



Energy consumption and heat recovery of an industrial fluidized catalytic cracking process based on cost savings

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Abstract

Energy consumption is a significant issue in operation design for low-cost sustainable production and is accomplished by heat integration giving overall environmental advantages via reducing carbon emissions. Heat recovery is a beneficial tool that determines the minimum cooling and heating demand through recovery and re-use of energy within the process. Thus in this study, process of heat recovery and energy consumption of the fluidized catalytic cracking (FCC) is investigated to recover most of the external energy and reducing the environmental effect in addition to maximizing the productivity with minimum overall cost of the process. Where the performance of the FCC units plays a major role on the overall economics of refinery plants and improvement in operation or control of FCC units, it will result in dramatic economic benefits. The heat integration process is done based on experimental information from pilot scale, mathematical modeling developed and commercial process reported in our earlier study.

Keywords FCC · Heat integration · Energy consumption · Heat recovery · Cost savings

Introduction

One of the assignments with which chemical engineers constantly tend to be involved is scaling up of pilot plant tests to full-scale production. Because of the high operating cost of a pilot plant, this progression is beginning to be outperformed in different cases by generating the full-scale unit depending on the operation of a small-scale plant named a microplant. With a specific end goal making such manner effectively, an exhaustive comprehension of the chemical kinetics and transport restrictions is essential [1, 8]. The evaluation of the optimal cooling and heating prerequisites (the minimum) uncover imperative energy savings. For example, Union Carbide in the United States of America and Imperial Chemical Industries in the United Kingdom have both detailed the consequences of various case examinations that reference 30–50% energy savings in contrast with traditional practice [5]. CO₂ contents in the air have risen from 270 to 380 ppm by 2006. The essential human source of CO₂ in the air is generated by consuming petroleum derivatives

towards energy generation and transportation opportunities. To keep away from or lessen global warming, an important concentration in all CO₂ emissions should be accomplished (25%–40% by 1990 and 80%–95% by 2050) [6, 8]. Nevertheless, more productive usage of energy consumption reduces the negative effects of CO₂ discharges into the environment.

Industrial petroleum refining units generate huge amount of heat, which is normally ejected to the environment using either air or cooling water frameworks. Some processes are employed to recover such energy as a part of integration process system as well as for heating in local and commercial operations via hot water network [12].

Energy saving is very significant in process design, and estimating minimum cooling and heating requirement is very important in energy savings issues. Heat integration is a very beneficial tool used in calculating the cost of initial design and waste heat recovery provides environmental benefits for handling unit operators. A wide range of units, such as oil refinery and other industrial processes generate a large amount of heat that is discarded to the environment without reusing in other processes. Such behavior takes into account the recovery of some energy and a part of it is used in process integration. The cooling and heating units are called utility units involving hot utility and cold utility. Hot utility includes furnace, boilers, hot water, steam, and

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generators. Cold utility includes cold water from external source. In recovery system, the process streams exchange heat so as to reduce the cold and hot utility requirements of the heat exchangers, which are the only units in a heat recovery system. Heat exchanger is a unit in which heat is transferred from the hot fluid to the cold fluid. Conventional outline strategies start by planning the reactor, the separation system, the heat exchanger and lastly end by using utilities for providing residual requirements [2, 5, 11, 13].

More recently, we [7] employed the FCC reactor to optimize the process conditions of the operation for the purpose of maximizing the conversion as well as the octane number while minimizing the coke amount in the regenerator based on pilot plant experiments at a very high temperature (460 °C–540 °C) requiring high energy consumption in the process. Also, energy consumption for the lab scale is an unimportant issue (ignored) and no extra utility was needed as the quantities of reactants and products were little at laboratory scale, hence recovery issues were not a point in the laboratory-scale operation. However, the design when scaled-up to a commercial level [7] offers the chance of energy savings via appropriate process integration. In commercial operation, heat recovery and energy consumption should be taken into account to decrease environmental effect in addition to reducing the treatment operation cost. Therefore, the main focus of this study is to maximize heat recovery of an industrial fluidized catalytic cracking reactor utilizing the mathematical model developed earlier [7] and then analyze the heat integration process while minimizing overall annual cost of such process.

The experimental data

The experimental results have been taken from literature [16]. A brief description of the materials, apparatus and experimental procedure used for getting the experimental results are as follows.

The reaction takes place in plug flow reactor (65 cm (reactor length) 1.5 cm (reactor diameter) and 65 cm³ (catalyst bed volume)) with zeolite as a catalyst (pore volume of 0.37 (cm³/g) and 338.8 (cm²/g) surface area with particle diameter of 75 μm having a spherical shape) that should be loaded before each run. The catalyst-to-oil weight ratio (CTO) was adjusted by keeping the catalyst amount loaded and the feedstock (which is vacuum gas oil (VGO)) under the following characterizations: density (at 20 °C) is 0.919 (g/cm³), Conradson carbon residue (CCR) is 4.13 (wt%), molecular weight of 400 (g/mol) and sulfur content of 1.61 (wt%) rate constant. Steam was passed over the catalyst bed at a fixed flow rate at a set reaction temperature for 20 min before each run. Liquid products were collected in a glass receiver after condensation and the gaseous products

were collected in a gas collecting bottle by water displacement. The spent catalyst was stripped by steam for 30 min to recover the entrapped hydrocarbons. Subsequently, the catalyst bed was heated to 680 °C to burn off the coke with oxygen and the intermediate CO was converted to CO₂ by a CO converter. The total amount of the coke on the spent catalyst was determined by a CO₂ infrared detector. The process flow diagram of the experimental setup is shown in Fig. 1. Further information relating to the experimental work can be found in Xiong et al. [16].

Energy consumption and heat recovery of the FCC unit

In the pilot plant-scale process, heat recovery was not an issue and the energy consumption was negligible, while in industrial processes, energy consumption will be a big issue and heat recovery must be taken into account. A heat integrated system was considered for reducing overall energy consumption (hence reducing environmental effect). However, it is important to incorporate for heat exchangers in the process system. Evaluating the required design is an important issue to minimize the energy consumption and to maximize the energy recovery hence consequently minimizing the capital investment.

Figure 2 describes the heat integration process proposed for FCC unit. Many of utility units operated in series including heaters and coolers and the utility hot units regulate the final temperature of the cold fluid to the required temperature of the reaction, while the utility cold units regulate the final temperature of the hot fluid to a required temperature of the next process.

