



## Process Systems Engineering and Process Safety

## Study on co-cracking performance of different hydrocarbon mixture in a steam pyrolysis furnace☆

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## ABSTRACT

Co-cracking is a process where the mixtures of different hydrocarbon feedstocks are cracked in a steam pyrolysis furnace, and widely adopted in chemical industries. In this work, the simulations of the co-cracking of ethane and propane, and LPG and naphtha mixtures have been conducted, and the software packages of COILSIM1D and SimCO are used to account for the cracking process in a tube reactor. The effects of the mixing ratio, coil outlet temperature, and pressure on cracking performance have been discussed in detail. The co-cracking of ethane and propane mixture leads to a lower profitability than the cracking of single ethane or single propane. For naphtha, cracking with LPG leads to a higher profitability than single cracking of naphtha, and more LPG can produce a higher profitability.

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## 1. Introduction

Ethylene and propylene are important raw materials of the petrochemical industries for other chemical products and mainly produced by the cracking of petroleum distillates. The cracking is a process where the different hydrocarbons are transformed into light hydrocarbons such as paraffin, olefin and so on in a steam pyrolysis furnace (thereafter called furnace). At present, there are several different cracking routines applied in the industries [1], such as catalyst cracking [2,3], hydrocracking and steam cracking.

The steam cracking has a wide range of applications in the chemical industries, and it is also the most energy-consuming process in chemical industries. It is found that the radiation section of a furnace consumes approximately 65% of total process energy [4]. In a furnace, a gaseous or liquid hydrocarbon feedstock like naphtha, LPG, ethane and propane is thermally cracked with steam instead of oxygen. At present, the naphtha is a widely used feedstock and its consumption continues to increase in China. Nowadays, it is one of the most important feedstocks of the industrial furnace [5].

A steam pyrolysis furnace is mainly composed of two sections: a convection section and a radiation section. Some investigations have been performed in the convection section [6–9], however, more

attention has been paid to the radiation section in the past few decades, in which the mixtures of hydrocarbon feedstocks are thermally cracked into olefins, aromatics, methane hydrogen and so on. In the cracking process, the product yield distribution is strongly affected by the hydrocarbon partial pressure, the coil outlet temperature, and the resident time, especially the type of feedings. High olefin yields tend to high temperature, low partial pressure, and high resident times [10]. To predict the effect of these parameters on the olefin yields, a good reaction model is essential, and the modeling methods can be divided into three kinds: empirical model, molecular reaction model and free radical reaction model. A detail model based on the individual molecular reactions had been outlined by Van Damme *et al.* [11]. Van Geem *et al.* [12] carried out an investigation of the cracking simulation of ethane by a 1-D model and a 2-D reactor model, and obtained the effect of the temperature profiles on yields in steam cracking. Marco *et al.* [13] established a kinetic modeling study of ethane cracking for optimal ethylene yield. These models based on free radical reactions just apply to the chemical experiments in the laboratory without taking into consideration the real application problems in ethylene plants. The molecular reaction model was proposed by Kumar *et al.* [14], in which one primary reaction model with a first-order primary step and 21 secondary reactions were included. The Kumar model has been widely used in later researches [15–17]. For instance, Hu *et al.* [18,19] carried out a coupled simulation of a steam cracker with Kumar model to study the flow, combustion and heat transfer process in the furnace and obtained product yields and coke rates.

Simulation, control and optimization of ethylene cracking furnace can be carried out based on these reaction models. Masoumi

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et al. [20] developed a static and a dynamic cracking model to predict the optimal temperature profiles along the reactor and study the performances of temperature control loop for different controller parameters and disturbances. Mehdi et al. [21] carried out a dynamic optimization of a cracking reaction in a tube reactor by a 1-dimensional pseudo-dynamic model and improved operating profit by 13.1%. Shu et al. [22] proposed a distributed parameter model for a tube reactor in an ethylene cracking furnace based on 3-D combustion monitoring, and predicted the distributions of the heat flux and the tube temperature.

In the past years, more attention has been paid to the cracking of single feedstock like naphtha, ethane and propane. As the crude becomes heavier and heavier all over the world, the cost of the cracking of hydrocarbon feedstocks for ethylene and propylene increases gradually, especially in China whose demand for ethylene and propylene is high and increases gradually. Thus, the selection and optimum allocation of raw materials have received more attention, and the co-cracking of different feedstocks becomes more and more common. Co-cracking is a process where several different feedstocks are cracked together. Many researchers think that thermal cracking is a chemical reaction based on free radical mechanism, and thus an easily cracking and high-selectivity hydrocarbon can promote decomposition of other difficulty cracking components in the co-cracking process [23].

In the present work, the simulations of co-cracking of ethane and propane, and naphtha and LPG mixtures are performed with Coilsim1D [24,25] in tube reactors of two different furnaces: one is an industrial USC-M furnace and the other USC-U furnace. This paper aims to study the performance of the co-cracking of ethane and propane, and naphtha and LPG mixtures in the tube reactor of a furnace, and the effects of the mixing ratio of two different feedstocks, the coil outlet temperature and the coil outlet pressure on the product distribution are explicitly discussed.

## 2. Mathematical Models and Physical Properties

Coilsim1D [24,25] is developed by Laboratory for Chemical Technology in Ghent University. And it is a commercial steam cracking software developed to simulate the steam cracking of the hydrocarbons in a tube reactor including the simulations of the tube reactor and the run length of industrial steam cracking tubes. The simulation is performed based on a free-radical reaction mechanism and a 1-dimension reactor model. The software package SimCo is used for feedstock reconstruction using a neural network method and a Shannon entropy optimization method.

### 2.1. Reactor model

The steam cracking is a non-isothermal, non-adiabatic and non-isobaric process. A 1-D reactor model has been used in this work ignoring the effect of gradients in the radial direction. In a 1-D reactor model, a set of steady state continuity equations for feed components are solved simultaneously with the energy equation and the momentum equation. These equations have been listed in Table 1.

To perform a simulation of the cracking model the reactor conditions should be needed. In Coilsim1D, four methods have been specified: the temperature and pressure profile, the heat flux profile [26], the external wall temperature profile and the outlet conditions. In each way the coil inlet temperature, coil inlet pressure and the dilution of simulation are all required.

### 2.2. Feedstock model

The detail feedstock components are needed for a free-radical reaction model, but not for a molecular reaction model. During the past years, several analytical techniques to generate a detailed molecular

composition have been developed and improved such as gas chromatography (GC), gas chromatography-mass spectrometry (GC-MS) and high performance liquid chromatography (HPLC) [24]. However these techniques are error-prone and time-consuming for chemical industries. In general, only several global characteristics are measured on line called the commercial indices such as the specific density, the PIONA weight fractions and a set of ASTM boiling points. Thus, many methods have been proposed to get the information of detailed compositions from the commercial indices [24,27,28].

