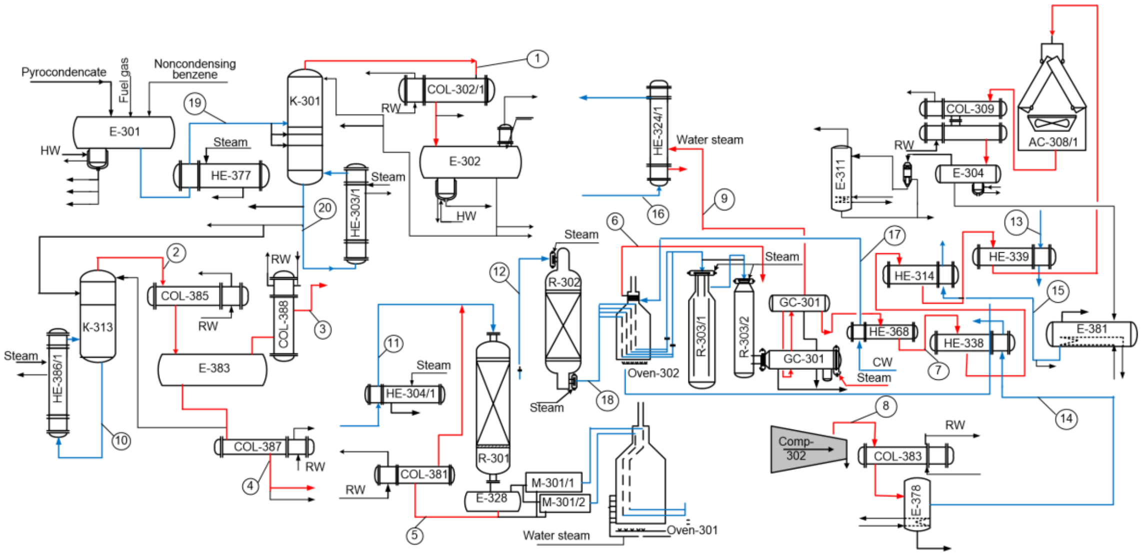

Figure 1: The flowchart of the existing process (COMP – compressors; R – reactors; K – distillation columns; M – mixers; HE – heat exchangers; GC – gas cooler; COL – coolers; AC – air coolers; RW - recycled water; HW – hot water; CW – cold water)

The raw material for the benzene production is a benzene-toluene xylene (BTX) fraction, which is allocated from pyrocondensate in a distillation column K-301. The rectification of pyrocondensate in column K-301 is performed under vacuum in order to maintain the temperature of cube lower than 180 °C to avoid the polymerization of unsaturated hydrocarbons. Heat transfer in the extraction column of benzene-toluene-xylene fraction is carried out by circulating the bottoms liquid of column K-301. The benzene-toluene-xylene fraction is fed on the first stage of hydrogenation. The bottom product of column K-301 is fed to the column K-313. The rectification in column K-313 is performed under vacuum in order to maintain the temperature of cube lower than 170 °C to avoid polymerization of unsaturated hydrocarbons. The heat transfer in the column K-313 is provided by circulating of the column bottoms liquid. The benzene-toluene-xylene fraction after the coagulator is mixed with a hydrogen-consisted gas from the compressor M-302 and passes through the heater T-304 to the hydrogenation reactor of benzene-toluene-xylene fraction of the P-301 1st stage. Hydrogenize from the 1st stage of the reactor R-301 bottom comes to the separator. The hydrogenized liquid from the separator is fed to the M-301/1,2 mixer. A mixture of hydrogenize from the 1st stage and hydrogen-consisted gas from the M-301/1,2 mixer are heated in the P-301 furnace, and enters the 2nd stage of the R-302 hydrogenation reactor. The mixture of hydrogen-consisted gas with the aromatic recycle and hydrogenize from 2nd stage is heated in P-302 oven and enters the R-303/1 reactor. The stream from the top of R-303/1 reactor enters the R-303/2 reactor. The hydrodealkylation from R-303/2 reactor enters the GC-301 gas cooler and then comes to T-368. Hydrodealkylation from T-368 heat-exchanger sequentially passes through T-338 heat-exchangers, where it heats the hydrogen fraction coming to P-301; T-314, where it heats the feed stream of stabilizer column K-305; T-339, where it heats hydrogen-consisted gas from the compressor Comp-301 for the unit of hydrocracking and hydrodealkylation. Hydrodealkylate from the heat-exchanger T-339 is cooled in coolers T-308 and T-309, and enters to the separator. Hydrogen fraction from compressor Comp-302 passes the water cooler T-383 and enters to the separator; then it comes to the hydrogenation reactors R-301, R-302.

