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Process Control and Coordinating Optimisation for the Multi-Feed Demethaniser in Ethylene Complex

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In the chemical process, it is essential to control the product quality and optimal utilisation in limited energy. To achieve the goal of energy savings during the process control, three kinds of process control and coordinating optimisation strategies are proposed and compared in a two-input-two-output system. The first strategy contains both setting value control of inputs and optimisation of two set points to acquire the control of the objective output. The second strategy is realised by cascade control of the expected objective output and coordinating optimisation of the energy to acquire optimal set point of the setting value control of the other input. For the third one, the objective output and the other output are both controlled by cascade control, and the set point of the other output is optimised to achieve energy-saving goal. A case study on the multi-feed demethaniser in ethylene complex is employed to identify the efficiency of the process control and coordinating optimisation strategy. It turns out that the combination of the process control of the overhead composition and coordinating optimisation of reboil flow rate control is the most efficient on energy savings in proposed strategies during the improvement of the product quality.

1. Introduction

In the chemical process, the product quality of the distillation usually changes along with the requirements of the market, or it is controlled at constant values under disturbances. And the adjustment of corresponding set points of the temperature or composition controllers is efficient. Meanwhile, it is essential to achieve the optimal utilisation of the limited energy. Strong anti-interference ability and short adjustment period can ensure the robustness of the controller and product quality, which are significant for the design of the control system. Therefore, the investigation on the process control and coordinating optimisation strategy can not only guarantee the quality of the product, but also achieve the energy-saving goal during the life cycle.

The existing energy-saving measures on the distillation can be separated into several branches: heat integration of the distillation (Linnhoff et al., 1983) which has been regarded as the helpful method for the economic developments in energy, water and better utilisation of resources (Klemeš et al., 2013), modelling and optimisation of the distillation (Osuolale and Zhang, 2014) and more recently (Luo et al., 2015;), retrofit design to increase the production capacity (Tavan et al., 2016). But the investigations are limited in the fixed product quality, and authors seldom consider the changes of control objective and control strategies. Furthermore, the existing control strategies of the chemical process mainly fix on the design of the control configurations (Koggersbøl et al., 1996) and selection of the control methods. A comprehensive and cost-effective study on combination of the process control and coordinating optimisation for actual chemical process is still very lacking, which can not only quickly achieve the final objective output as required, but also realise the optimal utilisation of the system energy.

In this work, to study the process control and coordinating optimisation strategy, a simple two-input-two-output system is introduced and analysed in section 2. To identify the efficiency of the proposed method, a similar multi-feed demethaniser system is employed and discussed in section 3. Meanwhile, process control and coordinating optimisation strategies are conducted and compared to select the optimal method to realise energy savings during improving the product quality.

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2. Strategy analysis and methodology

In the distillation column, to achieve control of the temperature and key components at the overhead and bottom, the reflux and the bottom reboil flow rate are always selected as operating variables, respectively. Therefore, in terms of the control of the overhead and bottom, distillation column can be regarded as a two-input-two-output system. The coupling relationship usually exists between inputs and outputs, as shown in Figure 1 and relative equation of Eq(1).

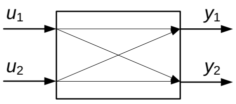

Figure 1: Two-input-two-output system

$$\begin{cases} y_1 = f_{S1}(u_1, u_2) \\ y_2 = f_{S2}(u_1, u_2) \end{cases} \tag{1}$$

Bristol(1966) proposed the steady relative gain array(RGA) method to analysis the interaction of variables in the MIMO system. Based on the analysis results of RGA, proper control strategy can be investigated. Hence, one can suppose the proper control strategy, in which inputs u_1 and u_2 are used to control outputs y_1 and y_2, respectively. And there are two kinds of control strategies to achieve the control objective, which are open-loop and close-loop control, as shown in Table 1.

Table 1: Control strategies for the two-input-two-output system

Controlled variable	Open-loop	Close-loop
y_1	Setting control of input u_1	Cascade control of output y_1
y_2	Setting control of input u_2	Cascade control of output y_2

In the distillation column, outputs y_1 and y_2 represent the mole fraction of the key component or temperature of the overhead and bottom, and inputs u_1 and u_2 represent the reflux and reboil flow rate of the bottom, respectively. Based on the different control requirements, there are two common control objectives, which are the setting value control to keep the y_1 constant against disturbances and changing the set point of y_1 to the new objective y_1^{obj}. While both of them can be treated as the case of controlling the objective output at required value. There are three strategies that can control output y_1 at required value y_1^{obj} and realise the optimal utilisation of the system energy.