In this work, the simulation package SimCo is used to generate the detailed compositions based on a number of commercial indices. This soft contains two methods for the molecular reconstruction. One is the Shannon Entropy method which is the main method for all kinds of feedstock, and the other is an artificial neural network (ANN) which is only used for naphtha fractions [29]. ANN can only be used when the naphtha is indeed within range and all necessary commercial indices are available from the input file. The inputs of this neural network are the initial, 50% and final boiling point of the ASTM D86 boiling point curve, the global PIONA weight fractions and the specific density.

### 2.3. Physical properties

Usually, the steam cracking is always carried out at relatively low pressures and high temperatures. Hence, the process gas can be considered as an ideal gas mixture in the tube reactor, and thus the density of the gas mixture can be calculated by the ideal gas law:

$$\rho = \frac{p}{RT \sum (Y_i/M_i)} \quad (4)$$

The viscosity of the mixture is given by:

$$\mu = \sum_i \frac{X_i \mu_i}{\sum_j X_j \phi_{ij}} \quad (5)$$

In this equation,  $\mu_i$  is the viscosity of the pure component  $i$  and is defined as [30]: where  $P_{ci}$  and  $T_{ci}$  are the critical pressure and the critical

**Table 1**  
Balance equations

|                                                                                                                        |     |
|------------------------------------------------------------------------------------------------------------------------|-----|
| The steady state continuity equation                                                                                   |     |
| $\frac{dF_j}{dz} = \left( \sum_{k=1}^n v_{kj} r_{v,k} \right)$                                                         | (1) |
| The energy equation                                                                                                    |     |
| $\sum_j F_j c_{pj} \frac{dT}{dz} = \omega q + \Omega \sum_k R_{v,k} (-\Delta_f H_k^0)$                                 | (2) |
| The momentum equation                                                                                                  |     |
| $\frac{dp_t}{dz} = -\alpha \left( \frac{2f}{d_t} + \frac{\xi}{\pi r_b} \right) \rho v^2 - \alpha \rho v \frac{dv}{dz}$ | (3) |

**Table 2**  
Comparison between the modeling results and the experimental data (naphtha)

| Feedstock                 | Naphtha               |       |                          |
|---------------------------|-----------------------|-------|--------------------------|
|                           | Modeling results /wt% |       | Experimental data /wt%   |
|                           | COT                   | P/E   |                          |
| Ethylene                  | 21.29                 | 22.14 | 23.29                    |
| Propylene                 | 16.63                 | 16.72 | 17.59                    |
| P&E                       | 37.92                 | 38.86 | 40.88                    |
| High value-added products | 49.14                 | 50.4  | 49.45                    |
|                           |                       |       | Average relative error/% |
|                           |                       |       | 6.8                      |
|                           |                       |       | 5.2                      |
|                           |                       |       | 6.1                      |
|                           |                       |       | 1.25                     |

**Table 3**

Comparison between the modeling results and the experimental data (LPG)

| Feedstock                 | LPG                   |       |                        |                          |
|---------------------------|-----------------------|-------|------------------------|--------------------------|
|                           | Modeling results /wt% |       | Experimental data /wt% | Average relative error/% |
|                           | COT                   | P/E   |                        |                          |
| Ethylene                  | 42.04                 | 36.3  | 38.86                  | −0.8                     |
| Propylene                 | 5.83                  | 6.78  | 7.27                   | 13.2                     |
| P&E                       | 47.87                 | 43.08 | 46.13                  | 1.42                     |
| High value-added products | 52.19                 | 46.57 | 51.7                   | 3.5                      |

**Table 4**

Commercial indices of naphtha

|                            |        |
|----------------------------|--------|
| Specific density(20 °C)    | 0.715  |
| ASTM D86 boiling point /°C |        |
| IBP                        | 45     |
| 10%                        | 69     |
| 50%                        | 92     |
| 90%                        | 161    |
| 95%                        | 169    |
| FBP                        | 191    |
| POINA /wt%                 |        |
| P                          | 29.94  |
| I                          | 41.26  |
| O                          | 1.36   |
| N                          | 16.01  |
| A                          | 10.71  |
| S                          | 0.0307 |

temperature of the component  $i$ , and  $T_r$  is the ratio of the gas temperature and the critical temperature. In Eq. (5), the interaction parameter  $\Phi_{ij}$  is calculated by

$$\phi_{ij} = \frac{[1 + (\mu_i/u_j)^{1/2} (M_i/M_j)^{1/4}]^2}{\{8[1 + (M_i/M_j)]\}^{1/2}}. \quad (6)$$

**Table 5**

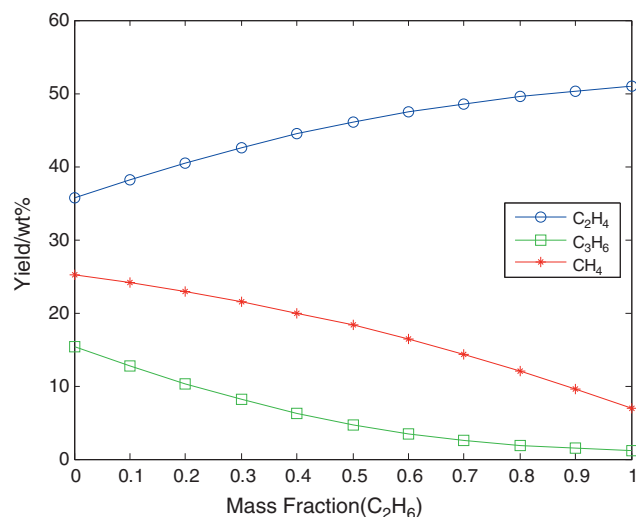
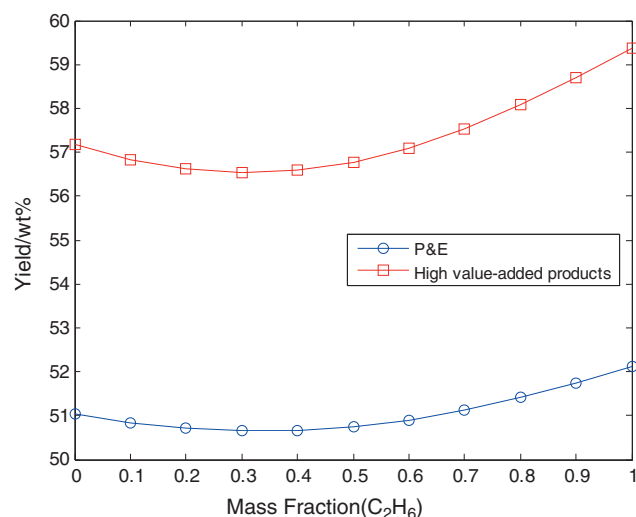
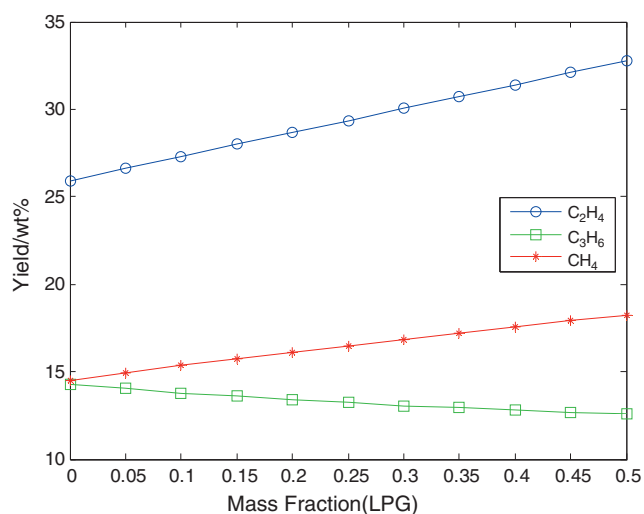
Feedstock compositions of LPG