In general, heat exchanging is working in series with heating and cooling systems. The heater controls the final temperature of the cold liquid to the needed reaction temperature, and the cooler alters the final temperature of the hot liquid to prerequisites of the subsequent stage of the procedure. It is shown in Fig. 2 that the feed stock is pumped by a pump (PU) into a heat exchanger 1 (H.E.1) and heated from T_{in} to T_{in1} it is fed into a heat exchanger 2 (H.E.2) to further heating from T_{in1} to T_{in2} to the required temperature of the reaction T_R then it is fed to the furnace. The reaction occurs inside a reactor (R), and after completion the hot product stream leaves the reactor and goes to the fractionators (FR). Then it is cooled from T_{out} to T_{out1} via heat exchanger 1 (H.E.1) by contacting with the main feed stock and further cooling via heat exchanger 2 (H.E.2) from T_{out2} to T_{out3} . The final product temperature is cooled from $T_{out,1}$ to T_{F1} via cooler (C1) by contacting with a cold water stream at temperature $T_{W,1}$ that is heated into $T_{W,2}$. The final product of the other stream temperature is cooled from $T_{out,3}$ to 2

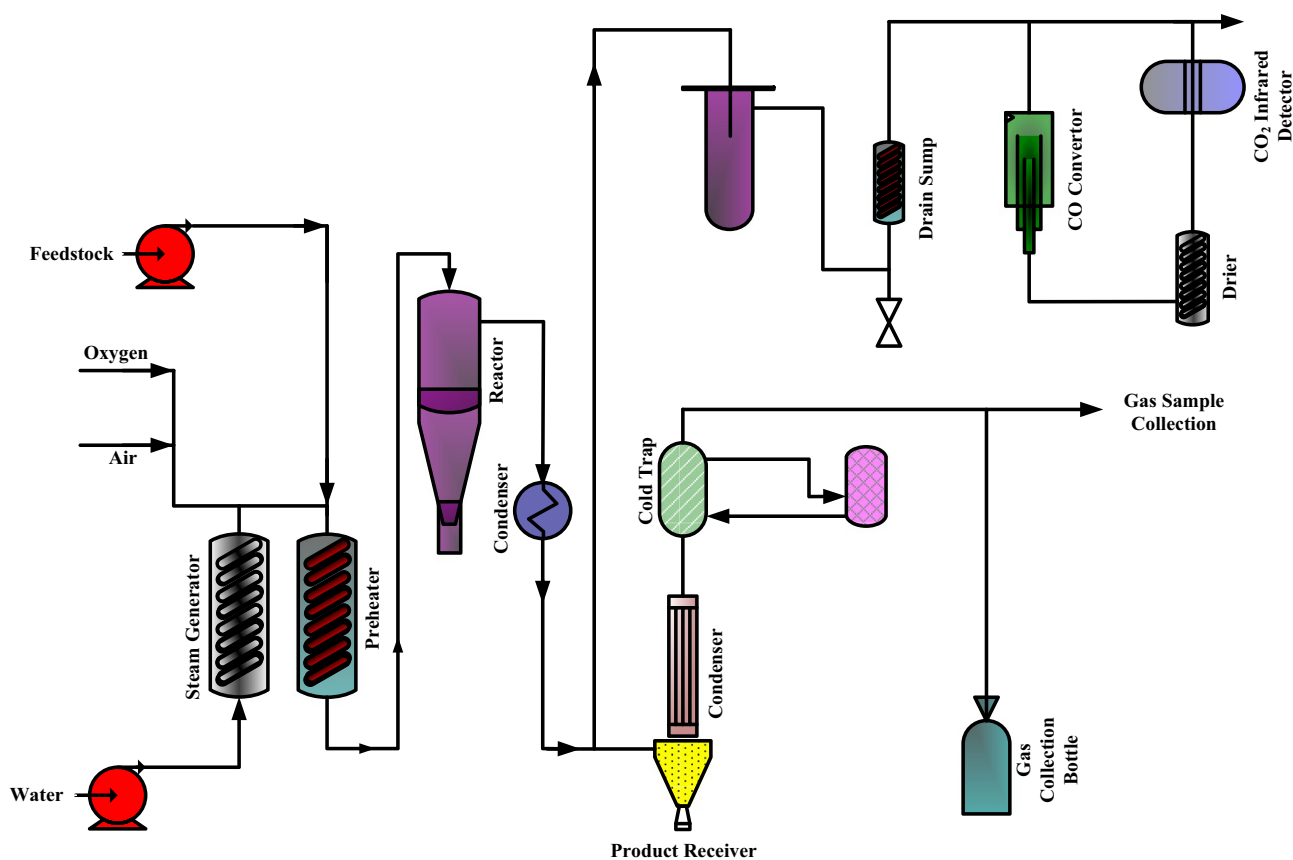


Fig. 1 Process flow diagram of the experimental setup. Adapted from [16]

via cooler (C2) by contacting with the cold water stream at temperature $T_{W,3}$, which is heated into $T_{W,4}$.

Process model equations

The main concern of this study is to minimize energy consumption and maximize heat recovery of an industrial fluidized catalytic cracking unit. More recently, we [7] have developed a new mathematical model for the fluidized catalytic cracking operation taking into consideration the complex hydrodynamics of the reactor regenerator system based on a new six-lump kinetic model for the riser. The best kinetic parameters of the relevant reactions have been evaluated employing the optimization technique depending upon the experimental results taken from literature. The optimal operating conditions (mainly, reaction temp (T), catalyst-to-oil ratio (CTO) and weight hourly space velocity (WHSV) on the product composition were investigated and the optimal kinetic parameters obtained from the pilot plant scale were used to develop an industrial fluidized catalytic cracking process. The optimal operating conditions based on maximum conversion of vacuum gas oil with minimum cost in addition to maximizing the octane number of gasoline and coke content deposited on the

catalyst within the regenerator, have also been studied. Thus, mass balance equations, energy balance equations, reaction rate equations, catalyst deactivation with the characteristics of the catalyst bed utilized, riser hydrodynamics, regenerator model, dense bed modelling, dilute phase modelling, physical and chemical properties of the reactants and products, equipment and procedure, scale-up of FCC reactor with their optimization processes and results related to the optimal design and operation of such unit can be found with more detail in Jarullah et al. [7]. Therefore, the equations related to the heat exchangers of an industrial FCC reactor utilizing the optimal results obtained previously [7] can be stated as follows:

Heat exchanger 1 (H.E.1)

The feed stock is pumped through a pump (PU) then it is heated via heat exchanger 1 (H.E.1) from T_{in} to T_{in1} with product1 that leaving the fractionators while product1 is cooled from T_{out} to T_{out1} . The heat transfer rate of each stream is shown in Fig. 3 and can be described as follows:

$$Q_{1_{VGO}} = (\rho_{VGO} c_{p_{VGO}} Q_{VGO})(T_{in1} - T_{in}), \quad (1)$$

$$Q_{1_{prod.}} = (\rho_{LCO} c_{p_{LCO}} Q_{LCO})(T_{out} - T_{out1}), \quad (2)$$

Fig. 2 Process of heat integration proposed for FCC reactor

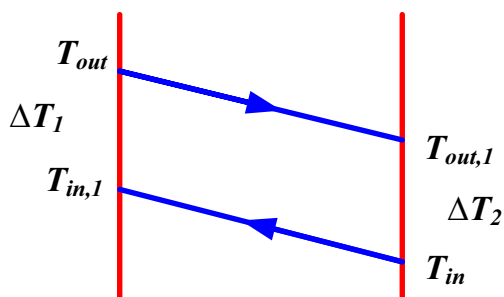
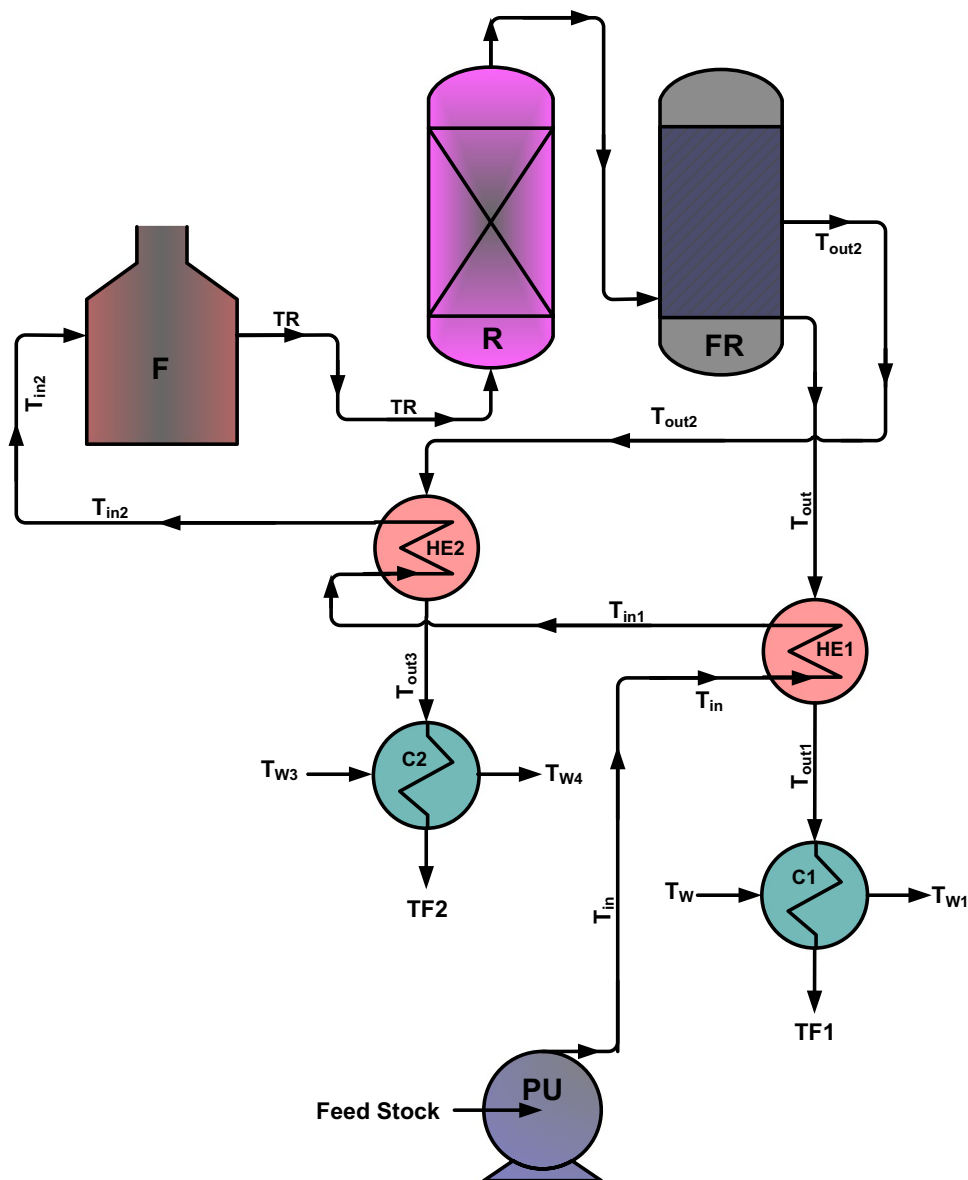


Fig. 3 Temperature gradient of H.E.1

$$\Delta T_{LMI} = \frac{\Delta T_1 - \Delta T_2}{\ln\left(\frac{\Delta T_1}{\Delta T_2}\right)}, \tag{5}$$

$$\Delta T_1 = T_{out} - T_{in,1} \tag{6}$$

$$\Delta T_2 = T_{out,1} - T_{in} \tag{7}$$

where T_{in} is the inlet temperature of vacuum gas oil into the heat exchanger 1 (H.E.1), °C; $T_{in,1}$ is the outlet temperature of phenol from heat exchanger 1 (H.E.1), °C; T_{out} is the inlet temperature of hot product mixture into the heat exchanger 1 (H.E.1), °C; $T_{out,1}$ is the outlet temperature of hot product mixture from heat exchanger 1 (H.E.1), °C; Q_{VGO} is the volumetric flow rate of feed stock, cm^3/s ; Q_{LCO} is the volumetric flow rate of light cycle oil, cm^3/s ; A_1 is the heat transfer area

$$Q_{VGO}^1 = Q_{prod}^1, \tag{3}$$

$$Q_{VGO}^1 = U_1 A_1 \Delta T_{LMI}, \tag{4}$$

of heat exchanger 1(H.E.1); ρ_{VGO} is the density of VGO, g/cm^3 ; cp_{VGO} is the specific heat capacity of VGO, $J/g K$; ρ_{LCO} is the density of LCO, g/cm^3 ; cp_{LCO} is the specific heat capacity of LCO, $J/g K$; U_1 is the overall heat transfer coefficient for heat exchanger 1(H.E.1), $W/m^2 K$; ΔT_{LM1} is the log mean temperature difference for heat exchanger 1(H.E.1); $Q1_{VGO}$ is the heat duty of feed stock in heat exchanger 1(H.E.1), W ; and $Q1_{prod.}$ is the heat duty of product mixture in heat exchanger 1(H.E.1), W .

