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DESIGN OF CONTROL STRUCTURES OF CUMENE PROCESS

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จิตติมา ทิพย์สุขุม : การออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์ของกระบวนการผลิตคิวมีน. (DESIGN OF CONTROL STRUCTURES OF CUMENE PROCESS) อ. ที่ปรึกษาวิทยานิพนธ์หลัก: ผศ. ดร.มนตรี วงศ์ศรี, 152 หน้า

ปัจจุบันนี้การแข่งขันทางด้านอุตสาหกรรมเกิดขึ้นอย่างมากมาย ดังนั้นจึงต้องมีการควบคุมคุณภาพของสินค้าให้ได้มาตรฐาน นอกจากนั้นแล้วยังต้องมีการควมคุมต้นทุนในการผลิตให้อยู่ในระดับที่ต่ำ จึงได้มีการนำสารตั้งต้นและพลังงานกลับมาใช้ภายในกระบวนการอีกครั้ง ซึ่งจะส่งผลให้โครงสร้างของกระบวนการมีความซับซ้อนมากยิ่งขึ้น ดังนั้นการควบคุมการะบวนการโดยรวมให้มีสมรรถนะที่ดีจึงจัดว่าเป็นสิ่งที่จำเป็น และมีความสำคัญในการดำเนินกระบวนการผลิตให้เป็นไปตามวัตถุประสงค์ ทั้งในด้านความปลอดภัยและด้านคุณภาพของผลิตภัณฑ์ นอกจากนี้ยังนำมาซึ่งการใช้สารตั้งต้นและการใช้พลังงานอย่างคุ้มค่าอีกด้วย ซึ่งงานวิจัยนี้มีความสนใจที่จะนำกระบวนการผลิตคิวมีน ซึ่งจะประกอบไปด้วยสายป้อนกลับและมีการแลกเปลี่ยนพลังงานระหว่างกันภายในกระบวนการ มาศึกษาการออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์ตามขั้นตอนของวงศ์ศรี(2009) โดยในงานวิจัยนี้ใช้โปรแกรมไฮซีสเพื่อจำลองกระบวนการผลิตคิวมีนทั้งที่สภาวะคงตัวและสภาวะพลวัต จากนั้นทำการออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์สำหรับกระบวนการคิวมีนทั้งหมด 5 โครงสร้าง และประเมินสมรรถนะของโครงสร้างการควบคุมที่ออกแบบเปรียบเทียบกับโครงสร้างการควบคุมพื้นฐานของลูเบน (2010) ด้วยตัวรบกวนกระบวนการ 2 ชนิดคือ การรบกวนอัตราการไหลของสารและการรบกวนทางความร้อน ผลที่ได้พบว่าโครงสร้างการควบคุมที่ออกแบบมีสมรรถนะที่ดีเมื่อเปรียบเทียบกับโครงสร้างการควบคุมที่ออกแบบโดยลูเบน สามารถกำจัดตัวรบกวนที่เข้าสู่ระบบและสามารถรักษาคุณภาพของผลิตภัณฑ์ได้ โดยเปรียบเทียบจากปริพันธ์ของค่าคลาดเคลื่อนสัมบูรณ์และพลังงานที่ใช้ทั้งหมดน้อยกว่า ดังนั้นจึงสรุปได้ว่าการออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์โดยใช้ขั้นตอนของวงศ์ศรี (2009) สามารถออกแบบโครงสร้างการควบคุมที่มีสมรรถนะที่ดีได้

ภาควิชา วิศวกรรมเคมี ลายมือชื่อนิสิต
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Nowadays, industries are very competitive. Therefore, the quality of products must be control according to industry standards. Moreover in order to control the unit cost of production to be low, so material and energy recycle are beneficial in economic aspect. They are the causes that make the process even more complicated, so the good control structure is necessary and is the important things in order to maintain process operation, safety and product quality. This research will focus on the control structure designed of cumene process consisting of recycle stream and energy integration by using procedure of Wongsri (2009). In this study, HYSYS is used to simulate cumene process at steady state and dynamic. After that we design new plantwide control structure for cumene process and evaluated the dynamic performance compare by integral absolute error (IAE) with base case control structure (Luyben 2010). Two types of disturbances: material and thermal disturbances are used to evaluate the performance. The result shows that new design control structures have comparable performance with the base case since they can handle entering disturbance of the process and can maintain product quality.

Department : Chemical Engineering	Student's Signature ....
Field of Study : Chemical Engineering	Advisor's Signature ....
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CONTENTS

PAGE

PAGE

PAGE

LIST OF TABLES

PAGE

LIST OF FIGURES

PAGE

PAGE

PAGE

CHAPTER I

INTRODUCTION

1.1 Importance and Reasons of Research

Nowadays, industries are very competitive both in quality and cost of production. Therefore, production process should have high quality and high efficiency. The process should always operate under the design condition, use little energy, low waste production and meet required specification of the products. In the real situation, the process will not operate smoothly. All factors do not meet the designed condition. The process always changes due to disturbance from the external factors and the internal factors. However, no matter what factors cause the change, in case of having deviation or disturbance come in to the process, the effect should be eliminate from the process as soon as possible process will have the least deviation from the designed condition. The appropriate choice of control structure among many possible control structures is important tasks in control system. In the past, Umeda, Kuriyama & Ichidawa (1987) presented the unit-based control system design methodology for control complete plants. The effect of recycles impacts on this method. Later Luyben, Tyreus and Luyben (1997) presented a general heuristic design procedure. Their procedure generated an effective plantwide control structure for an entire complex process flowsheet and not simply individual units. The nine step of the proposed procedure center around the fundamental principles of plantwide control: energy management, production rate, product quality, operational, environmental and safety constraints, liquid level and gas-pressure inventories, makeup of reactants, component balances and economic or process optimization.