| Feedstock                                | Mole fraction/% |
|------------------------------------------|-----------------|
| H <sub>2</sub>                           | 0.09            |
| CH <sub>4</sub>                          | 1.18            |
| C <sub>2</sub> H <sub>6</sub>            | 8.46            |
| C <sub>3</sub> H <sub>8</sub>            | 25.66           |
| C <sub>3</sub> H <sub>6</sub>            | 0.03            |
| <i>i</i> -C <sub>4</sub> H <sub>10</sub> | 35.42           |
| <i>n</i> -C <sub>4</sub> H <sub>10</sub> | 28.66           |
| <i>i</i> -C <sub>5</sub> H <sub>12</sub> | 0.45            |
| <i>n</i> -C <sub>5</sub> H <sub>12</sub> | 0               |
| C <sub>4</sub> =                         | 0.05            |

**Table 6**

Reactor conditions

| Parameters                         | U-shape tube | M-shape tube |
|------------------------------------|--------------|--------------|
| Mass flow rate /kg·h <sup>−1</sup> | 300          | 1600         |
| Steam/hydrocarbon                  | 0.55         | 0.3          |
| CIT /°C                            | 610          | 660          |
| CIP × 10 <sup>−5</sup> /Pa         | 2.48         | 2.5          |
| COP × 10 <sup>−5</sup> /Pa         | 1.84         | 1.97         |
| COT /°C                            | 850          | 850          |

**Fig. 1.** Effect of mass fraction (C<sub>2</sub>H<sub>6</sub>) on CH<sub>4</sub>, C<sub>2</sub>H<sub>4</sub>, C<sub>3</sub>H<sub>6</sub>.**Fig. 2.** Effect of mass fraction (C<sub>2</sub>H<sub>6</sub>) on the total yields of P&E and high value-added products.**Fig. 3.** Effect of mass fraction (LPG) on CH<sub>4</sub>, C<sub>2</sub>H<sub>4</sub>, C<sub>3</sub>H<sub>6</sub>.

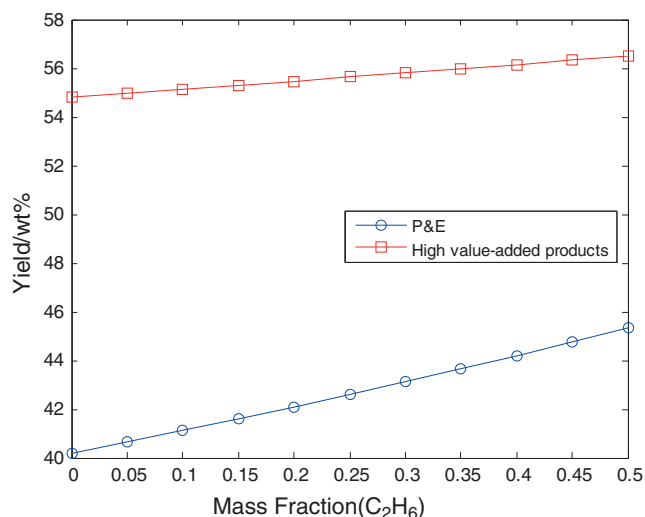


Fig. 4. Effect of mass fraction (LPG) on the total yields of P&E and high value-added products.

The thermal conductivity is calculated in the same way as the viscosity:

$$\lambda = \sum_i \frac{X_i \lambda_i}{\sum_j X_j \phi_{ij}} \quad (7)$$

The heat capacity of the mixture is computed based on a mass fraction average of the heat capacities of pure species as

$$C_p = \sum_i Y_i C_{pi} \quad (8)$$

### 3. Model Validation

To verify the reliability of the theoretical model, some simulations of naphtha cracking and LPG cracking are performed with Coilsim1D, and

different convergence conditions of COT and P/E are specified. Model validation is carried out by comparing the modeling results with the experimental data obtained from the petrochemical industries.

Table 2 compares the modeling results and the experimental data of naphtha cracking. As can be seen, the modeling results are all a bit smaller than the experimental data expect for the yield of high value-added products with the convergence condition of P/E, and the results with the convergence condition of P/E are closer to the experimental data than that with the convergence condition of COT. The average relative errors of the product yields are all below 7%. The comparison between the modeling results and the experimental data of LPG cracking is shown in Table 3. As shown, the average relative errors of product yields are small except for the error of propylene yield, and the results with the convergence condition of COT are closer to the experimental data.

In general, the modeling results are in good agreement with the experimental data and can reflect the product yields in the real steam pyrolysis furnace.

### 4. Model Setting

#### 4.1. Reactor geometry

The simulations of U-shape and W-shape tube reactors are performed in this work. Some parameters of tubes should be specified like tube layer, tube number, length of the coil, tube internal diameter and wall thickness.

#### 4.2. Feedstock setting

Ethane and propane, and naphtha and LPG mixtures are chosen as the feedstocks for the USC-M furnace and the USC-U furnace. The detailed information of naphtha and LPG is presented in Tables 4 and 5.

#### 4.3. Reactor conditions

The simulation model provides a set of ordinary (non-linear) differential equations which need to be solved. As mentioned before, the reactor conditions determine the solution of these equations. In this work the outlet conditions are specified as cracking severity

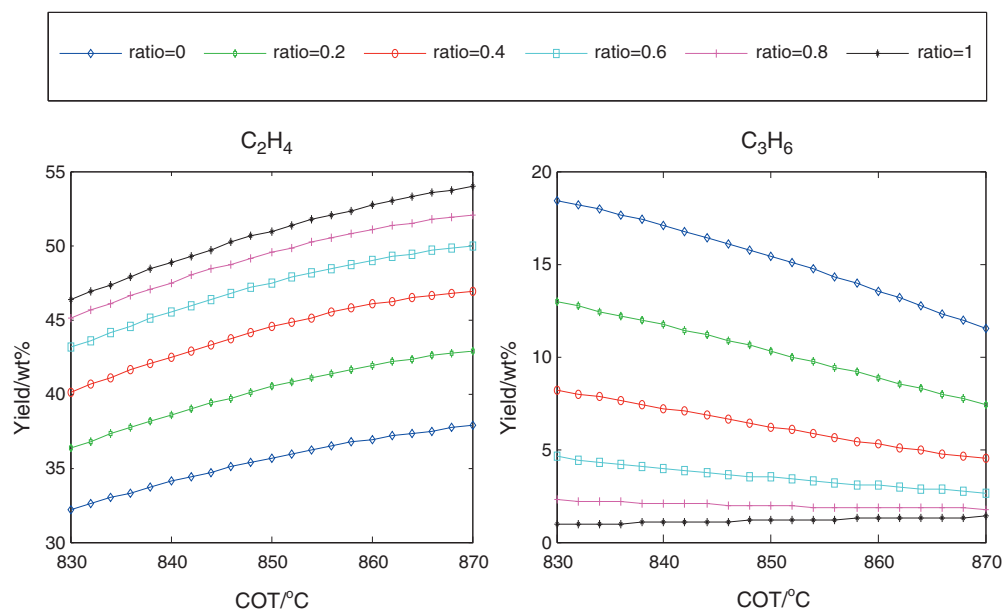


Fig. 5. COT effect on ethylene and propylene yields (C<sub>2</sub>/C<sub>3</sub>).