Strategy 1: setting control of inputs u_1 and u_2, and optimisation of two controllers. Output y_1 is equal to the objective output y_1^{obj}, and optimisation of the system energy cost can achieve the optimal set points for inputs, u_1^{sp} and u_2^{sp}, by calculating the model Eq(1) and optimisation problem Eq(2).

$$\begin{cases} \min Q = f_1(u_1, u_2) \\ y_2 \in [y_{2,\min}, y_{2,\max}], y_1 = y_1^{\text{obj}} \\ u_{1,2} \in [u_{\min}, u_{\max}] \end{cases} \tag{2}$$

Strategy 2: the cascade control of output y_1, the setting control of the input u_2, and coordinating optimisation of the input u_2. The cascade control of the y_1 can easily achieve the objective output y_1^{obj} by the main-loop control. Output y_2 will arrive at a new output y_2', when the set point of setting controller u_2 is constant. Based on the cascade control of output y_1, if one can achieve the optimal set point of the input u_2^{sp}, the required output y_1^{obj} and optimal energy utilisation of the system can be realised. Therefore, one should select the optimal objective function of energy cost, operating variable input u_2 and constraints, such as the problem of Eq(3). When solving the Eqs(1) and (3), the optimal set point u_2^{sp} can be calculated.

$$\begin{cases} \min Q = f_2(u_2) \\ u_1 = f_{c1}(y_1, y_1^{\text{obj}}) \\ y_2 \in [y_{2,\min}, y_{2,\max}], y_1 = y_1^{\text{obj}} \\ u_{1,2} \in [u_{\min}, u_{\max}] \end{cases} \tag{3}$$

Strategy 3: cascade control of outputs y_1 and y_2, and coordinating optimisation of output y_2. Based on the cascade control of output y_1, y_1=y_1^{obj} can be achieved. The set point of the sub-loop controller is fixed by main-loop controller because the set point y_2 is constant. Under the cascade control of output y_1, if one can achieve the optimal output y_2^{sp}, the optimal inputs u_1^{opt} and u_2^{opt} can be calculated and the optimal utilisation of the energy will be realised. Therefore, after selecting the optimal objective function of energy cost, operating variable y_2, equations of controllers and constraints, the optimisation problem like Eq(4) can be proposed. And solving the problems containing Eq(1) and Eq(4) can get the optimal set point of y_2^{sp}.

$$\begin{cases} \min Q = f_3(y_2) \\ u_1 = f_{c1}(y_1, y_1^{obj}) \\ u_2 = f_{c2}(y_2) \\ y_2 \in [y_{2,\min}, y_{2,\max}], y_1 = y_1^{obj} \\ u_{1,2} \in [u_{\min}, u_{\max}] \end{cases} \tag{4}$$

Based on the above analysis of the control and optimisation strategies, it is convenient and meaningful to combine the process control and coordinating optimisation. Three kinds of process control and coordinating optimisation strategies named as Cases 1 to 3 can be proposed as shown in Table 2. Based on the differences of operating conditions in control and optimisation, the steady-state input-output relationship in each strategy can be developed in Figure 2. There are three routes, A→C, A→B→C and A→D→C. The coordinates of the operating points A-D are listed in Figure 2, in which points A and C are the primary point and optimal objective point, respectively. According to the requirement of product quality and optimisation results, outputs y_1 and y_2 vary from $y_{1,A}$ to y_1^{obj}, $y_{2,A}$ to y_2^{sp},respectively.

Table 2: Control and coordinating optimisation strategies for the two-input-two-output system

Label: route	Control and coordinating optimisation strategies
Case 1: A→C	1) optimisation of the u_1, u_2; 2) setting control of the optimal inputs u_1^{sp} and u_2^{sp}.
Case 2: A→B→C	1) cascade control of output y_1=y_1^{obj} and setting control of input u_2; 2) coordinating optimisation of the optimal input u_2^{sp}.
Case 3: A→D→C	1) cascade control of output y_1=y_1^{obj} and cascade control of output y_2; 2) coordinating optimisation of the optimal output y_2^{sp}.

