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Abstract: Two approaches are applied for reducing energy consumption in an actual delayed coking plant converting vacuum residue into lighter and higher value hydrocarbon products. Pinch analysis is applied with a view to optimising the pinch temperature approach and structure of the existing seven heat exchangers' network. This resulted in about 27% reduction of the total annual cost corresponding to an annual saving of about US$1.9 million. Exergy analysis is also conducted on other main process plant equipment which revealed a possible annual saving of US$0.75 million through generation of medium pressure steam in the blowdown tower.

Keywords: delayed coker; pinch analysis; process optimisation; exergy analysis; energy efficiency.

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1   Introduction

Delayed coking is a reliable, well tested, and widely used deep conversion technology for processing heavy petroleum fractions to higher value lighter fractions in order to satisfy stringent market demands. This process is however energy intensive and is consequently associated with undesirable emissions. Improvement of energy efficiency in delayed coking units (DCU) targets both an increased profitability and a better environmental performance. Pinch analysis emerged as one of the most common tools for heat transfer task integration in process plants (Zhang et al., 1981). It enables the realisation of maximum heat recovery among process streams in a heat exchange network and therefore the optimisation of external heating and/or cooling requirements subject to the minimisation of the total annual cost of the heat transfer network. This includes both the annualised capital cost and the operating additional heating and/or cooling costs.

Earlier publication of the concept of pinch analysis (Ostrovsky et al., 1985) gave two four-stream examples to illustrate its potential and its possible industrial applicability both at the network design stage and in the retrofitting of existing networks. It appealed to a wide variety of chemical industries especially the oil and gas industry. Imperial Chemical Industries was among the first companies to adopt pinch analysis for the preheating section in a crude oil distillation unit (Linnhoff, 1993) with a view to increasing the refining capacity by 20%. The revamp did not only improve heat conservation in order to minimise the additional energy requirements associated with the increased capacity but it also avoided the installation of a new furnace for which no plant space was available in the layout. The modification of the existing heat exchangers' configuration enabled to achieve significant reduction in external energy use.

The application of pinch technology has been expanded through using a variety of analytical (Mehdizadeh Fard et al., 2017), graphical, and computational tools for

optimising energy recovery and therefore minimising utility consumption in process plants. Although pinch technology has been successfully applied to improve the thermal efficiency of heat exchange networks, it is not applicable to the analysis of the energy efficiency of other process equipment or the efficiency of configuring high pressure streams (Ghannadzadeh and Sadeqzadeh, 2017). This is because it is only concerned with the quantity of heat (enthalpy) rather than its quality (entropy). Accordingly, pinch analysis per se is not sufficient for conducting comprehensive energy efficiency studies. Improving integrated process thermal efficiency must also consider exergy analysis of the individual equipment since it enables the determination of the feasible energy recovery options under actual plant conditions. Exergy refers to the maximum useful work possible during a process that brings the system into equilibrium with a heat reservoir to reach maximum entropy.

Typical industrial applications of exergy analysis covered many industries including sulphuric acid plants, refrigeration units, air separation schemes, and alternative desalination process technologies (Shahzad et al., 2019). It has been applied not only to new designs of power plants but also in upgrading the energy efficiency of existing facilities (Kamate and Gangavati, 2009). In the renewable energy sector, exergy analysis has been applied to all types of solar heat pumps, solar energy storage, and in the production of biofuels (Feroskhan et al., 2021). Exergy analysis is not incompatible with pinch analysis. Combining pinch and exergy analyses has been first implemented in closed cycle gas turbines (Dhole and Zheng, 1995) before being applied to more complex process systems. Recently, exergy analysis has been also used in optimising the operation of heat exchanger networks (Gilani and Morosuk, 2022).

The petroleum and petrochemicals industries, being intensive energy consumers, have benefited from combined pinch and exergy analyses. It has been used for studying typical hydrocracking units, diesel hydrotreating units (Bandyopadhyay et al., 2019), an amine regeneration tower for sour gas sweetening (Ibrahim et al., 2022), and the Claus sulphur recovery unit in gas plants. Many studies focused on optimising the delayed coker unit's (DCU) energy consumption. Retrofitting of a DCU in a Chinese refinery achieved 37% savings in energy consumption (Chen et al., 2004). The feed preheating section in another DCU has been improved using Pro II simulation software (Yang et al., 2013). This involved application of the traditional pinch analysis together with the calculation of the exergy destruction for the preheat section. This suggested the increase of the preheated feed temperature by more than 40°C which is currently applied in modern DCU preheat section designs.

The present work presents a case study integrating pinch and exergy analyses with a view to improving the energy efficiency of a recently operated DCU in an oil refinery plant in Egypt.

2   Process description

The capacity of the studied DCU is 25,000 BPSD. It converts heavy vacuum residue into higher-value light hydrocarbon liquids and petroleum coke through thermal cracking. The liquids and gases produced by the thermal cracking reactions are subsequently used as feedstock for downstream units for further processing and refining.