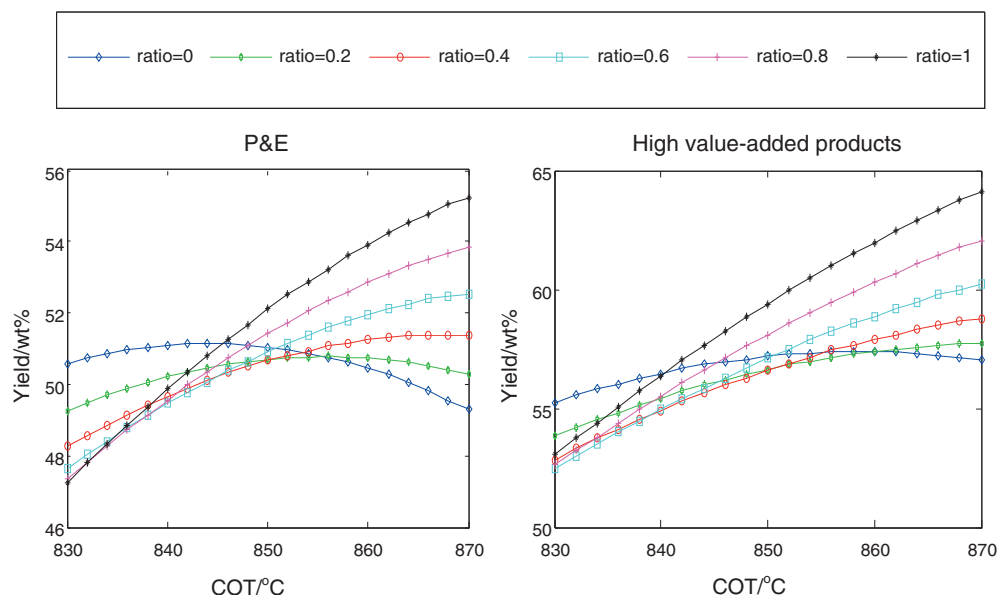


Fig. 6. COT effect on the yield of P&E and high value-added products(C2/C3).

indices such as the propylene/ethylene ratio, the methane/propylene ratio, a key component conversion, or a fixed yield of ethylene or methane.

Calculations are performed with the shooting method [31] which can calculate the inlet pressure and the heat flux profile corresponding to the specified cracking severity indices. For an industrial furnace, a pressure related to severity index (either COP or the ethylene to ethane yield ratio) and one of the temperature related severity index (propylene to ethylene yield ratio or COT) are available at the reactor outlet, so they can be set as boundary conditions in a simulation for an industrial furnace, then the inlet pressure and temperature become a settable variable to get the desired outlet specification.

The outlet conditions of the single tube reactor consist of mass flow rate, hydrocarbon/steam, the coil inlet temperature and pressure (a settable value), the coil outlet pressure and temperature. The detailed values are provided in Table 6.

## 5. Results and Discussion

The co-cracking of ethane and propane, and LPG and naphtha mixtures in the USC-M and USC-U furnaces was studied, respectively. The effects of the mixing ratio, the coil outlet temperature (COT) and the coil outlet pressure (COP) were discussed on the yield distributions of methane, ethylene and propylene, and some indices were also taken into consideration, *e.g.* the summarized yields of ethylene and propylene, the total yields of the high value-added products ( $H_2$ ,  $C_2H_4$ ,  $C_3H_6$ ,  $C_4H_6$ ,  $C_6H_6$ ) which determined the profitability of the industrial furnace.

### 5.1. Effect of mixing ratio of two different feedstocks

In the cracking process, ethane tends to be transformed to ethylene and propane tends to be transformed to propylene, and this corresponds to Fig. 1. It can be seen that ethylene yield increases

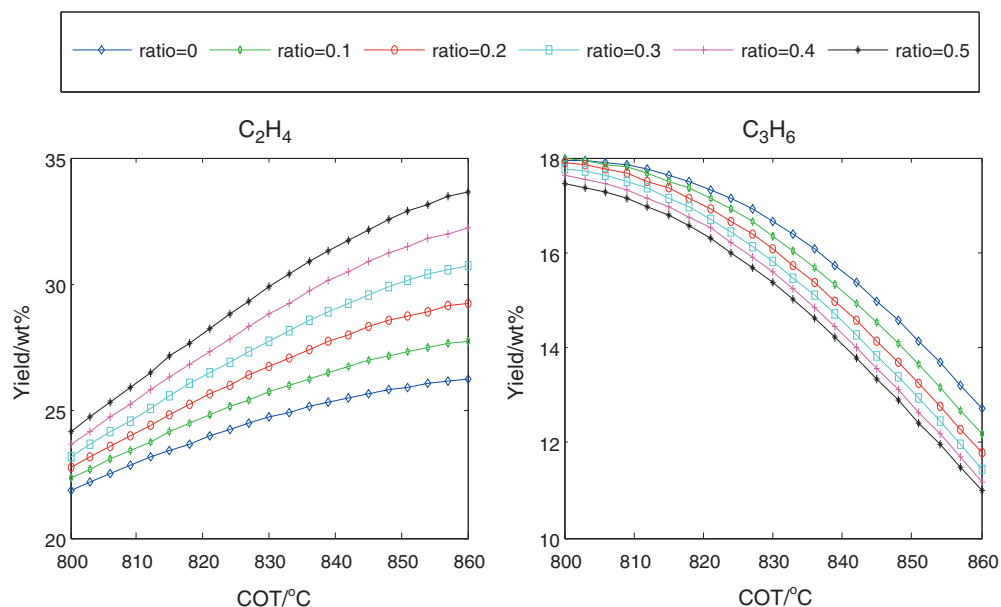


Fig. 7. COT effect on ethylene and propylene yields (LPG/NAP).