3. Data extraction process

In the review of technological processes, the basic parameters of technological streams included in the integration process have been determined and the stream table of the process was obtained (Table 1) according to methodology described in book by Klemeš (2013).

Table 1: The process streams data

№	Type	TS, °C	TT, °C	G, kg/h	C, kJ/(kg·°C)	CP, kW/°C	r, kJ/kg	α, kW/m^2·K	ΔH, kW
1	hot	67.1	27	23,840	1.70	11.23		0.764	450.11
2	hot	126	96	2,509	6.93	4.83		0.764	144.47
3	hot	96	30	580	2.70	0.44		0.764	28.71
4	hot	90	30	1,109	2.39	0.74		0.142	44.29
5	hot	146	41	2,769	2.34	1.80		0.714	189.11
6	hot	680	163	11,024	2.30	7.04		0.467	3,641.29
7	hot	220	35	23,120	2.12	13.61		0.544	2,517.01
8	hot	65	25	21,860	11.09	67.36		0.484	2,694.60
9	hot	215	215	2,554			1,924	0.818	1,364.97
10	cold	133	145	37,419	2.56	26.56		0.173	324.00
11	cold	20	41	91,750	4.71	119.91		0.818	2,506.16
12	cold	130	278	15,388	2.14	9.16		0.818	1,355.07
13	cold	52	104	12,100	11.22	37.70		0.500	1,960.13
14	cold	32	62.2	21,860	2.73	16.55		0.375	499.30
15	cold	16	46.8	8,963	3.59	8.93		0.500	274.82
16	cold	192	211	93,781	2.80	72.99		0.818	1,364.97
17	cold	142	161.6	26,300	4.32	31.58		0.714	619.03
18	cold	270	565	15,014	2.91	12.11		0.818	3,573.45
19	cold	126	135	30,590	2.06	17.51		0.173	157.58
20	cold	146	159	90,950	2.29	57.85		0.173	752.11

The following flows were under consideration: 1 - Vapours of BTX-fraction from the K-301; 2 - Vapours of C9 fraction from the K-301; 3 - The gas phase of hydrocarbons from E-383; 4 - C9 fraction; 5 - Recycle hydrogenize; 6 - Cooling gas from the P-302; 7 – Hydrodealkylate; 8 - Hydrogen-containing gas from M-302; 9 - Vapour from gas cooler; 10 - Bottoms liquid from K-313; 11 - Hydrogen-containing gas to P-301; 12 – Hydrogenize; 13 - Hydrogen-containing gas from E-312; 14 - Hydrogen-containing gas from E-378; 15 - Hydrodealkylate to K-305; 16 - Bottoms liquid from K-306; 17 - Cold cooling water from the pyrolysis department; 18 - Hydrogenize of the 2nd stage; 19 – Pyrocondensate; 20 - Bottoms liquid from K-301.

Basing on the obtained data a Grid Diagram (Linnhoff and Flower, 1978) for the existing process was obtained. In Figure 2 the Grid Diagram, which demonstrates the process streams and existing heat exchangers connections between them is presented. The analyses of the obtained data showed that the heat consumption of the existing process for cold utilities (Q_c) comes to 8,146 kW and 10,494 kW for hot utilities (Q_H), the heat recovery (Q_{REC}) equals to 2,930.90 kW if to consider all the heat exchangers involved in the flowchart of the processes for all four units. Such high power consumption needs Process Integration and more energy recuperation between the process streams.

4. The Heat Integration

The Process Integration of the HEN requires proper determination of energy and capital targets. The main parameters, which influence the HEN capital cost includes number of heat exchangers, their heat exchanger area, the type of heat exchangers and pressure drop of heat carriers, the installed capital cost of heat exchanger and re-piping. The proper energy and cost targeting needs correct estimation of the minimum temperature difference (ΔT_{min}) between heat source and sink streams (Nordman, 2005). The annualized capital cost plotted together with annual utility cost for the range of minimum temperature difference, what corresponds to the different process modifications, provide the optimum energy consumption and indicate the corresponding ΔT_{min} (Smith, 2005).