Heat exchanger 2 (H.E.2)

The feed stock stream leaving the H.E.1 is passed through H.E.2 for raising the feed temperature. The stream of product mixture 2 out of FR is used to heat the feed into heat exchanger 2 (H.E.2). In this case, the feed is heated from T_{in1} to T_{in2} and at the same time the product mixture 2 is cooled from T_{out2} to T_{out3} as shown in Fig. 4. The equations used in heat exchanger 2 (H.E.2) are:

$$Q2_{VGO} = (\rho_{VGO}cp_{VGO}Q_{VGO})(T_{in2} - T_{in1}), \tag{8}$$

$$Q2_{prod.} = (\rho_{GLN}cp_{GLN}Q_{GLN})(T_{out,2} - T_{out,3}), \tag{9}$$

$$Q2_{VGO} = U_2A_2\Delta T_{LM2}, \tag{10}$$

$$Q2_{VGO} = Q2_{prod.}, \tag{11}$$

$$\Delta T_{LM2} = \frac{\Delta T_3 - \Delta T_4}{\ln\left(\frac{\Delta T_3}{\Delta T_4}\right)}, \tag{12}$$

$$\Delta T_3 = T_{out,2} - T_{in,2}, \tag{13}$$

$$\Delta T_4 = T_{out,3} - T_{in,1}, \tag{14}$$

where T_{in2} is the feed stock outlet temperature into heat exchanger 2 (H.E.2), $^{\circ}C$; $T_{out,3}$ is the outlet temperature of product mixture 2 from heat exchanger 2 (H.E.2), $^{\circ}C$; $T_{out,2}$ is the inlet temperature of product mixture 2 from heat exchanger 2 (H.E.2), $^{\circ}C$; A_2 is the heat transfer area of heat exchanger 2 (H.E.2), m^2 ; ρ_{GLN} is the density of gasoline

(GLN), g/cm^3 ; cp_{GLN} is the specific heat capacity of GLN, $J/g K$; U_2 is the overall heat transfer coefficient for heat exchanger 2 (H.E.2), $W/m^2 K$; $\Delta T_{LM,2}$ is the log mean temperature difference for heat exchanger 2 (H.E.2), K ; $Q2_{prod.}$ is the heat duty of product mixture 2 in heat exchanger 2 (H.E.2), W ; and Q_{GLN} is the volumetric flow rate of gasoline, cm^3/s .

Cooler 1 (C1)

The outlet stream product mixture from heat exchanger 1 (H.E.1) will be cooled through a cooler (C1) from $T_{out,1}$ to T_{F1} using water as a cold fluid at temperature $T_{W,1}$ heated to $T_{W,2}$, described in Fig. 5. The equations used in cooler 1 are:

$$Q3_w = m_w cp^w (T_{W,2} - T_{W,1}), \tag{15}$$

$$Q3_{prod.} = (\rho_{LCO}cp_{LCO}Q_{LCO})(T_{out,1} - T_{F1}), \tag{16}$$

$$Q3_w = Q3_{prod.}, \tag{17}$$

$$Q3_w = U_3A_3\Delta T_{LM,3}, \tag{18}$$

$$\Delta T_{LM,3} = \frac{\Delta T_5 - \Delta T_6}{\ln\left(\frac{\Delta T_5}{\Delta T_6}\right)}, \tag{19}$$

$$\Delta T_5 = T_{out,1} - T_{W,2}, \tag{20}$$

$$\Delta T_6 = T_F - T_{W,1}, \tag{21}$$

where T_{F1} is the outlet final temperature of product mixture from cooler, $^{\circ}C$; $T_{W,1}$ is the inlet temperature of water into cooler, $^{\circ}C$; $T_{W,2}$ is the outlet temperature of water from cooler, $^{\circ}C$; m_w is the mass flow rate of cooling water, g/s ; cp^w is the specific heat capacity of water, $J/g K$; A_3 is the heat transfer area of cooler, m^2 ; U_3 is the overall heat transfer coefficient for cooler, $W/m^2 K$; $\Delta T_{LM,3}$ is the log mean temperature difference for cooler, (K) ; $Q3_w$ is the heat duty of water in cooler, W ; and $Q3_{prod.}$ the heat duty of product mixture in cooler, W .

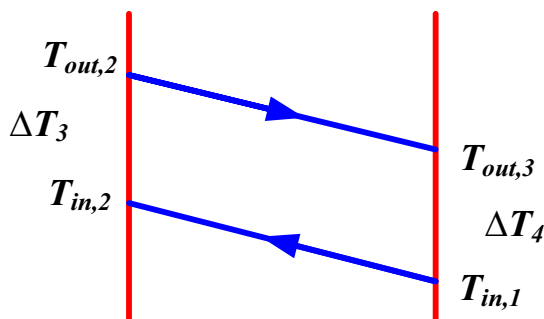


Fig. 4 Temperature gradient of H.E.2

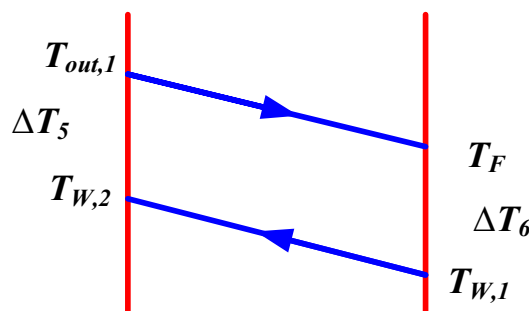


Fig. 5 Temperature gradient of cooler 1 (C1)