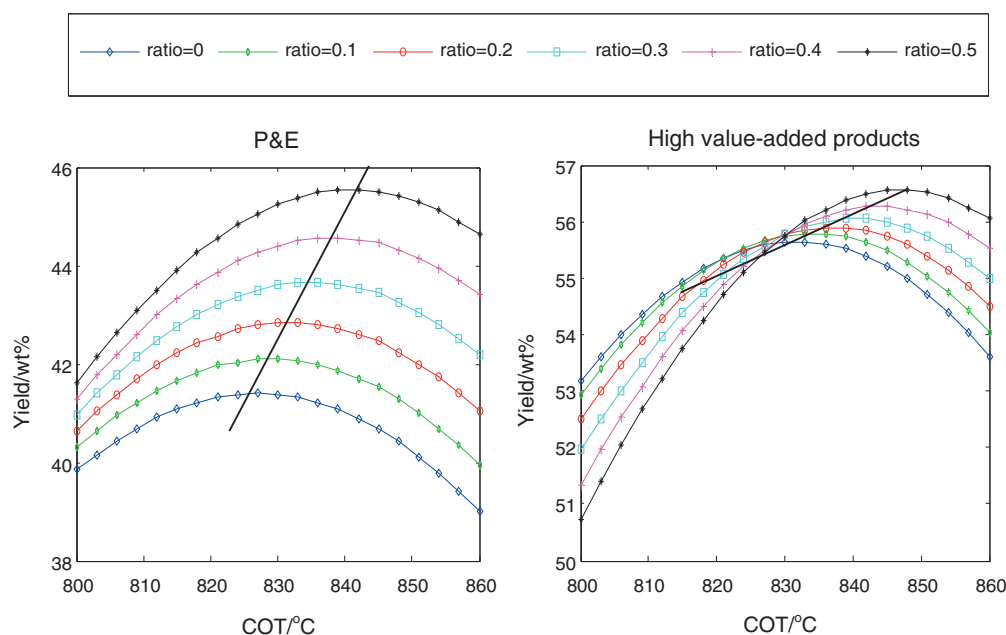


Fig. 8. COT effect on the yield of P&E and high value-added products (LPG/NAP).

with increasing proportion of ethane, and the growth rate tends to be gentle. Methane yield and propylene yield decrease, and propylene yield loss becomes slower gradually. Fig. 2 indicates the summarized yields of ethylene and propylene and the total yields of high value-added products. It can be seen that a minimal yield exists and the maximum point can be obtained when the mass fraction of ethane reaches 1.0. Furthermore, the two curves change gently, and the gap between minimum and maximum is within 3%. Thus, the co-cracking of ethane and propane mixtures has a negative effect on the profitability of the industrial furnace.

The product yield profiles with the mass fraction of LPG are shown in Figs. 3 and 4. Considering the actual production process, the mass fraction of LPG is defined between 0 and 0.5. As can be seen in Fig. 3, the ethylene and methane yields increase with increasing proportion of LPG, while the propylene yield decreases. This can be explained by

the easy cracking of LPG. LPG is lighter than naphtha and can be cracked easily to generate more small molecules which can promote the cracking of naphtha to obtain a higher ethylene yield. The gap between the minimum and maximum of ethylene yield can even reach 7%. Fig. 4 illustrates that the summarized yields of ethylene and propylene and the total yields of the high value-added products increase gradually, especially for ethylene and propylene, and the total yields increase by 5%. Consequently, LPG can have a positive effect on the cracking of naphtha and improve the profitability of the industrial furnace.

## 5.2. Effect of the coil outlet temperature

According to the industrial data, the coil outlet temperature varies from 830 °C to 870 °C for the co-cracking of ethane and propane mixtures. Figs. 5 and 6 show the product yield profiles with the coil outlet

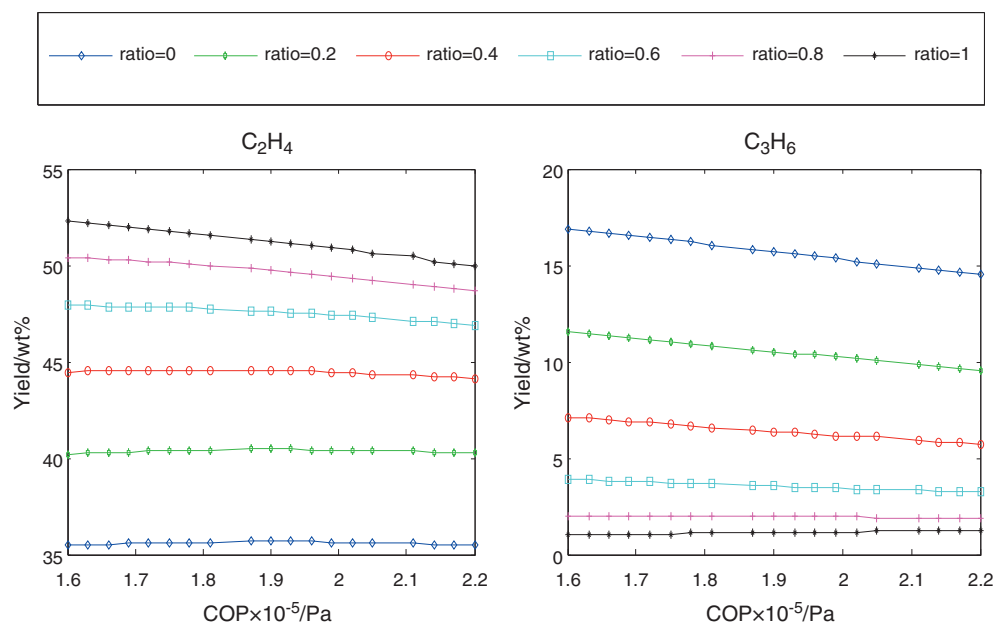


Fig. 9. COP effect on the yield of ethylene and propylene yields (C<sub>2</sub>/C<sub>3</sub>).

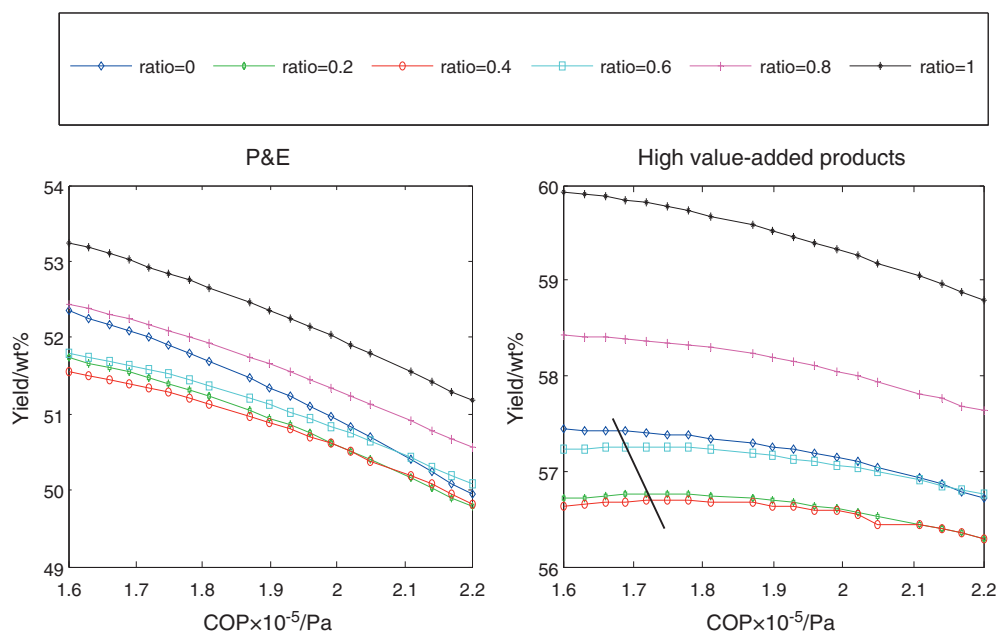


Fig. 10. COP effect on the yield of P&E and high value-added products (C<sub>2</sub>/C<sub>3</sub>).