The capital cost of the heat exchanger is defined by Eq(1):

$$\text{Capital cost} = A_T + B_T (S)^c \tag{1}$$

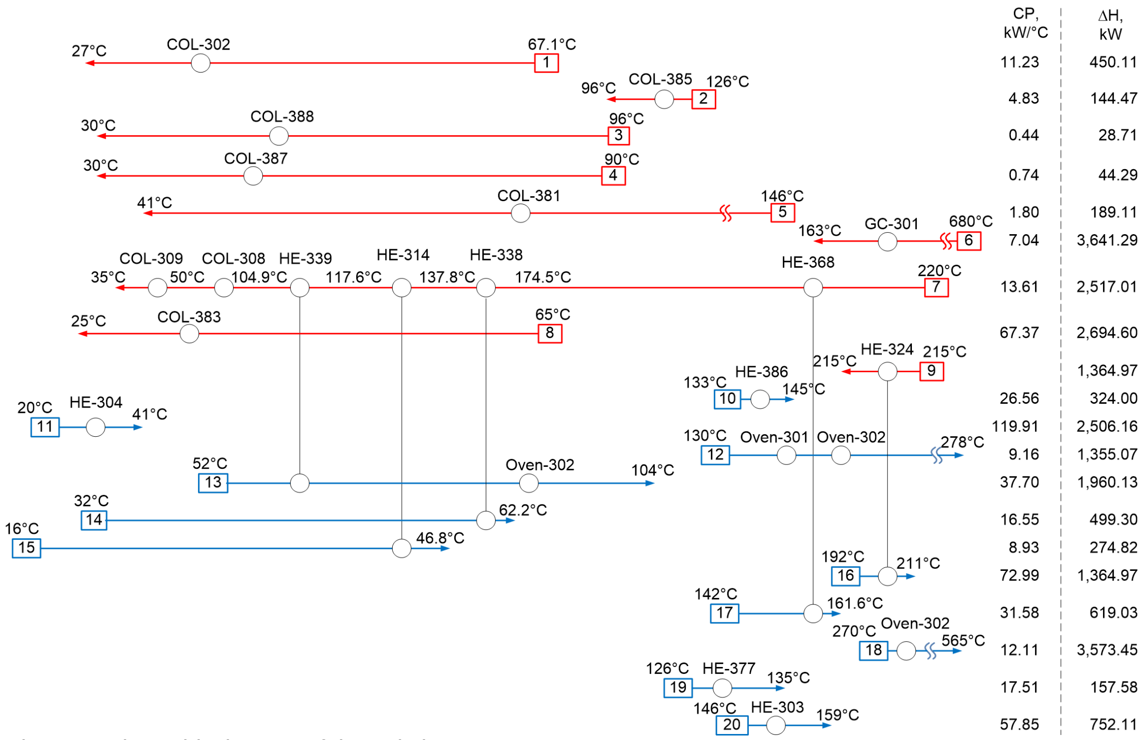

Figure 2: The Grid Diagram of the existing process

where: A_T is the cost of installing of shell-and-tube heat exchanger, A_T = 10,000 USD; B_T is the rate equivalent to the cost of 1 m² heat transfer surface area, and for shell-and-tube heat exchangers, B_T = 4,000 USD; S is the heat transfer surface area, m²; C is the factor reflecting the non-linear dependence on the value of the cost of the heat exchanger to the heat transfer surface, and was taken C = 0.87. The selected prices of hot and cold utilities are not stable and affect the total cost target. In the present work the cost of hot utilities used in the process are taken equal to 316 USD/(kW·y) assuming 8,000 h/y activity, and for the cold utilities it was set to 10 USD/(kW·y).

Using «PINCH 2.0» (Tovazhnyansky et al., 2003) software the Cost Curves were obtained and the optimum minimum temperature difference value was determined. For the observed process ΔT_{min} is equal to 16 °C. Basing on the targeted temperature difference (ΔT_{min} = 16 °C) the Composite Curves for source and sink streams were built (Figure 3).