Figure 1   DCU simplified flow diagram
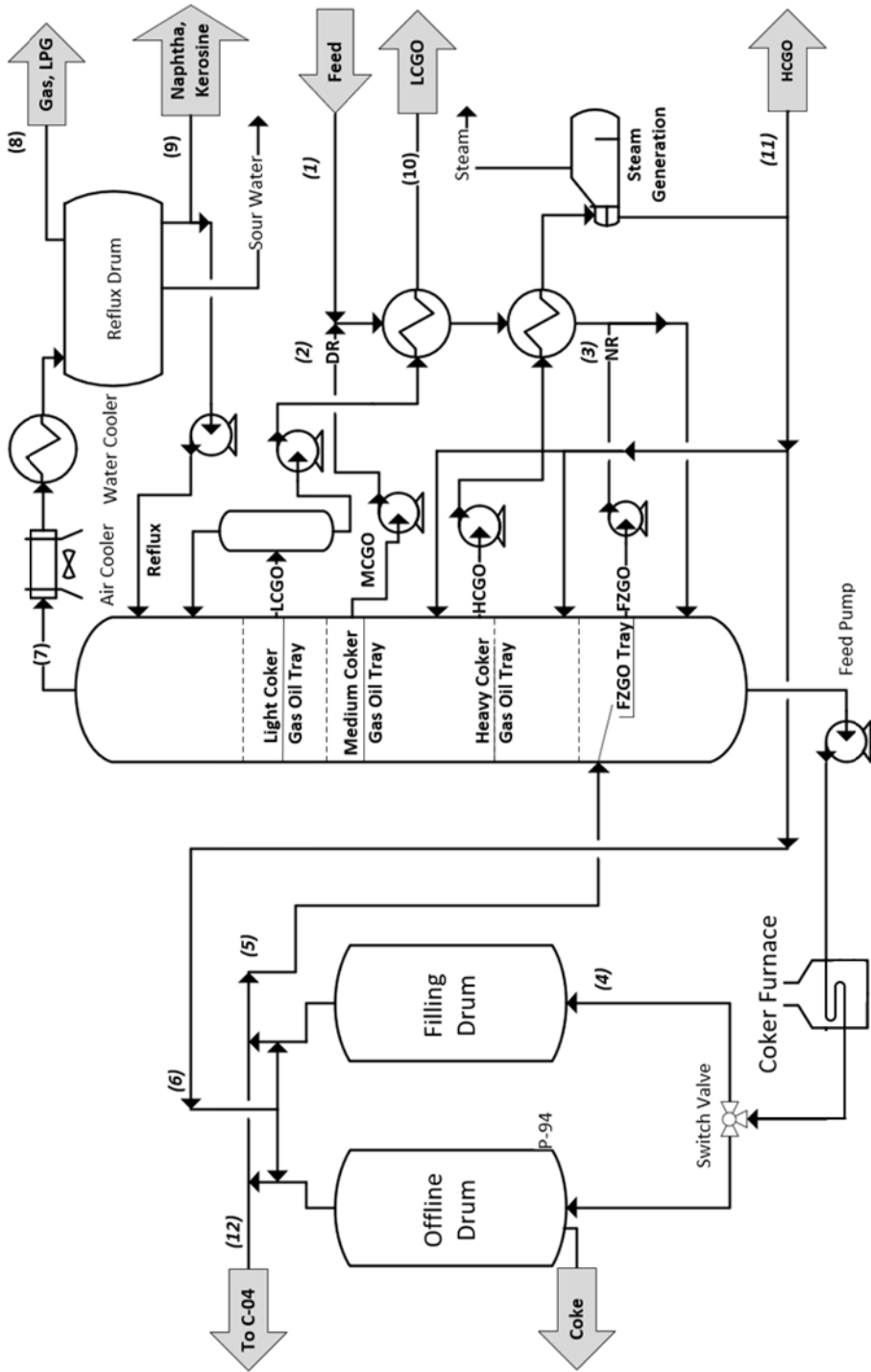
A simplified flow diagram of the DCU is given in Figure 1. Stream (1) feeding the unit is 80% hot feed, which comes directly from the vacuum distillation unit at 200°C and 20% cold feed from storage at 170°C. The total feed is combined with the distillate recycle, DR (2), which is medium coker gas oil, MCGO from the coker fractionator. The combined feed is preheated and then combined with the gas oil recycle (natural recycle, NR) (3), before entering the bottom of the coker fractionator.

The bottom section of the coker fractionator provides adequate surge capacity to maintain uninterrupted flow to the furnace charge pumps. The combined stream including fresh feed (1), distillate recycle, DR (2), and natural recycle, NR (3) is fed to the coker furnace and heated to the thermal cracking temperature of 496°C. The furnace effluent (4) flows into the bottom of an initially empty coke drum (online), while its paired drum is offline for coke removal and preparation for returning to coking service. Inside the online coke drum, the oil feed is converted into cracked hydrocarbon vapours and petroleum coke. The coke drum effluent vapour (5) is quenched with heavy oil (6), to terminate the coking reaction.

The coke drum vapours (5) are then directed to the main fractionation section. Lighter components are rectified to obtain the desired intermediate products. Two final products are obtained from the fractionation section via two side strippers namely heavy coker gas oil, HCGO (stripper not shown), which is sent to the hydrocracker (11) to crack it into lighter products. The second product is light coker gas oil, LCGO, which is sent to the diesel hydrotreating unit (10).