temperature. In Fig. 5, the ethylene yields increase with increasing coil outlet temperature for different mass fractions of ethane. As the steam cracking is an endothermic reaction, a higher temperature can promote the cracking reaction and lead to a higher ethylene selectivity. However, the propylene yields decrease when the mass fraction of ethane is below 0.8. Moreover, the downtrend becomes more and more obvious with increasing mass fraction of propane. This can be explained by a high propylene concentration. When the mass fraction of propane is high, more propylene can be generated and a high temperature can also accelerate the decomposition of propylene. However, the propylene yields increase slowly when the mass fraction of ethane exceeds 0.8 since less propane generates less propylene, leading to a low propylene concentration. Also, a low propylene concentration will reduce the decomposition of propylene.

From Fig. 6 the optimum summarized yields of ethane and propane exist within the experimental temperature when the mass fraction of ethane is within 0.2. The smaller mass fraction of ethane, the lower coil outlet temperature of the optimum yields. The same goes for high value-added products but the two indices always increase when the proportion of ethane exceeds 0.2.

Figs. 7 and 8 give an overview of the product yield profile with the coil outlet temperature for the co-cracking of LPG and naphtha mixtures. As the same as the co-cracking of ethane and propane mixtures, the ethylene yields increase and the propylene yields decrease with increasing temperature for different mass fractions of LPG. In addition, the optimum yields can be obtained within the experimental temperature for the summarized yields of ethylene and propylene and the total yields of high value-added products. It can also be seen that the coil

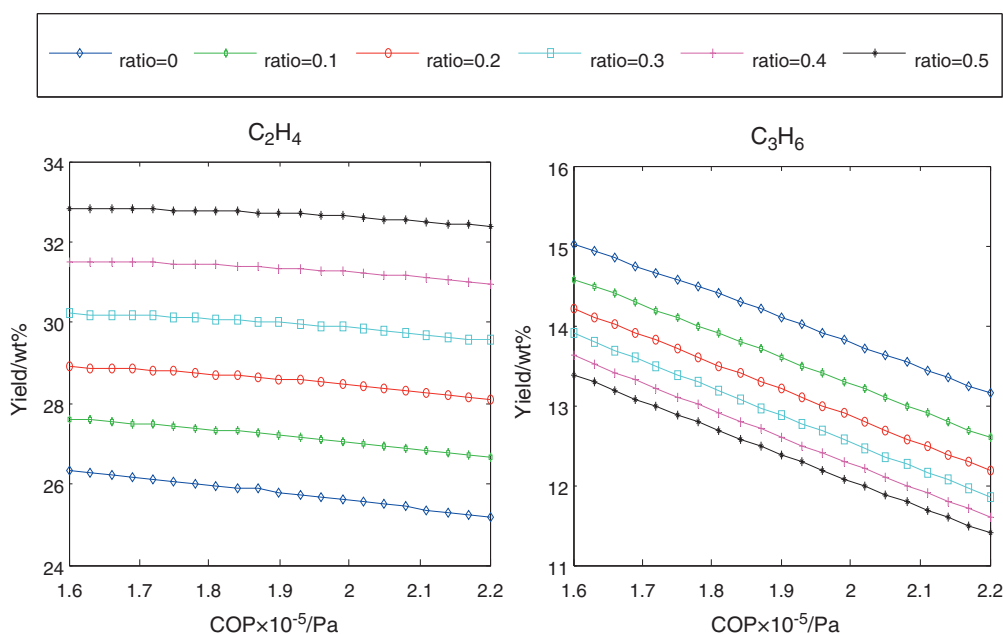


Fig. 11. COP effect on ethylene and propylene yields (LPG/NAP).

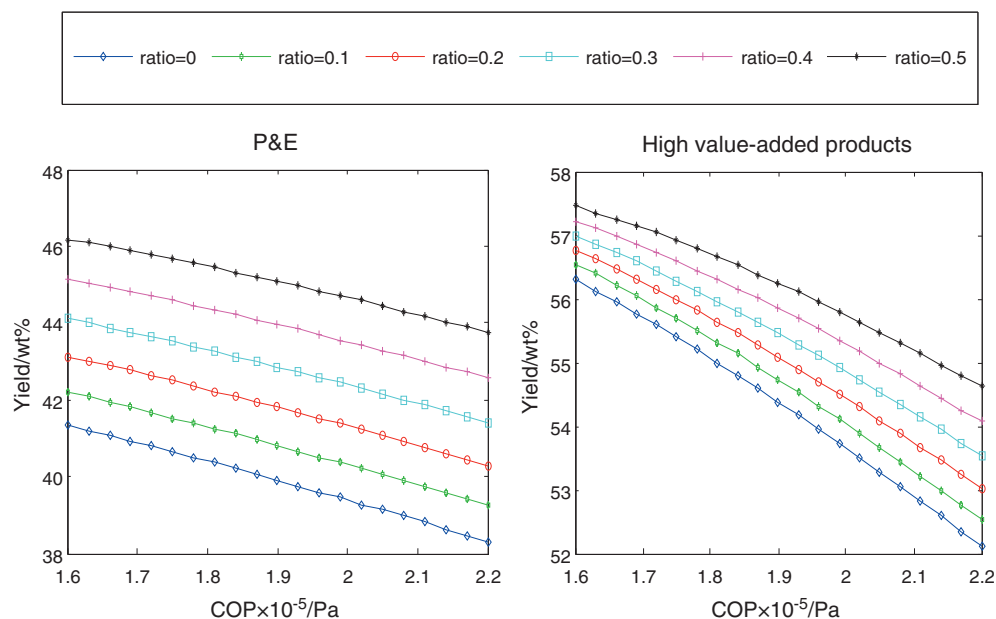


Fig. 12. COP effect on the yield of P&E and high value-added products(LPG/NAP).

outlet temperature at the point of the optimum yields rises with increasing mass fraction of LPG.