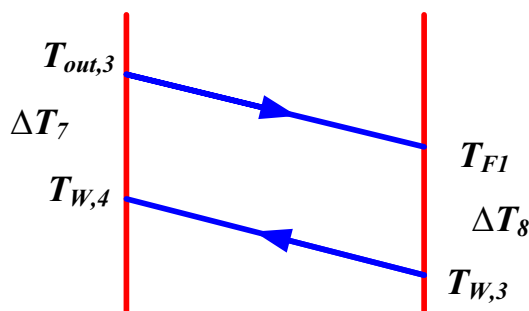


Fig. 6 Temperature gradient of cooler (C2)

Cooler 2 (C2)

The outlet stream product mixture 2 from heat exchanger 2 (H.E.2) will be cooled through a cooler (C2) from $T_{out,3}$ to T_{F2} using water as a cold fluid at temperature $T_{W,2}$ heated to $T_{W,4}$, described in Fig. 6. The equations used in cooler 2 are:

$$Q_{4w} = m_{W2} cp^W (T_{W,2} - T_{W,1}), \quad (22)$$

$$Q_{4prod.} = (\rho_{LCO} cp_{LCO} Q_{LCO}) (T_{out,1} - T_F), \quad (23)$$

$$Q_{4w} = Q_{4prod.}, \quad (24)$$

$$Q_{4w} = U_4 A_4 \Delta T_{LM,3}, \quad (25)$$

$$\Delta T_{LM,4} = \frac{\Delta T_8 - \Delta T_7}{\ln \left(\frac{\Delta T_8}{\Delta T_7} \right)}, \quad (26)$$

$$\Delta T_7 = T_{out,3} - T_{W,4}, \quad (27)$$

$$\Delta T_8 = T_{F1} - T_{W,3}, \quad (28)$$

where T_{F1} is the outlet final temperature of product mixture 1 from cooler 2, °C; $T_{W,3}$ is the inlet temperature of water into cooler 2, °C; $T_{W,4}$ is the outlet temperature of water from cooler 2, °C; m_{W2} is the mass flow rate of cooling water, gm/s; cp^W is the specific heat capacity of water, J/g K; A_4 is the heat transfer area of cooler 2, m²; U_4 is the overall heat transfer coefficient for cooler 2, W/m² K; $\Delta T_{LM,4}$ is the log mean temperature difference for cooler 2, (K); Q_{4w} is the heat duty of water in cooler 2, W; $Q_{4prod.}$ is the heat duty of product mixture in cooler 2, W.

The total heat transfer area (A_t , m²) can be calculated as follows:

$$A_t = A_1 + A_2 + A_3 + A_4. \quad (29)$$

Furnace 1 (F1)

Furnace is needed to further raise the feed temperature to obtain the reaction temperature T_R . In this case, the feed

stock is fed into furnace (F1) separately to preheat the feed from T_{in2} to the required temperature of the reaction T_R . The equations of furnace used are:

$$QF = (\rho_{VGO} cp_{VGO} Q_{VGO}) (T_R - T_{in2}) \quad (30)$$

where Q_{VGO} is the heat duty of VGO in furnace, W; and QF is the total heat duty in furnace, W.

It is known that the physical properties such as density, heat capacity, etc., are temperature dependent that can be calculated (in each equipment) using the following equation:

$$T_{av} = \frac{T_{in} + T_{out}}{2}, \quad (31)$$

where T_{in} and T_{out} are inlet and outlet temperatures for item of each equipment, respectively.

Optimization problem formulation

The optimization process as a function of the cost models of the whole process is presented here. The overall annual process cost can be calculated using the following relationship [14]:

$$OAPC = ACC + OPC, \quad (32)$$

where OPAC is the overall annual process cost (\$/year).

The annualized capital cost (ACC, \$/year) can be calculated from the total capital cost (TCC, \$), which includes the cost of the main equipment in the unit process such as reactor, compressor, heat exchanger, pump and furnace:

$$\text{Annualized capital cost (ACC, \$ / yr)} = \text{Total capital cost (TCC, \$)} \times \frac{i(1+i)^N}{(1+i)^{N-1}}, \quad (33)$$

N is number of years and i is the fractional interest per year; $N = 10$ years, $i = 5\%$ [14].

The total capital cost (TCC, \$) can be calculated from the following equation [15]:

$$\begin{aligned} \text{Total capital cost (TCC, \$)} \\ = \text{Capital cost of installed equipment (CC, \$)} \\ \times 1.4, \end{aligned} \quad (34)$$

$$\begin{aligned} \text{Capital cost of installed equipment (CC, \$)} \\ = \text{reactor cost (} C_r \text{)} + \text{heat exchanger} \\ \text{Cost (} C_{\text{heatexch.}} \text{)} + \text{pump cost (} C_{\text{Pump}} \text{)} \\ + \text{Furnace cost (} C_{\text{Furn.}} \text{)}, \end{aligned} \quad (35)$$

The operating cost (OPC) in Eq. (32) above is determined utilizing the following equation [15]:

$$\text{OPC (\$/year)} = \text{Variable operating cost (VOPC)} \quad (36)$$

Variable operating cost (VOPC, \$/yr)

$$= \text{Heating cost } (C_{\text{Heating}}) + \text{pumping cost } (C_{\text{Pumping}}) + \text{cooling cost } (C_{\text{cooling}}) \quad (37)$$

- Heating cost (C_{Heating}) (\$/yr):

$$C_{\text{Heating}} (\$/\text{year}) = (Q_F(\text{kW})) \left(\frac{0.06\$}{\text{kWh}} \right) \left(\frac{24\text{h}}{1\text{day}} \right) \left(\frac{342\text{days}}{1\text{year}} \right) \quad (38)$$

- Pumping cost (C_{Pumping}) (\$/yr):

$$C_{\text{pumping}} (\$/\text{year}) = (Q_p(\text{kW})) \left(\frac{0.06\$}{\text{kWh}} \right) \left(\frac{24\text{h}}{1\text{day}} \right) \left(\frac{342\text{days}}{\text{year}} \right) \quad (39)$$