The overhead vapour from the fractionator (7) is routed to the gas recovery plant to separate the fractionator section wet gas and un-stabilised oil into main products. The first product is sweetened fuel gas (8), which is separated from the sponge oil absorber and treated in an amine treatment tower C-13 in the gas plant. It is then directed to the fuel gas unit outside the DCU. The second product is LPG (8), which is separated from the top product of the debutaniser tower in the gas plant. It is then directed to a naphtha hydrotreating unit outside the DCU.

The third product is naphtha and is separated from the top of the rerun column in the gas plant. It is directed to the naphtha hydrotreating unit. The last product from the gas plant is kerosene (heavy naphtha), which is separated from the bottom of the rerun column in the gas plant. It is directed to the diesel hydrotreating unit. Naphtha rerun tower C-14 is only used for adjusting the final boiling point of heavy naphtha (9) in case it is higher than 170°C. Usually, this tower does not exist in all DCUs, as other units rely only on manipulating the naphtha final boiling point by adjusting the temperature of the top of the main fractionator C-01. Thus, naphtha becomes on spec from debutaniser tower directly and there is no kerosene produced (as it is converted to LCGO from the main fractionator). In the current case, there is a rerun tower that helps to bring naphtha products on spec earlier, soon after startup. This tower separates naphtha, which represents 92% of the feed from the top and the slight heavy portion from the bottom.

The two coke drums operate intermittently (online and offline), quenching and decoking take place in the offline drum before returning to the online mode. The blowdown tower C-04, shown in Figure 2, is used to separate the offline coke drum vapours during steaming out and quenching in a 16-hour cycle, while the online drum is connected to the main fractionator. During the quenching step, offline coke drum vapours (12) leave the drum at 427°C. These vapours are mainly steam with small amounts of stripped hydrocarbons. High temperature vapours are mixed with cold water (13) and

combined (14) before entering the tower to reduce the temperature from 427°C to 200°C. They are then combined with heavy slop oil, HSO (15) to reduce the temperature of the tower inlet stream (16) to around 188°C.

Figure 2   Blowdown tower C-04
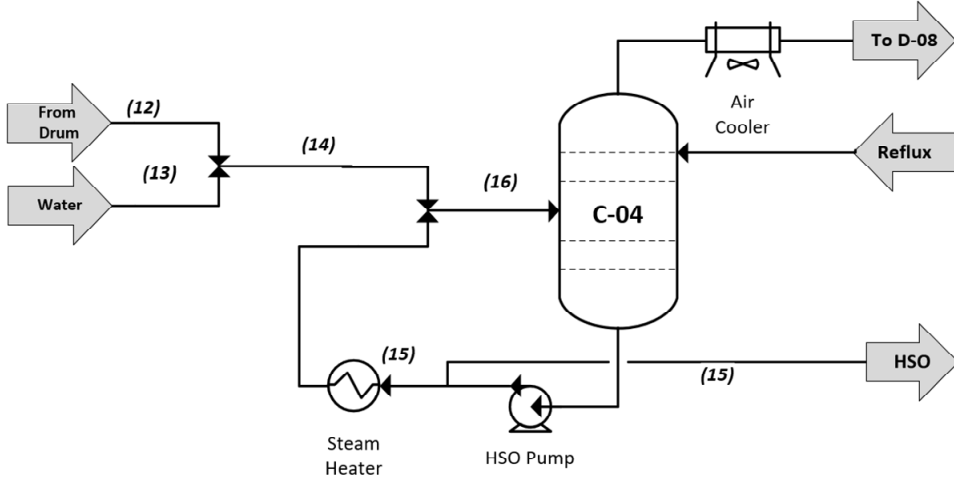
3   Methodology

Pinch analysis is used to optimise the HEN while exergy analysis is applied to the other equipment in the DCU. Building a simulation model is the first step to identify all of the unit stream properties and equipment required for pinch and exergy analyses. Besides addressing the optimal thermal design configuration, some modifications are proposed for the existing DCU.

3.1   Simulation model

The model is built using Aspen HYSYS software V11 to validate all stream properties in the DCU. It is then calibrated based on plant design data to ensure reproduction of the DCU products' yields since the vacuum residue (1) is combined with the two recycle streams (2) and (3). The final calibrated model is the base case for further energy studies.

3.2   Pinch analysis

It is required to raise the temperature of the combined feed to the coking drum from 190°C up to 496°C. Heating is typically accomplished in two steps: the preheating train and the fired heater. The main cooling requirement in the DCU is the main fractionator pump around. The fractionator inlet (5) temperature at the bottom tray is 424°C and should drop down to around 127°C at the tower top.

In order to optimise the HEN design, the base case (simulated by Aspen HYSYS) is combined with the main fired heater in a new model in Aspen Energy Analyser Software, V11. The program's cost calculations are based on a 10% rate of return, a five-year plant

life, and 365.25 operating days for calculating the operating costs. The following steps are followed:

a   A range of minimum approach temperature ΔT_{min} from 5°C to 40°C is studied to determine heating and cooling requirements. The area and number of shells are also calculated. The capital and operating costs for each assumed ΔT_{min} are computed.

b   The total annualised cost TAC accounting for both the capital cost CC and operating cost OC for each assumed ΔT_{min} is calculated.

c   The ΔT_{min} equivalent to lowest TAC is selected for the design of the final HEN.