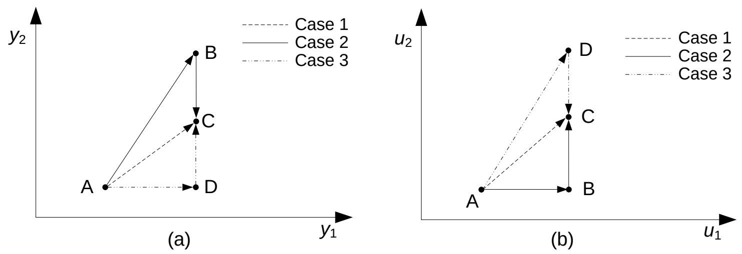

Figure 2: Diagrams of the input and output relationship; a) outputs y_1 and y_2; b) inputs u_1 and u_2

Coordinates of the operating points:

Points	*Coordinates (a)*	*Coordinates (b)*
A	($y_{1,A}$,$y_{2,A}$)	($u_{1,A}$,$u_{2,A}$)
B	(y_1^{obj},y_2')	($u_{1,B}$,$u_{2,A}$)
C	(y_1^{obj},y_2^{sp})	($u_{1,C}$,u_2^{sp})
D	(y_1^{obj},$y_{2,A}$)	($u_{1,D}$,$u_{2,D}$)

Case 1: route A→C, firstly, optimise the model and find out the optimal set points of the two open-loop setting controllers. Secondly, directly change the set points of the open-loop setting controllers of u_1 and u_2 from A($u_{1,A}$,$u_{2,A}$) to C(u_1^{sp},u_2^{sp}) as shown in Figure 2b. While exact values of u_1^{sp} and u_2^{sp} cannot be easily achieved in the chemical plant to achieve the objective output y_1^{obj}. Therefore, it is difficult to conduct this strategy, and the route A→C is impossible in actual chemical process.

Case 2: route A→B→C, the objective output y_1^{obj} is controlled by main-loop controller in the cascade control of output y_1, and input u_2 is controlled by setting controller at a constant value $u_{2,A}$. Based on the process control strategy, output y_1 can reach the objective value y_1^{obj}, and input u_1 arrives at input $u_{1,B}$. Because of the coupling of the system, even though the set point of the setting value control of u_2 is constant, output y_2 varies along with the change of the input u_1 to a new steady state output y_2'. The original output condition point A($y_{1,A}$,$y_{2,A}$) will arrive at a new output point B(y_1^{obj},y_2') in Figure 2a, where input u_2 is at a new point B($u_{1,B}$,$u_{2,A}$), but input u_2 is still at the old value $u_{2,B}$. To achieve the final optimisation of the energy utilisation of the plant, the coordinating optimisation of set point of the setting control of input u_2 should be conducted, based on which the optimal set point u_2^{sp} can be achieved. Because of the coupling, the cascade control of output y_1 will make input u_1 arrive at the new point $u_{1,C}$ along with the input controller's set point to u_2^{sp}. Output point B(y_1^{obj},y_2') will finally arrive at point C(y_1^{obj},y_2^{sp}), as output route B→C shown in Figure 2a. Input point B($u_{1,B}$,$u_{2,A}$) will turn to the point C($u_{1,C}$,u_2^{sp}), as shown in Figure 2b.

Case 3: route A→D→C, both controllers are selected at the cascade control mode. During the process control, the set point of output y_1 is switched to the objective y_1^{obj}, while the set point of output y_2 is unchanged, like the output route A→D which is from point A($y_{1,A}$,$y_{2,A}$) to point D(y_1^{obj},$y_{2,A}$), as shown in Figure 2a. When the set point of y_2 is optimised, the optimal output y_2^{sp} can be achieved and switched. While the set point of the old controller on output y_1 is unchanged. Under the control of the controllers, the inputs of the system will arrive at the optimal objective point C($u_{1,C}$,u_2^{sp}) from point D in Figure 2b. The route of the control and coordinating optimisation strategy is the curve A→D→C, as shown in Figure 2a. Compared with Case 1, Cases 2 and 3 can be easily realised and repeated in life cycle. Therefore, Cases 2 and 3 will be employed and mainly discussed in next Section.

3. Application on demethaniser system

3.1 Demethaniser description

Multi-feed demethaniser is a special unit in ethylene complex and is used to extract the methane and hydrogen from the ethylene cracking gas. Model of ethylene complex has been referred, and validity of the simulation has also been identified in our early work (Wu and Luo, 2016) in gPROMS software. The high temperature ethylene cracking gas transfers heat with the bottom reboiler to provide heat energy for the demethaniser. At the same time, the cracking gas flows through the cold box system and is divided into several streams in different energy levels, such as feed streams F1–F4 with different temperature and pressure. There are four feed streams in the multi-feed demethaniser, as shown in Figure 3.