- Cooling cost (C_{cooling}) (\$/year) can be estimated by the following relationship with a price of cooling water (0.00375 \$/kg) [8]:

$$C_{\text{cooling}} (\$/\text{year}) = \left(m_w \left(\frac{\text{kg}}{\text{h}} \right) \right) \left(\frac{0.00375\$}{\text{kg}} \right) \left(\frac{24\text{h}}{1\text{day}} \right) \left(\frac{342\text{days}}{\text{year}} \right). \quad (40)$$

Others capital cost (CC, \$) for each term can be calculated as follows [5, 15]:

- Reactor cost (C_r) (\$):

$$C_r (\$) = \left(\frac{M\&S}{280} \right) 101.9 D_r^{1.066} L_r^{0.802} (2.18 + F_c) \quad (41)$$

$$F_c = F_m F_p \quad (42)$$

D_r and L_r are the reactor diameter and length, respectively. $M \& S$ is Marshal and Swift index for cost escalation ($M \& S = 1536.5$) [3].

F_c , F_m and F_p are dimensionless factors that are function of the construction material and operating pressure ($F_m = 3.67$, $F_p = 3.93$) [5, 4].

- Heat exchanger cost ($C_{\text{heatexch.}}$) (\$):

$$C_{\text{heatexch.}} (\$) = \left(\frac{M\&S}{280} \right) 210.78 A_t^{0.65} (2.29 + F_c), \quad (43)$$

$$F_c = F_m (F_d + F_p). \quad (44)$$

- Pump cost (C_{Pump}) (\$):

$$C_{\text{Pump}} (\$) = \left(\frac{M\&S}{280} \right) 9.84 \times 10^3 F_c \left(\frac{Q_p}{4} \right)^{0.55}, \quad (45)$$

$$F_c = F_m F_p F_T. \quad (46)$$

- Furnace cost ($C_{\text{Furn.}}$) (\$):

$$C_{\text{Furn.}} (\$) = \left(\frac{M\&S}{280} \right) 5.52 \times 103 Q_F^{0.85} (1.27 + F_c), \quad (47)$$

$$F_c = F_m + F_p + F_d, \quad (48)$$

where Q_F is the heat duty of the furnace, W ; F_m , F_p , F_c , F_d and F_T are dimensionless factors that are functions of the construction material, operating pressure and temperature in addition to the design type.

The optimization problem can be stated as shown below:

Given	Inlet temperature of feed stock $T_{in,0}$, outlet product mixture 1 T_{out1} , outlet product mixture 2 T_{out2} reaction temperature T_R , inlet water temperature $T_{W,1}$, $T_{W,3}$ volumetric flow rates of feed stock (Q_{VGO})
Optimize	T_F , $T_{W,2}$
So as to minimize	The total annual cost of the process (OAPC)
Subjected to	Process constraints and linear bounds on all decision variables

The optimization problem can mathematically be represented as follows:

Min OAPC

T_F , $T_{W,2}$

s.t. $f(x(z), u(z), v)$ (model, equality constraints)

$T_F^L < T_F < T_F^U$ (inequality constraints)

$T_{W,2}^L < T_{W,2} < T_{W,2}^U$ (inequality constraints)

$\Delta T_{W,2}^L < \Delta T_{W,2} < \Delta T_{W,2}^U$ (inequality constraints)

$\Delta T_F^L < \Delta T_F < \Delta T_F^U$ (inequality constraints)

$T_F = T_F^*$ (equality constraints),

where $\Delta T_{W,2}$ is the temperature difference between inlet and outlet temperature of water in the cooler. Practically, the best temperature difference between inlet and outlet water in the cooler is 5–25 °C. ΔT_F is the temperature difference between inlet and outlet temperature of feed stock in the furnace. T_F^* is the target final temperature of the product. The optimization solution method used by gPROMS is a two-step method known as feasible path approach. The first step performs the simulation to converge all the equality constraints (described by f) and to satisfy the inequality constraints. The second step performs the optimization (updates the values of the decision variables such as the kinetic parameters) [10]. The optimization problem is posed as a nonlinear programming (NLP) problem and is solved using a successive quadratic programming (SQP) method within gPROMS software.

Results and discussion

Kinetic parameters estimation

The optimal set of kinetic parameters of an industrial fluidized catalytic cracking reactions was estimated utilizing the objective function via minimizing the sum of squared error between the experimental results and the model results including all the data related to the product compositions, which are presented in Jarullah et al. [7]. The kinetic parameters have been calculated employing nonlinear approach, where such regression is used to calculate the order of the reactions [VGO reactions order (n_1), LCO reactions order (n_2), GLN reactions order (n_3), LPG

reactions order (n_4)], activation energies and pre-exponential factors for all reactions related to the FCC reactions, simultaneously. These kinetic parameters were obtained accurately among all results based on average absolute error of less than 5%, and hence can be confidently utilized for reactor design, operation and control. The optimal kinetic parameters with the optimal operating conditions results are summarized in Table 1 for convenience (more details related to the kinetic parameters, maximum conversion and octane number, minimum coke content, etc., can be found in Jarullah et al. [7].

Figure 7 shows the parity plots of the experimental and predicted yields of LCO, GLN, DG and coke in an industrial fluidized catalytic cracking unit. All the plots reveal straight lines with slopes close to unity. Hence, excellent

Table 1 Optimal kinetic parameters and operating conditions obtained for industrial FCC process