3.3   Exergy analysis

In the absence of nuclear, electrical, surface tension, and magnetic effects, exergy, EX is the summation of the physical exergy, EX_{Ph}, chemical exergy, EX_{Ch}, kinetic exergy, EX_{Kin} and the potential exergy, EX_{Pot} (Abdollahi-Demneh et al., 2011). Normally, kinetic and potential exergies have very low values compared to other types of exergy and thus can be ignored. Only chemical and physical exergies are taken into consideration. For a flowing stream, the physical exergy is calculated as (Dincer and Marc, 2020)

$$EX_{Ph} = (H - H_0) - T_0(S - S_0) \tag{1}$$

where H, S, and T are the enthalpy, entropy, and temperature. The subscript 0 represents the reference conditions at which the exergy is taken as zero. The chemical exergy is calculated from

$$EX_{Ch} = \sum X_i EX_{qi} + RT_0 \sum X_i \ln X_i{}^\circ {}_i \tag{2}$$

where X_i is the mole fraction, EX_{qi} is the standard chemical exergy (kJ/mole). For pure substances, the standard chemical exergy can be easily obtained from updated tables, R is the universal gas constant, T_0 is the standard normal temperature (298.15 K), and γ_i is the activity coefficient. The activity coefficient is taken as 1 since for petroleum fractions the deviation for chemical exergy is only about 0.005% (Rivero et al., 1999).

Equation (2) can be used independently if the composition is known. Unfortunately, for petroleum fractions this is not applicable, because the majority of the components used are hypothetical components rather than pure components. The Szargut model, modified by Rivero et al. (1999) is used to obtain the standard chemical exergy for the pseudo-components.

$$EX_{qi} = \sum Z_i EX_{qj} + NHV_i \beta_i \tag{3}$$

where Z_i is the mass fraction of water and metals (V, Ni, Fe) in the hypothetical component. Normally, there is no water in the hypothetical component as it is defined as an individual component, NHV_i is the net heating value of the hypothetical component, and β_i is the chemical exergy correction factor calculated from equation (4).

$$\beta_i = 1.0401 + 0.1728(Z_{H_2}/Z_C) + 0.0432(Z_{O_2}/Z_C) + 0.2169(Z_S/Z_C) \tag{4}$$

where $Z_{H_2}, Z_c, Z_{O_2}, Z_S$ are the mass fractions of H_2, C, O_2, and S respectively. The exergy associated with the work performed on or by the system, it is calculated from

$$EX_W = W \tag{5}$$

where W is the amount of work done. For the heat transfer process, the associated exergy, EX_Q is obtained from

$$EX_Q = Q\left(1-\left(T_0 / T\right)\right) \tag{6}$$

where Q is the amount of transferred heat. The exergy balance equation can be summarised as:

$$\sum EX_{in} + \sum \left(1 - T_0 / T\right)Q = \sum E_{out} + \sum W + E_{loss} \tag{7}$$

The total exergy in or out is the summation of the physical exergy, EX_{Ph} and chemical exergy, EX_{Ch}.

The standard chemical exergy of all hypothetical components in the model is obtained by applying equations (2), (3) and (4). Exergy efficiency is then calculated for the main equipment in the DCU.

4   Results and discussion

4.1   Simulation model and pinch analysis

The two fluid packages used in the simulation model are the Peng Robinson (all hypothetical and pure components are used except for the amine treatment tower), and acid gas chemical solvent (for light components and some of the hypothetical components in the light boiling range. It is applied for the amine treatment tower).

Table 1   Model calibration targets

Model calibration targets	
Drum outlet temperature, °C	446
Drum outlet pressure, kPa	204.3
Coke sulphur, %	5.43
Furnace residence time, sec	60
Advanced kinetic parameters modified	
Modified fraction of unconverted residue	0.02
Sulphur to hydrocarbon cracking ratio	1
Aromatics coking activity	100
Tuning parameters altered on simulation	
Crack to H_2S	38.71
SG parameters (350–650 °F)	0.02812
SG parameters (650+)	–0.003

Calibration of the coker model including the coke drum in which the thermal cracking reaction takes place has been undertaken in order to match the unit design data. Other kinetic parameters were modified to ensure the strength of the model's response to

changes in Conradson carbon residue (CCR) and the drum pressure. Table 1 shows the calibration targets, advanced kinetic parameters and tuning parameters.

Ten towers, including distillation towers, strippers, and absorbers are simulated in addition to a heat exchanger network, eight air coolers, a fired heater, 14 separators, 24 pumps, and a two-stage compressor. The simulation model was validated by comparing the results with the design conditions.

The data of the original HEN was used for pinch analysis calculations for a range of ΔT_{min} from 5°C to 40°. Aspen energy analyser was used to carry out the calculations for alternative configurations of the network structure. Two configurations (A and B) gave the lowest total annual cost, TAC at the optimum ΔT_{min}. Figure 3 indicates that this optimum ΔT_{min} is between 10°C and 12°C. Configuration B is selected as the final configuration because it not only has a lower TAC, but it also gave a fired heater inlet temperature compatible with delayed coking requirements. Table 2 compares the performance of configurations A and B with the original unit design. Configuration B achieved a reduction in the TAC by US$0.06/sec which is equivalent to US$1.893 million per year. Similar cost savings associated with a HEN retrofitting project has been obtained for a large capacity crude distillation unit in Nigeria where a saving of US$1.6 million per year has been achieved (Waheed and Oni, 2015).