### 5.3. Effect of the coil outlet pressure

In Fig. 9, ethylene yield increases firstly and then decreases with increasing coil out pressure when the mass fraction of ethane is small, and an optimum yield exists. However, ethylene yield decreases when the mass fraction of ethane exceeds 0.5, and the change is small. For propylene, the yield decreases with the coil outlet temperature when the mass fraction of ethane is small, and the yield increases with the coil outlet temperature when the mass fraction of ethane becomes larger. Overall, the variation trend is slow and the coil outlet pressure has a little influence on ethylene and propylene yields.

In Fig. 10, the summarized yields of ethylene and propylene decrease with the coil outlet pressure for different mass fractions of ethane because of a rising number of molecules in the cracking reaction. Corresponding to Fig. 2, the summarized yields of ethylene and propylene decrease firstly and then increase with increasing mass fraction of ethane. For high value-added products, the total yields decrease when

the mass fraction of ethane is zero or exceeds 0.8, and when the mass fraction of ethane is between 0 and 0.8, an optimum value can be obtained.

Figs. 11 and 12 illustrate the product yield profile with the coil out pressure for the co-cracking of LPG and naphtha mixtures. As can be seen, the yields all decrease with the coil outlet pressure increasing for ethylene, propylene, P&E and high value-added products, and a high pressure can restrain the cracking reaction. However, the change of ethylene yield is small, while the yields of other products decrease dramatically.

### 5.4. Interaction effect of mixing ratio and coil outlet temperature

Figs. 13 and 14 give an overview of the interaction effect of the mixing ratio and the coil outlet temperature on the yields of P&E and high value-added products. As can be seen in Fig. 13, the biggest profitability is obtained for the co-cracking of ethane and propane mixtures when COT is at point of 870 °C and the mass fraction of ethane is 1. From Fig. 14, an optimum profitability exists within the experimental temperature when the mass fraction of LPG is the highest. Thus, the

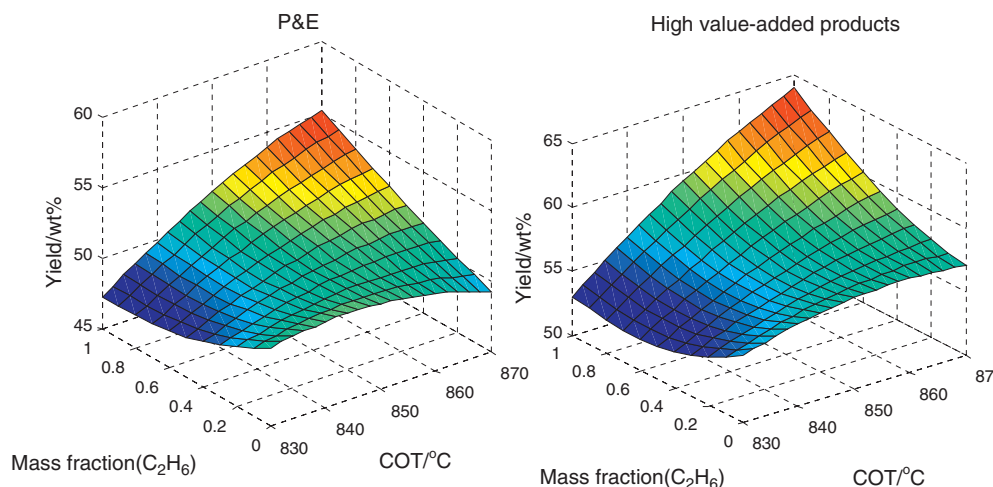


Fig. 13. Effect of mass fraction ( $C_2H_6$ ) and COT on the yields of P&E, high value-added products ( $C_2/C_3$ ).

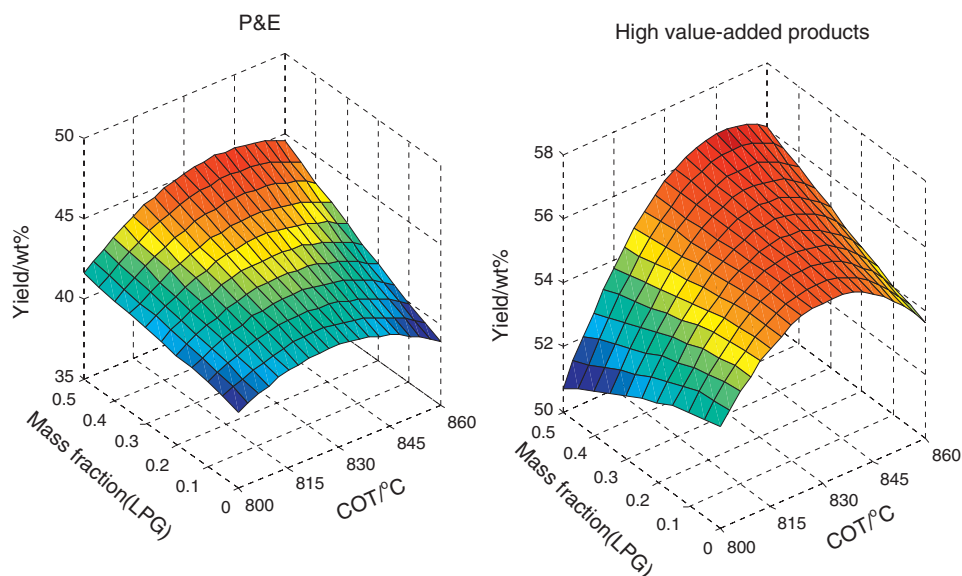


Fig. 14. Effect of mass fraction (LPG) and COT on the yields of P&E, high value-added products (LPG/NAP).

co-cracking of ethane and propane mixtures has a negative effect on the profitability of the steam pyrolysis furnace, however, LPG can improve the profitability of the cracking of naphtha.

#### 5.5. Interaction effect of mass ratio and coil outlet pressure

The interaction effect of the mixing ratio and the coil outlet pressure on the yields of P&E and high value-added products are shown in Figs. 15 and 16. As can be seen, a high coil outlet pressure leads to a low profitability. For the co-cracking of ethane and propane mixtures, the minimum yield exists with increasing mass fraction of ethane corresponding to Fig. 2. Consequently, the results can be summarized that the co-cracking of ethane and propane mixtures has a negative effect on the profitability of the industrial furnace, however, LPG can promote the profitability of the cracking of naphtha.