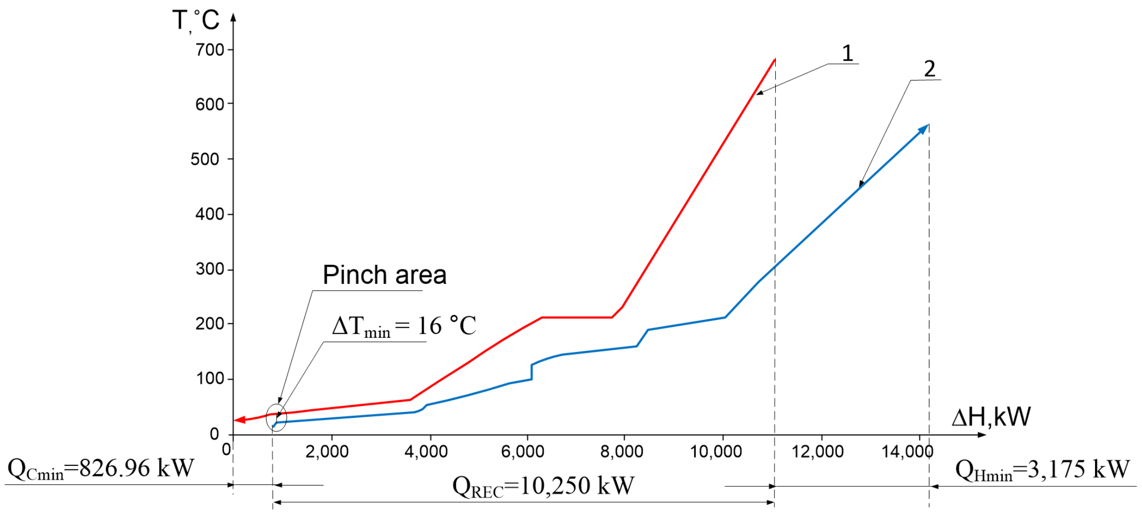

Figure 3: The Composite Curves of the integrated process for ΔT_{min} = 16 °C: 1 – Hot Composite Curve, 2 – Cold Composite Curve

The proposed retrofit project has the recuperation load of Q_{REC} = 10,250 kW, the minimum cold utilities are Q_{Cmin} = 826.96 kW, and minimum hot utilities are Q_{Hmin} = 3,175 kW, what comparing with the existing process shows the significant increase of the recuperation heat and decrease of the heat loads.
For the flowsheet retrofit according to the achieved targeted loads the HEN structural changes were carried out for several possible solutions according to the principles of Pinch Analysis (Smith et al., 2000).