Table 2   Comparison between configurations A and B

Design no.	Configuration A		Configuration B		Original design	
	Cost index	*% of target*	*Cost index*	*% of target*	*Cost index*	*% of target*
Heating, $/s	0.13	102.3	0.13	102.3	0.23	180.0
Cooling, $/s	−0.02	103.8	-0.02	103.8	−0.07	−186.9
Operating, $/s	0.11	103.3	0.11	103.3	0.16	145.9
Capital cost, $	6,443,356	133.7	5,637,595	116.9	7,148,077	148.3
Total, $/s	0.17	*111.4*	0.16	106.9	0.22	*146.5*

Figure 3   TAC vs. ΔT_{min} (see online version for colours)
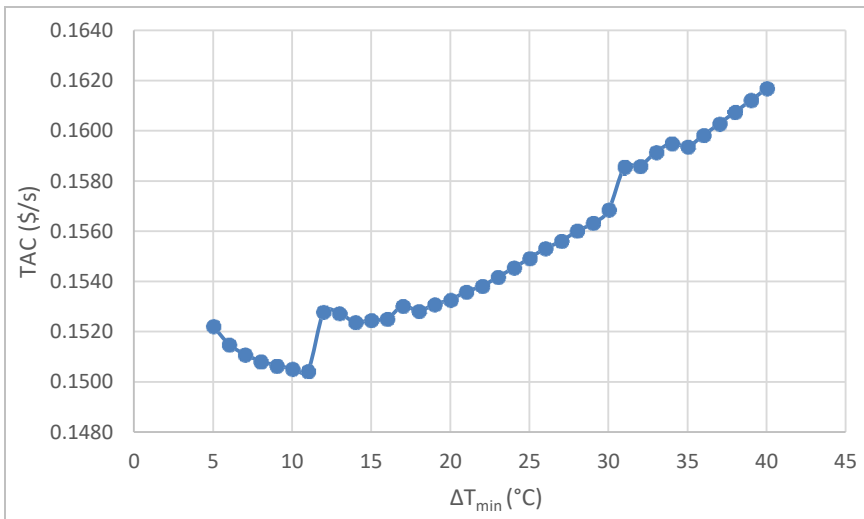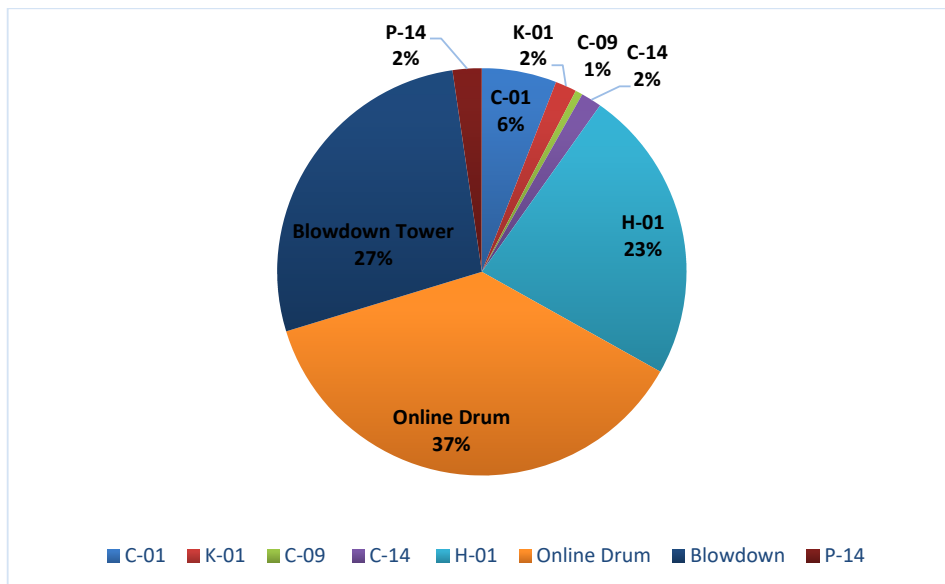
4.2 Exergy analysis

The exergy analysis for the main process equipment involves the calculation of both physical and chemical exergy destructions, in addition to exergy efficiency. Table 3 shows the exergy efficiency calculations.

Table 3 Exergy efficiency of main equipment

Equipment	Description	Total efficiency, %	Physical efficiency, %	Chemical efficiency, %
C-01	Main fractionator	99.80	77.80	100.00
K-01	Wet gas compressor	99.60	53.40	
C-09	Debutaniser tower	100.00	81.00	
C-14	Rerun tower	99.50	9.70	
C-13	Amin treatment tower	99.90	95.50	
P-01	Feed pump	100.00	98.20	
H-01	Coking heater	99.20	237.60	98.50
D-01	Online drum	98.77	89.40	98.90
C-04	Blowdown tower	58.30	1.50	79.70
P-14	Jet pump (coke cutting pump)	80.30	66.50	100.00

Figure 4 summarises the total exergy destruction of the main equipment with exergy destruction values greater than 1 GJ/h. It is clear from the figure that the top four exergy destruction equipment are the blowdown tower, the online coke drum, the blowdown tower C-04 and the main fractionator. These results are similar to those reported for an atmospheric distillation unit (Rivero et al., 2004) where the top three exergy destruction equipment were the two fired heaters and a fractionator.