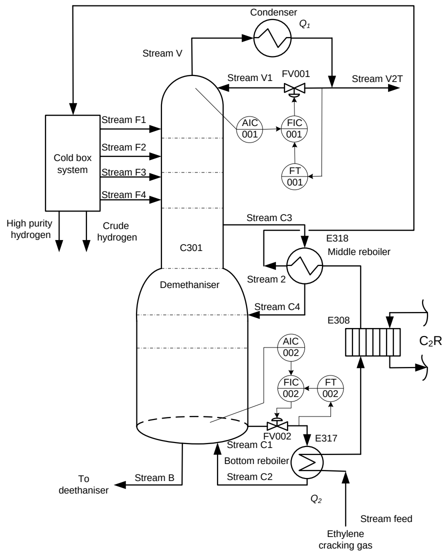

Figure 3: Flow sheet of the multi-feed demethaniser system

For the demethaniser system, both mole fraction of ethylene at the overhead and mole fraction of methane at the bottom should be taken into consideration during the operation of the demethaniser in the separation of the cracking gas. Considering that the temperature of demethaniser is the lowest among distillation columns in ethylene complex system, the methane contained in bottom product is difficult to be extracted in other columns, with the result that the amount of methane found in the final ethylene product will be high and the product quality will be affected. Therefore, to lower down the content of methane found in the bottom stream, more flow rates of stream C2 should be reboiled to gain more heat from the ethylene cracking gas. At the same time, the increase of the bottom heat will bring more hot vapour stream flowing into the overhead. To keep the content of ethylene at a low level, the reflux rate should be increased to provide more extra cooling capacity to lower the temperature of the overhead. Obviously, the coupling exists between the control of the overhead and bottom key components. For the distillation system, the reflux is usually selected to control the overhead temperature or component, and the bottom reboil flow rate is usually used to adjust the bottom temperature and mole fraction of the key component. According to the different production projects, the product quality usually changes as required. The demethaniser is very similar with the two-input-two-output system referred before, and can adopt the above process control and coordinating optimisation method to achieve the energy-saving goal.

3.2 Process control and coordinating optimisation strategy of demethaniser

To achieve the required overhead product quality and decrease the energy cost of the demethaniser system as much as possible, the process control of the overhead key component and coordinating optimisation of the bottom reboil flow rate are conducted in this section. The diagrams of cascade control strategies are used to control the overhead component of ethylene and bottom component of methane, as shown in Figure 3. Main-loop and sub-loop controllers are all listed in Table 3. If the cascade control strategy only includes sub-loop controller, it is called the setting value control of the flow rate, which is the open-loop control of the component.

Table 3: Cascade control strategies of key components of the demethaniser

Controlled variables	Main-loop controller	Sub-loop controller
Overhead component of ethylene	AIC001	FIC001
Bottom component of methane	AIC002	FIC002

Demethaniser is more special than common distillations, in which the coupling exists between the cooling capacity and bottom reboiler. To potentially decrease the temperature of the cracking gas, more bottom heat Q_2 should be absorbed. Meanwhile, the increase of the bottom heat will directly result in the cooling capacity of the condenser reaching a new energy balance. Therefore, to achieve the optimal energy-saving goal of demethaniser system, the cooling capacity Q_1 should be small enough and be selected as the objective function. Based on the above control strategies, required model equations and constraints, the control and coordinating optimisation strategies, Cases 2 and 3 can be conducted. After that, steady-state and dynamic adjustments of mole fractions of overhead ethylene and bottom methane, flow rate of reflux and reboil flow rate are shown in Figure 4.

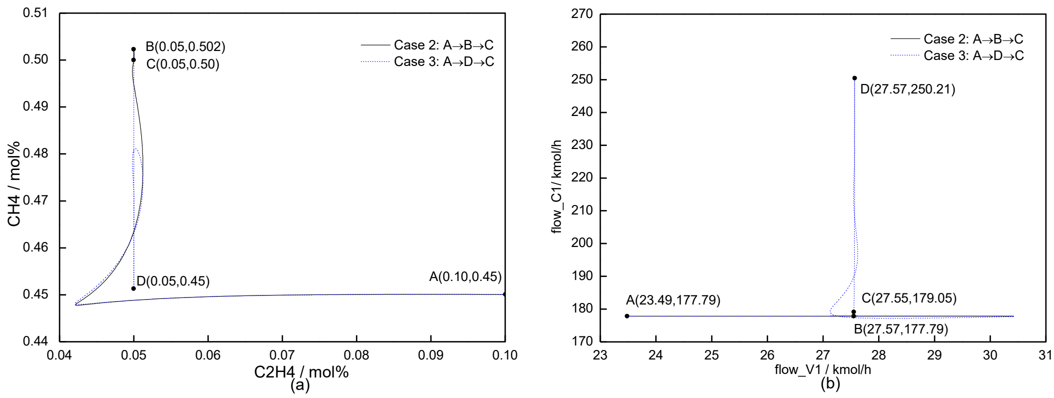

Figure 4: Dynamic relationships of the overhead and bottom key components and flow rates; a) overhead and bottom components of methane and ethylene; b) flow rates of Streams V1 and C1