Parameter	Symbol	Unit	Value
Order of vacuum gas oil concentration	n_1	(–)	0.925367
Order of light cycle oil concentration	n_2	(–)	1.000001
Order of gasoline concentration	n_3	(–)	0.999785
Order of liquefied petroleum gases concentration	n_4	(–)	0.999413
Activation energy for reaction VGO → LCO	E_1	kJ/mol	20431.1
Activation energy for reaction VGO → GLN	E_2	kJ/mol	23082.6
Activation energy for reaction VGO → LPG	E_3	kJ/mol	23082.6
Activation energy for reaction VGO → DG	E_4	kJ/mol	22271.8
Activation energy for reaction VGO → CK	E_5	kJ/mol	9006.57
Activation energy for reaction LCO → GLN	E_6	kJ/mol	49215.6
Activation energy for reaction LCO → CK	E_7	kJ/mol	19854.4
Activation energy for reaction GLN → LPG	E_8	kJ/mol	70463.8
Activation energy for reaction GLN → DG	E_9	kJ/mol	88051.1
Activation energy for reaction LPG → DG	E_{10}	kJ/mol	65992.4
Pre-exponential factor for reaction VGO → LCO	A_1	(cm ³ /g) 0.074633 s ⁻¹	8.15295 × 10 ⁶
Pre-exponential factor for reaction VGO → GLN	A_2	(cm ³ /g) 0.074633 s ⁻¹	391.828
Pre-exponential factor for reaction VGO → LPG	A_3	(cm ³ /g) 0.074633 s ⁻¹	1276.72
Pre-exponential factor for reaction VGO → DG	A_4	(cm ³ /g) 0.074633 s ⁻¹	1656.55
Pre-exponential factor for reaction VGO → Ck	A_5	(cm ³ /g) 0.074633 s ⁻¹	1204.11
Pre-exponential factor for reaction LCO → GLN	A_6	s ⁻¹	598.233
Pre-exponential factor for reaction LCO → CK	A_7	s ⁻¹	20986.8
Pre-exponential factor for reaction GLN → LPG	A_8	(cm ³ /g) 0.000215 s ⁻¹	3.0214 × 10 ⁷
Pre-exponential factor for reaction VGO → DG	A_9	(cm ³ /g) 0.000215 s ⁻¹	1.46191 × 10 ⁷
Pre-exponential factor for reaction LPG → DG	A_{10}	(cm ³ /g) 0.000215 s ⁻¹	28090.8
Sum of Square Errors	SSE	(–)	4.42621 × 10 ⁻⁷
Operating conditions	Symbol	Unit	Values
Optimal operating conditions obtained for industrial FCC process			
Reaction temperature	TR	K	820.012
Weight Hourly space velocity	WHSV	hr ⁻¹	2.002
Catalyst-to-oil ratio	CTO	(–)	10.00
Conversion	CV	(–)	87.6075
Octane number	Octane number	(–)	97.5722

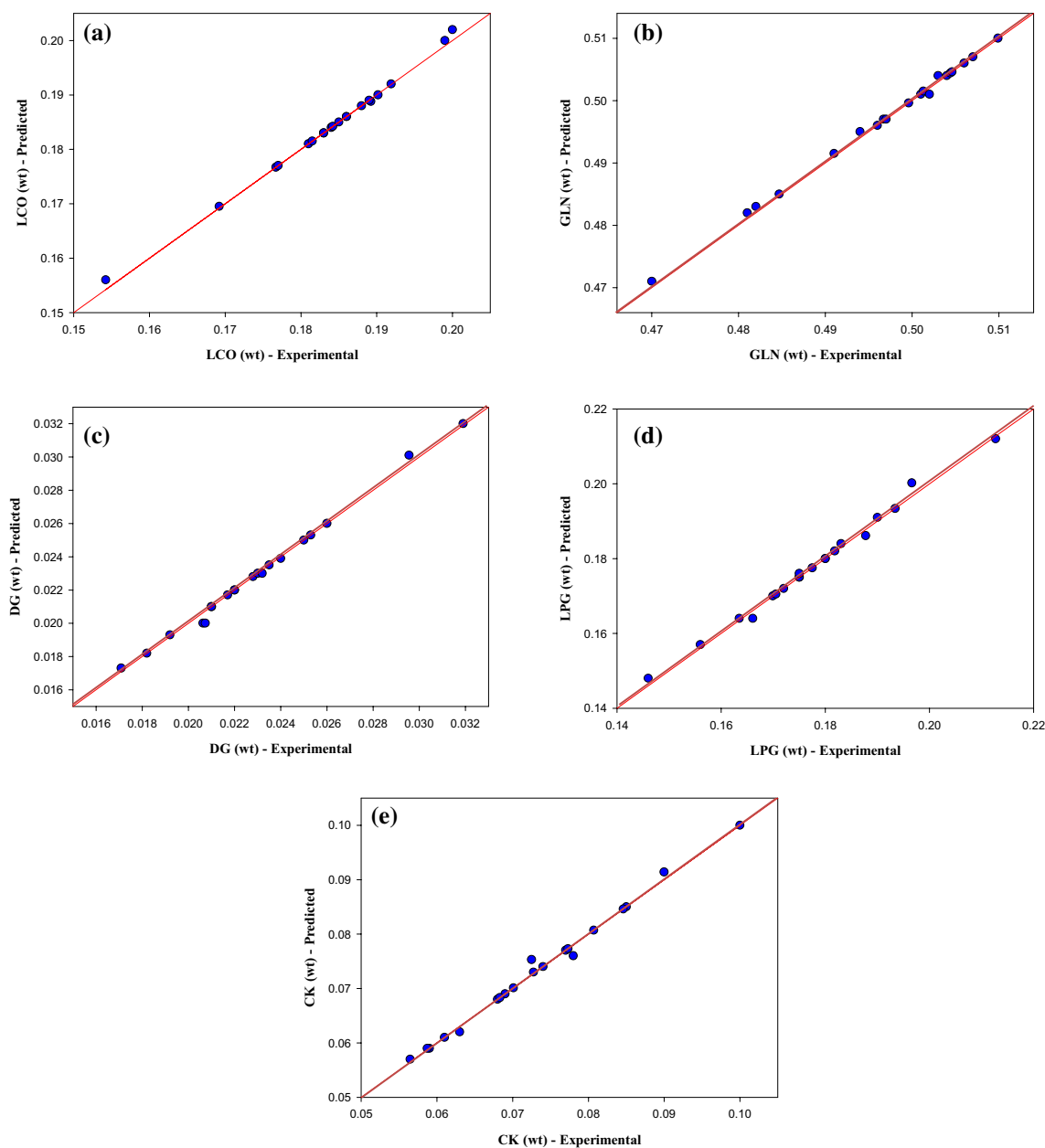


Fig. 7 Comparison between the experimental and simulated data for **a** light cycle oil, **b** gasoline, **c** dry gaseous, **d** liquefied petroleum gases, **e** coke

agreement was achieved between the pilot plant data of the unit and the simulated results of this study. Thus, the model can be employed for describing the performance of the industrial fluidized catalytic cracking reactions at several operating conditions for which experimental data are not available.