Figure 4 Total exergy destruction (see online version for colours)

4.2.1 *Chemical exergy analysis*

Calculation of the standard chemical exergy for each individual hypothetical component indicated that the chemical exergy efficiency is nearly 100% when performing chemical exergy balance for hydrocarbon streams in non-conversion process equipment where no chemical reaction occurs. This is clear in the main fractionator C-01, debutaniser tower C-09, Amine treatment tower, and Rerun Tower C-14, as well as the pumps P-01, P-14, and compressor K-01, which show only minor chemical exergy destruction (less than 0.5%). On the other hand, for the blowdown tower C-04, there is a 20% efficiency loss in chemical exergy as the inlet stream contains more than 97% water with the balance hydrocarbons. The reason of this exergy loss is that most of the tower heat input is introduced as steam with a high chemical exergy (9.5 kJ/mol) and is withdrawn from the tower as liquid water with a very low chemical exergy (0.9 kJ/mol).

The chemical exergy efficiency of the coke drum where the delayed coking reactions occur is around 99% (no waste stream rejected to the environment). The fired heater is considered as a conversion equipment where the chemical reaction results in converting fuel gas chemical exergy into physical exergy reflected as temperature increase within the process. Despite the fact that the flue gas stream is wasted from the stack to the environment, the chemical exergy efficiency is high, around 98.5%. This is because flue gas chemical exergy is very low compared to the chemical exergy of the hydrocarbon fuel gas stream.
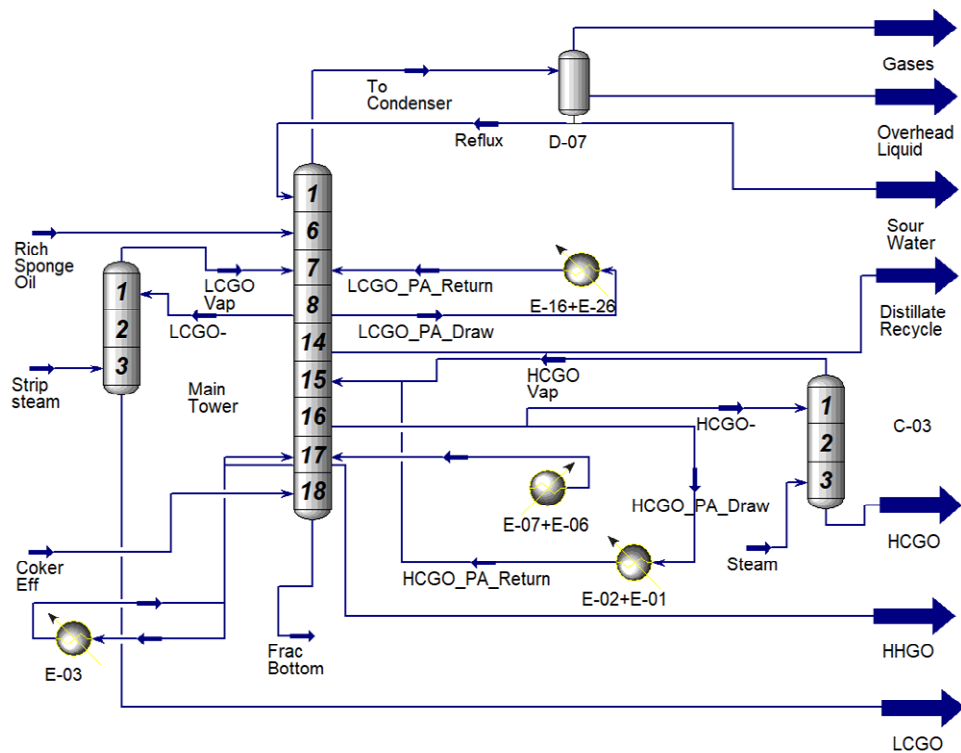
4.2.2 *Physical exergy*

Physical exergy destruction is more common within process equipment than chemical exergy. The reason for this is that pressure and temperature decreases are more common in process equipment than composition changes. In terms of physical exergy, the sour amine treatment tower has the highest exergy efficiency of more than 95% because it only involves absorption with no reboiler or condenser. It is followed by the debutaniser, the exergy efficiency of which is approximately 81% due to the presence of a reboiler and a condenser.

For the primary fractionator C-01, the tower has no reboiler as the tower feed has the highest temperature and is fed to the tower bottom as shown in Figure 5. Three pump around are located throughout the tower to control the temperature gradient. Pump around heat is easily recovered by preheating the feed and producing low and medium pressure steam. However, there is a significant flaw in the tower heat recovery system, as the top reflux of the tower uses an air cooler with an inlet temperature of 127°C associated with an exergy loss of around 13 GJ/h, while the total system physical exergy loss is 19 GJ/h. As a result, the exergy efficiency of C-01 is 78% compared to 93% if all the reflux duty had been recovered. An efficiency of 97.3% has been reported (Lei et al., 2016) for a DCU fractionator system having a slightly different configuration (disregarding the heat exchange equipment).