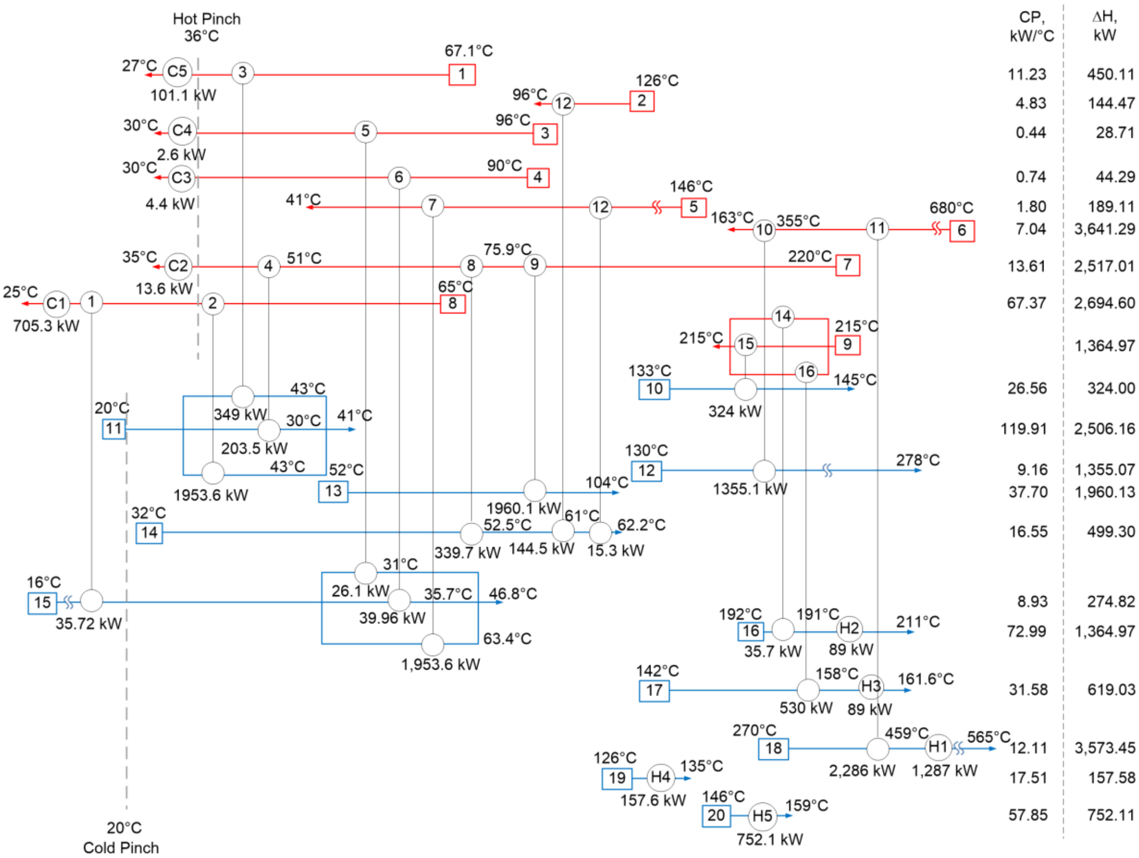

Figure 4: The Grid Diagram of the integrated processes, ΔT_{min} = 16 °C

Table 2: The existing and additional heat transfer area for the retrofitted process

	T_{HOTin}, °C	T_{HOTout}, °C	T_{COLDin}, °C	$T_{COLDout}$, °C	ΔT_{LM}, °C	Q, kW	S, m^2	Existing HE	Additional surface, m^2
HE1	36	35.47	16	20	17.68	35.72	11	T-388	32
HE2	65	36	20	43	18.84	1,953.59	432	T-324, T-338	-
HE3	67.1	36	20	43.27	19.66	349.10	57	T-304/1	17
HE4	50.95	36	20	30.17	18.29	203.48	44	-	44
HE5	96	36	20	31	26.09	26.10	4	-	4
HE6	90	36	20	35.79	31.31	39.96	15	-	15
HE7	137.35	41	20	63.36	42.08	173.86	18	-	18
HE8	75.91	50.95	32	52.53	21.09	339.71	92	T-383	2
HE9	220	75.91	52	104	58.31	1,960.13	164	T-368	55
HE10	355.25	163	130	278	52.03	1,355.07	111	T-309	-
HE11	680	355.25	270	458.79	142.59	2,286.22	69	T-385	10
HE12	126	96	52.53	61.26	53.40	144.47	14	T-377	-
HE13	146	137.35	61,26	62.2	79.88	15.25	1	-	1
HE14	215	215	192	199	19.29	510.95	75	T-386	35
HE15	215	215	133	145	75.84	324	37	-	37
HE16	215	215	142	158.79	64.24	530.02	26	-	26

The most energy and cost effective alternative was obtained and its Grid Diagram is presented in Figure 4. It demonstrated the integration of four units of benzene production plant, which are benzene-toluene-xylene fractionation unit, hydrogenation unit, hydrodesulphurization unit and hydrothermoprocessing unit, with the ΔT_{min} = 16 °C, Hot Pinch is 36 °C and Cold Pinch is 20 °C. The proposed retrofit needs 5 coolers and 5 heaters to provide the heat supply. For this process 15 heat exchangers are required. As all the used heat exchangers are of shell-and-tube type, it was proposed as much as possible to use the existing heat transfer capacities. The additional surface area for the existing heat exchangers is listed in Table 2. The heat transfer equipment for positions 4,5,6,7,13,15,16 requires purchasing the new units. For these positions in the present work the shell-and-tube units are considered, but also it is needed to analyze the possibility to use the compact heat exchangers, that will help to save the consuming energy, installation costs and material for production.

The obtained retrofit project needs totally 1,170 m^2 of heat transfer surface area, from which 296 m^2 is needed additionally to increase the existing heat transfer surface area. For the case the shell-and-tube units were under consideration. The possibilities to use compact heat exchangers for some positions and the application of the Shifted Retrofit Thermodynamic Diagram methodology (Yong et al., 2014) and an Extended Grid Diagram (Yong et al., 2015) can be implemented further for more heat recovery and higher utilities saving, what is going to be the subject for future work.

5. Conclusions

A Pinch Analysis and Pinch Integration methodologies were applied for the retrofit of four units of benzene production plant, namely benzene-toluene-xylene fractionation unit, hydrogenation unit, hydrodesulphurization unit and hydrothermoprocessing unit. The proposed retrofit allows reducing the power consumption by 7.3 MW. The hot utilities needed to the process operation are reduced by 69.94 % and cold utilities are 89.84 % less. The proposed retrofit of plant HEN applies the existing heat transfer equipment, and the additional heat transfer surface area comes to 296 m^2. The calculated annual income from the project implementation is 1.56×10^6 USD and the expected payback period is about 6 months.

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