Energy recovery and cost savings

Here, an industrial fluidized catalytic cracking operation with energy consumption and heat recovery choice

is regarded to reduce overall energy consumption (subsequently lessening environmental impact). Notwithstanding, various heat exchangers are added to the process. The goal is to provide a retrofit outline for the purpose of reducing the energy consumption and maximizing the heat recovery leading to reduce the capital investment. The heater controls the final temperature of the cold liquid to the required reaction temperature, and the cooler sets out the final temperature of the hot liquid to necessities of the following stage of the operation.

Table 2 Values of parameters employed in this model

Parameter	Unit	Value
$T_{in,0}$	°C	24
T_{out}	°C	347
T_{out2}	°C	192
$T_{w,1}$	°C	20
U_1	W/m ² K	150
U_2	W/m ² K	900

Table 3 Dimensionless constants used in this model

Dimensionless parameters	Furnace	Pump	Heat exchanger
F_m	0.75	1	3.75
F_p	0.15	1.9	0.625
F_d	1	0	1
F_T	0	1	0

In this study, the values of the constant parameters with factors, coefficients and dimensionless constants are listed in Tables 2 and 3. Process heat integration variables are evaluated and optimized which are summarized in Table 4.

Based on the results presented in Table 4, it is observed that the minimum total cost (C_t) and amounts of cooling water (m_w) with heating integration of the fluidized catalytic cracking process are less than those obtained without heating integration. Also, it is noticed that the cost saving is (34.87%) in comparison with the cost obtained without heating integration to reach reaction temperature and to minimize the final product temperature. To achieve the target final temperature of the product (26 °C), it is noticed that the amount of cooling water needed to reach the final temperature without heat integration is larger than that used with heating integration due to the heat recovery. The energy requirement is also taken into account in this work and shown in Table 4. It has been

noted that the energy saving obtained here is about 48% compared with those without heat integration. Such result gives a clear indication that the CO₂ emissions will be reduced by 48%, which has the added benefit of significantly reducing environmental impact.

The performance of an industrial FCC reactor

Fluidized catalytic cracking is considered one of the most significant operations where the large molecules (undesirable compound) are converted into more valuable compounds leading to achieve the market request via increasing the middle distillates yields [particularly car fuel (gasoline) and diesel fuel (light cycle oil)] in addition to light gases obtained by this process. Therefore, an increase in productivity of such cuts is considered a major issue in the FCC operations and plays a significant role in coping with these issues.

After obtaining the optimal kinetic parameters, the maximum conversion, maximum RON, minimum coke deposited and optimal operating conditions as recently reported [7] and continuing with the present work by employing the whole results obtained, Fig. 8 illustrates the product yield (productivity) of an industrial fluidized catalytic cracking process using VGO as a feedstock. Based on the results presented in this figure, it is clearly observed that the productivity of valuable products (mainly gasoline and diesel fuel) increased utilizing the optimal data obtained while VGO content decreased. This increase in the productivity of valuable products can be attributed to conversion of vacuum gas oil compounds and long molecules that are concentrated in this cut (VGO) to light compounds (GLN, LCO, LPG, DG) owing to catalytic cracking reactions in such fluidized reactor.

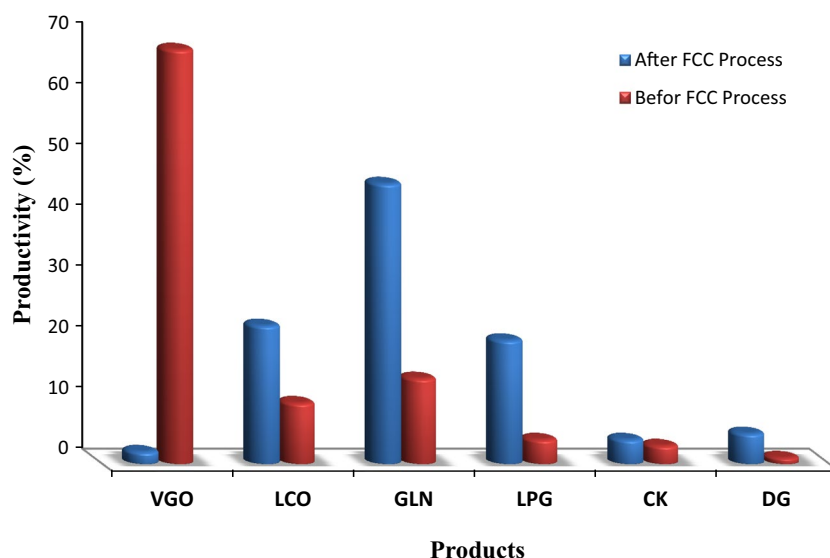
Conclusions

Process of heat integration of the FCC unit was investigated to recover most of the external energy and reducing

Table 4 Optimization results for heat integration system

Variables	Without heating integration	With heating integration	Decision variable type	Optimized value
A_r (m ²)	15.852884	68.744644	$T_{out,1}$ (°C)	231.22226
C_t (\$/year)	2.42532704E8	1.57962688E8	$T_{out,3}$ (°C)	125.18642
m_w (kg/hr)	1928.32956	1806.57972	T_{w1} (°C)	115.0
C_s (%)	–	34.87	T_{w2} (°C)	120
$Q1_{VGO}$ (W)	–	67098.45	T_{F1} (°C)	28
$Q2_{VGO}$ (W)	–	43591.605	T_F^* (°C)	26
$Q3_w$ (W)	–	138432.78	–	–
$Q4_w$ (W)	–	85681.945	–	–
Q_t (W)	–	3453183.5	–	–
ES (%)	–	48	–	–

Fig. 8 Product yield of an industrial FCC reactions obtained in this work



the environmental effect in addition to maximizing the productivity with minimum cost of the process. The energy recovery by heat integration between hot and cold streams from a large-scale fluidized catalytic cracking process was investigated and optimized. The energy consumption and heat recovery process was performed based on experimental information from pilot scale, mathematical modeling and commercial process and the minimum energy requirement and heat recovery were optimized. It was found that the heat integration process is very useful and efficient in commercial processes which maximizes production with minimum cost of the process. The cost savings and energy savings have been calculated to be 35% and 48%, respectively, in comparison with the process of without heat integration. Also, applying the optimal results obtained, the yield of product of fluidized catalytic cracking processes has clearly been increased.

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