Exergy efficiency for the rerun tower C-14 is very low as the inlet stream from the debutaniser bottom is a liquid at a low temperature of 115°C, while the tower top and bottom temperatures are higher. As a result, much duty is consumed in the reboiler, especially that 92% of the feed is required to be vaporised and condensed from the top reflux drum. For this reason, the reboiler exergy is six times the physical exergy of the inlet process stream. Also, there is no recovery of the reboiler duty through the tower top

air cooler, with an inlet temperature of 142°C. Accordingly, C-14's physical exergy efficiency is less than 10%. The best proposal is to improve the exergy efficiency of the system by first trying to adjust the main fractionator top temperature to produce on spec naphtha during normal operation directly from the debutaniser, without using the low exergy efficiency rerun column C-14. The benefit of this proposal is that no capital costs are incurred because it enables making use of the design margin available in the main fractionator reflux pump and the LCGO pump. The rerun column is still used in this proposal, but only during plant startup to expedite on-spec production.

Figure 5 Main fractionator C-01 flow sheet (see online version for colours)

The physical exergy efficiency of C-04 is only 1.5%, the lowest in the plant because around 15% physical exergy loss arises from mixing the cold stream (14) and the hot HSO (15) recycle feeding the tower. Furthermore, the top of the tower has an air cooler with an inlet temperature of around 175°C, which accounts for approximately 70% of the total physical exergy loss of the blowdown tower (not utilised for any heat recovery).

The two-stage compressor K-01 has two compression stages with an interstage water cooler. The physical exergy efficiency is 53.4%. The exergy destruction is caused by two main factors: the compressor efficiency, and the interstage cooler, which accounts for approximately 22.6% of the total exergy destruction. The physical exergy destruction efficiency for feed pump P-01 is 98.2%. Despite the high efficiency, an exergy destruction of around 0.14 GJ/h takes place in the discharge control valve in addition to the 0.46 GJ/h loss due to pump efficiency. As a result, it is recommended that the diameter of the pump impeller be reduced in order to save this wasted energy.

The endothermic reaction in the online coke drum causes a drop of about 50°C between its inlet and outlet. The energy used for the endothermic reaction causes some of the inlet stream's physical exergy to be transferred into chemical exergy in the outlet stream. For this reason, the physical exergy efficiency is less than 90%. The actual chemical exergy efficiency of a coke drum should be slightly lower than 99% mentioned in the previous section and this in order to account for the endothermic reaction exergy. In such cases, when there is a transfer between chemical and physical exergy, the total exergy destruction is very important as it reflects the net exergy efficiency which is 98.77%.

The physical exergy in the coke drum is converted to chemical exergy, whereas the opposite occurs in the fired heater. In the fired heater, the chemical exergy of the fuel gas is converted to physical exergy, which is reflected in the higher heater outlet temperature. However, in this case, the governing factor is the heater's total exergy efficiency, which is around 99.2%.

4.2.3 *Applying exergy recommendations*

The proposed modifications to the blowdown tower are discussed in detail as an example of exergy recommendations. The water is supplied to this tower intermittently for three hours in a 16-hour cycle, so it is preferable to use it for steam generation rather than in the HEN. It is proposed to improve this system utilising the very high temperature of the tower inlet to produce high pressure steam by replacing mixing of tower inlet (that cools hot vapours with cold water) with a steam generator as shown in Figure 6. The drum vapours are first introduced to the high-pressure steam super heater then to the saturated steam generator. A BFW pre-heater is also used to precool the tower overhead before the air cooler. This proposal produces 9,750 kg/h of superheated steam during the first hour of quenching in the 16-hour cycle, and about 10,000 kg/h of medium pressure steam during the second and third hours of quenching as the temperature drops.

Figure 6   Blowdown tower HPS and MPS proposal (see online version for colours)
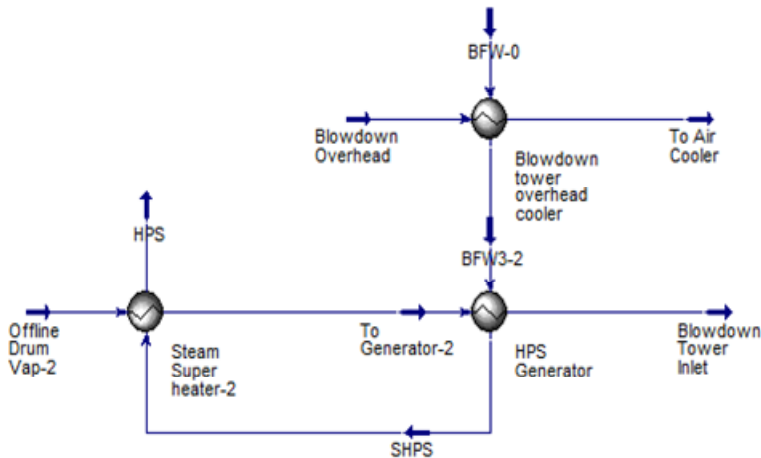
Despite that this proposal appears to be energy efficient, exergy analysis reveals that the physical exergy efficiency increased only from 1.5% to less than 23% since the inlet vapours to the tower are mostly steam, and the vapour phase accounts for the majority of the chemical and physical exergy of the steam. For maximum exergy recovery, the heat of the condensing steam should be recovered through a heat exchanger.