## 6. Conclusions

The simulations of the co-cracking of ethane and propane, and LPG and naphtha mixtures are carried out with Coilsim1D. The effects of

the mixing ratio, the coil outlet temperature and pressure on the yields of some key products are discussed. The following results can be concluded:

- The co-cracking of ethane and propane mixtures has a negative effect on the profitability of the industrial furnace. On the contrary, LPG can promote the generation of ethylene and decrease the propylene yield. In general, LPG has a positive effect on the profitability of the industrial furnace.
- In the co-cracking process of ethane and propane mixtures, the ethylene yield increases with increasing coil outlet temperature, while the propylene yield decreases when the mass fraction of ethane is below 0.8 and the propylene yields increase slowly when the mass fraction of ethane exceeds 0.8. The optimum summarized yields of ethane and propane can be obtained within the experimental temperature when the mass fraction of ethane is within 0.2. The smaller mass fraction of ethane, the lower coil outlet temperature of the optimum yields, and the same goes for high value-added products. For the co-cracking of LPG and naphtha mixtures, ethylene yield increases and propylene yield decreases with

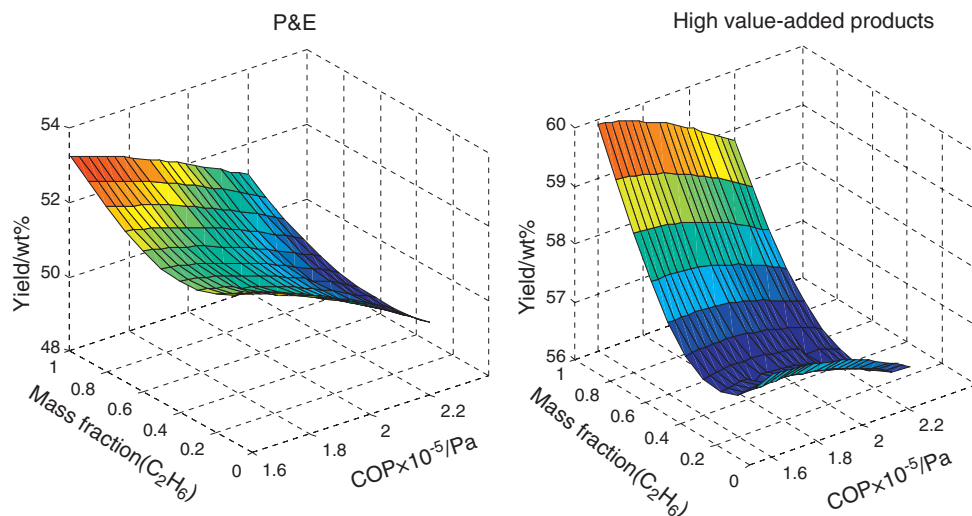


Fig. 15. Effect of mass fraction ( $C_2H_6$ ) and COP on the yields of P&E, high value-added products ( $C_2/C_3$ ).

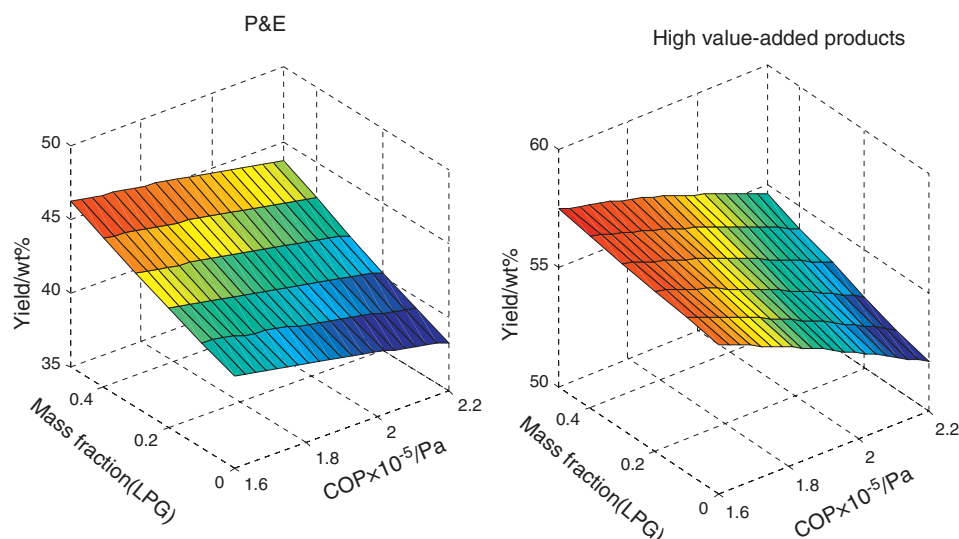


Fig. 16. Effect of mass fraction (LPG) and COP on the yields of P&E, high value-added products (LPG/NAP).

increasing coil outlet temperature, and an optimum yield exists within the experimental temperature for P&E and high value-added products.

- The most product yields decrease lightly with increasing coil outlet pressure in the co-cracking process of ethane and propane mixtures, and an optimum value exists for high value-added products when the mass fraction of propane is big enough. For the co-cracking of LPG and naphtha mixtures, the product yields all decrease with increasing coil outlet pressure, and the effect of the coil outlet pressure is apparent.

## Nomenclature

|                      |                                                                                                  |
|----------------------|--------------------------------------------------------------------------------------------------|
| $A_i, B_i, C_i, D_i$ | the coefficients of the heat capacity of pure species $i$                                        |
| $C_p$                | specific heat capacity of process gas, $\text{kJ} \cdot \text{kmol}^{-1} \cdot \text{K}^{-1}$    |
| $C_{p,i}$            | specific heat capacity of component $i$ , $\text{kJ} \cdot \text{kmol}^{-1} \cdot \text{K}^{-1}$ |
| $F_j$                | molar flow rate of component $j$ , $\text{kJ} \cdot \text{mol}^{-1}$                             |
| $f$                  | Fanning friction factor                                                                          |
| $n$                  | number of reactions                                                                              |
| $P_{ci}$             | the critical pressure of component $i$ , Pa                                                      |
| $p_t$                | the total pressure, Pa                                                                           |
| $q$                  | heat flux to the process gas, $\text{kJ} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$                |
| $R_{v,k}$            | net production rate for species $k$                                                              |
| $r_b$                | radius of the bend, m                                                                            |
| $r_{v,k}$            | reaction rate of reaction $k$ , $\text{kmol} \cdot \text{m}^{-3} \cdot \text{s}^{-1}$            |
| $T$                  | temperature of process gas, K                                                                    |
| $T_{ci}$             | the critical temperature of component $i$                                                        |
| $T_r$                | the ratio of the gas temperature and the critical temperature, K                                 |
| $V_{j,k}$            | stoichiometric coefficient of component $j$                                                      |
| $z$                  | axial position, m                                                                                |
| $\alpha$             | conversion factor                                                                                |
| $\Delta_f H_k$       | standard enthalpy of species $k$ , $\text{kJ} \cdot \text{kmol}^{-1}$                            |
| $\lambda$            | thermal conductivity of process gas, $\text{W} \cdot \text{m}^{-1} \cdot \text{K}^{-1}$          |
| $\lambda_i$          | thermal conductivity of component $i$ , $\text{W} \cdot \text{m}^{-1} \cdot \text{K}^{-1}$       |
| $\mu$                | viscosity of process gas, Pa·s                                                                   |
| $\mu_i$              | viscosity of component $i$ , Pa·s                                                                |
| $\rho$               | density of process gas, $\text{kg} \cdot \text{m}^{-3}$                                          |
| $\rho_i$             | density of component $i$ , $\text{kg} \cdot \text{m}^{-3}$                                       |
| $\phi_{ij}$          | the interaction parameter of component $i$ and component $j$                                     |
| $\Omega$             | cross sectional surface area, $\text{m}^2$                                                       |

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