For the Case 2 strategy, set point of the main-loop controller AIC001 is switched from 0.1mol% to 0.05mol%, and set point of the sub-loop controller FIC002 is unchanged. The mole fraction of ethylene is changed from point A to point B in Figure 4a, and the flow rate of the flow_V1 varies in the flow_V1 axis direction in Figure 4b. When the coordinating optimisation of the cooling capacity is employed, the set point of sub-loop controller FIC002 will reach the optimal point C. By comparison, Case 3 strategy is based on the cascade control of overhead and bottom, such as AIC001 and AIC002. The control of main-loop will drive the overhead and bottom component to the point D from point A. Coordinating optimisation of the cooling capacity can acquire the optimal set point of the bottom main-loop controller AIC002, and the optimal objective value of bottom component of methane can be achieved by process control, which is from point D to C, as shown in Figure 4a.

Figure 5 shows the dynamic changes of mole fraction of ethylene and methane, and flow rate of flow_V1 and flow_C1 in two cases. The controller can realise the control of overhead ethylene in two cases. With the different control strategies at the bottom, changes of the flow rate of flow_C1 and mole fraction of methane are very different, as shown in Figure 5a. During the process control, flow rate of flow_C1 is unchanged in Case 2, and the mole fraction of methane at the bottom is controlled and unchanged in Case 3. After the implementation of the process control and coordinating optimisation, the cooling capacity Q_1 and heat Q_2 in Case 2 and Case 3 reach the same steady state, respectively, as shown in Figure 5b. Compared with single control strategy in improving product quality of key component at the overhead, the increments of cooling capacity Q_1 and heat Q_2 achieved by Cases 2 and 3 decrease 9.4 % and 37.6 %. For the Case 2 and Case 3, both of them can realise the control objective and energy savings. However, compared with Case 3, increments of gross energy of Q_1 and Q_2 in Case 2 decrease 4.2 % and 29.1 % in 40 h. During the adjustment of improving the product quality at overhead, strategy Case 2 is more efficient than Case 3.
When a sinusoidal disturbance, 0.5·sin(6.28·t), is added to the temperature of the stream F1, mole fractions of ethylene and methane, reflux and reboil flow rate, cooling capacity Q_1 and heat Q_2 for two cases are shown in Figure 5. Considering the disturbance, Case 2 is at a better disturbance attenuation level than Case 3 strategy.

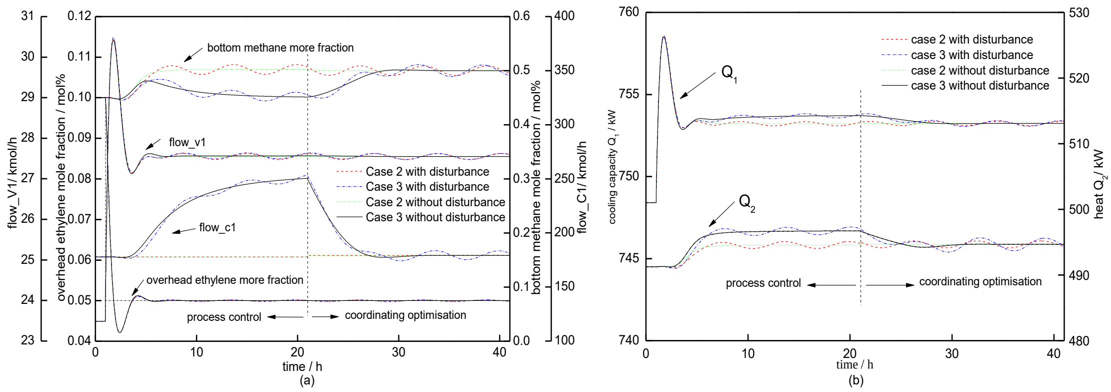

Figure 5: Changes of the demethaniser with and without disturbance; a) key mole fractions and flow rates at overhead and bottom; b) cooling capacity Q_1 and heat Q_2

4. Conclusions

This study shows that the control and coordinating optimisation strategy is efficient on the multi-feed demethaniser in ethylene complex to achieve the overhead product quality control and energy-saving goal. During the process of improving the product quality, the process control of the objective composition at the overhead and coordinating optimisation of setting value control of flow rate at the bottom reboiler is more efficient on energy savings than other strategies.

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