The limitation is that the condensation temperature of the coke drum steam is around 115°C (at the inlet pressure) which is not suitable for steam generation. To produce the maximum amount of steam, the quenching step should be accomplished at a higher pressure. Accordingly, the latent heat from the coke drum vapour could be recovered and used to generate low-pressure steam. Since the maximum design pressure of the coke drum is 7 kg/cm²g, the system could produce low-pressure steam as shown in Figure 7. This can be accomplished by raising the quenching pressure to 4.5 kg/cm²g and using it to generate low pressure steam at 175°C with a flow rate of 75,500 kg/h for the first three hours of water quenching in the 16-hour cycle.

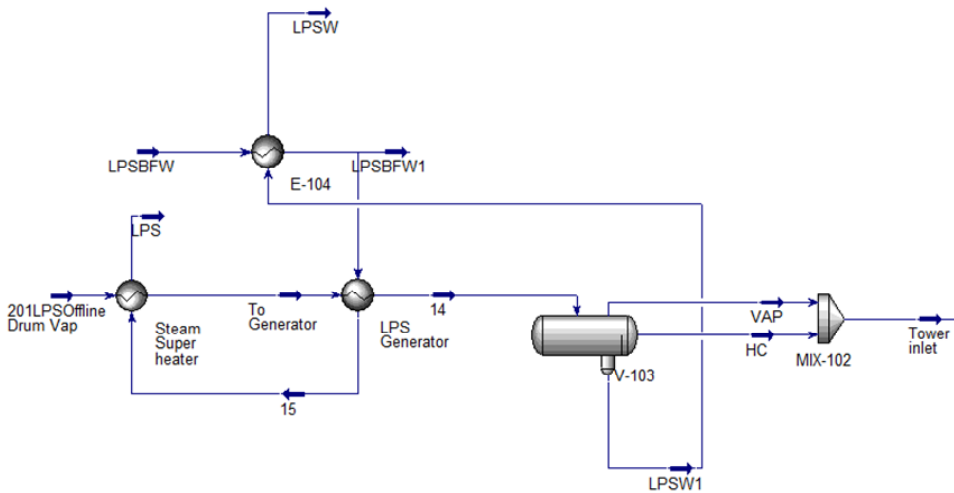**Figure 7**   LPS production proposal (see online version for colours)

Table 4   Blowdown tower alternatives

Case	Original	HPS+MPS	LPS
Total efficiency	58.3%	66.4%	99.2%
Physical efficiency	1.5%	22.6%	88.2%
Chemical efficiency	79.7%	82.9%	104.1%
Total exergy destructions, GJ/h	113.7	68.2	1.8
Saving due to steam production, $	-	156,774	769,381
Additional total installed cost, $	-	883,700	635,500

The physical exergy efficiency would thus be increased to 88.2% and the total exergy to more than 99%. This proposal saves exergy of around 56 GJ/h. Table 4 compares the original design with the above two proposals. The total installed cost is calculated using Aspen Process Economic Estimator V11, and the savings due to steam production are calculated using refinery steam cost data. Although the flow rate of low-pressure steam is

much higher than that of the high-pressure steam, the cost of high-pressure steam equipment is higher due to the high-pressure rating. Accordingly, the low-pressure steam proposal has the lowest investment cost and the highest profit, making it the better solution.

5 Conclusions

Energy analysis was carried out through both pinch and exergy analyses. Pinch analysis was used for the heat exchanger network. The network's optimum temperature approach with the lowest total annual cost is 10°C. The total annualised cost drops to US$0.16/sec compared to US$0.22/sec in the original design. This amounts to a US$1.9 million annual savings (including capital and operating costs).

Exergy analysis is applied to the other equipment in the DCU. The proposal of low-pressure steam, LPS production in the blowdown system is recommended as it has an exergy destruction value of 1.8 GJ/h while the high-pressure steam HPS production has an exergy destruction value of 68.2 GJ/h compared to the original design exergy destruction of 113.7 GJ/h. The LPS proposal recovered approximately 75.5 t/h of LPS which is equivalent to annual savings of around US$0.9 million.

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Nomenclature

A	Area, m^2
BFW	Boiler feed water
BPSD	Barrel per standard day
CC	Capital cost, $
CCR	Conradson carbon residue
DCU	Delayed coker unit
DR	Distillate recycle
EX	Exergy
FZGO	Flash zone gas oil
HCGO	Heavy coker gas oil
HPS	High pressure steam
HSO	Heavy slop oil
LCGO	Light coker gas oil
LPG	Liquified petroleum gas
LPS	Low pressure steam

Nomenclature (continued)

MCGO	Medium coker gas oil
Min	Minimum
MPS	Medium pressure steam
NHV	Net heating value
NR	Natural recycle
OC	Operating cost, $/y
PA	Pump around
SG	Specific gravity
T	Temperature, °C
TAC	Total annualised cost, $/y
V	Vanadium
W	Water
X	Mole fraction
Z	Mass fraction

Greek letters

β	Chemical exergy correction factor
γ_i	Activity coefficient

Subscripts

0	Reference environment
C	Carbon
Ch	Chemical
I	Component
in	Inlet
J	Component
Kin	Kinetic
out	Outlet
Pot	potential
Ph	Physical
Q	Amount of transferred heat
q	Standard chemical
W	Work