Cumene can be found in crude oil and is a part of processed high-octane gasoline. Cumene is an important intermediate for industrial products such as phenolic resins, epoxy, nylon-6 and polycarbonate resins etc. Generally, processes for cumene manufacture consist of a packed bed reactor (liquid or vapor phase) followed by a separation train that removes the light inerts (propane), recycles the unreacted benzene and separates the cumene product from heavies generated by further alkylation of cumene to diisopropylbenzene. The heavies are sent to a transalkylation reactor for reactions with benzene to recover cumene. In this research, it will focus on new plantwide control design procedure of Wongsri (2009) applied to cumene process. This study will design plantwide control structures of cumene process and simulate them by using HYSYS simulator to study about dynamic behavior and evaluate the performance of the designed structures. An effective designed structure can achieve the control objective to reduce the cost of production and operate the process within safety and environment constraint.

1.2 Objective of research

The objectives of this research are:

To validate plantwide control structures of cumene process by using Wongsri's procedure.

1.3 Scope of Research

The scope of this research can be listed below.

1.3.1 Simulation of the cumene process is performed by using HYSYS simulator.

1.3.2 Description of cumene process is given by William L. Luyben, 2010.

1.3.3 Plantwide control structures for cumene process are designed using new design procedure of Wongsri (2009).

1.3.3 Five control structures for cumene process are designed by Wongsri procedure.

1.4 Contribution of Research

The contribution of this research can be listed below:

1.4.1 Steady state and dynamic model have been simulated using by HYSYS simulator.

1.4.2 The new plantwide control structures of cumene process are designed by using new design procedure of Wongsri (2009).

1.4.3 The dynamic performance of the new plantwide control structures are compared with the work given by William L. Luyben, 2010.

1.4.4 Analysis of new control structures design procedure.

1.5 Research Procedure

Research Procedure of this research is as follows:

1.5.1 Study of plantwide process control theory.

1.5.2 Study of cumene process and concerned information.

1.5.3 Simulation of cumene process at steady state.

1.5.4 Simulation of cumene process at dynamic.

1.5.5 Study the new design procedure.

1.5.6 Design new plantwide control structures by using the new design procedure.

1.5.7 Evaluate the dynamic performance of the new designed control structures.

1.5.8 Collection and summarization of simulation result.

1.6 Research Contents

This thesis is divided into six chapters:

Chapter I is an introduction to this research. This chapter consists of importance and reasons for research, research objective, scopes of research, contributions of research and research procedures.

Chapter II reviews the work carried out on plantwide control, heat exchanger networks design, heat integrated processes and method of selection set of controlled variables.

Chapter III cover some background information of Luyben's theory concerning with plantwide control and new plantwide control structure design procedure of Wongsri (2009).

Chapter IV describes the process description and the design of heat exchanger networks for cumene process and process simulation via HYSYS.

Chapter V describes the design of plantwide control structures and dynamic simulation results.

Chapter VI presents the conclusion of this research and makes the recommendation for future work.

This is follow by:

References

Appendix A: Process stream data and equipment data

Appendix B: Parameter tuning of control structures

Appendix C: Fixture point theorem data

CHAPTER II

LITERATURE REVIEW

2.1 Plantwide control design

Luyben (1996) presented the problems in the development of the steady - state by finding the number of variables that will be required to complete the process. This number is called a design degree of freedom. However, for complex industrial processes often contain hundreds of variables and equations, and it is not a trivial task to ensure that the correct variables and equations have been defined.

Luyben, Tyreus and Luyben (1997) presented a problem in the design process. Nine steps of the proposed procedure center around the basic principles of control plantwide: energy management, production rate, quality of products, operational, environmental and safety constraints, fluid levels and inventories of gas pressure, makeup of reactants, component balances and economic or process optimization. This procedure was showed with three examples of the industry: vinyl acetate monomer process, Eastman process, and the HDA process. The procedure produced a workable plantwide control strategy for a given process design.

Kasemchainun (2003) presented control strategies for the design plantwide control structure of vinyl acetate monomer plant to achieve the objectives of this process can be implemented within the constraints of safety, environmental and operation. Three alternative design control structure, tested and compared the performance with Luyben's control structure. All of controls structures can operate within constraint achieve the objectives and a good control structure was quickly response to disturbance.

Dimian and Bildea (2008) present the conceptual design of chemical processes illustrated by case studies worked out by computer simulation. In addition, a preliminary design allows critical aspect in research and development and in researching subcontractors to be highlighted, well ahead of starting the actual plant design project. For alkylation of benzene by propylene to cumene illustrates the design of a modern process for a petrochemical commodity. Significant energy saving can be achieved by heat integration by using double-effect distillation and recovering the reaction heat as medium-pressure steam.

Suntisrikomol (2008) study of “Fixture Point Theorem” for hydridealkylation process (HDA) to select appropriate the set of controlled variables from a large number of candidate output. The fixture point control theorem state that the most disturbed points must be satisfactorily controlled by giving them consideration before other controlled variables. The maximum (scaled) gain is used to selecting and paring controlled variables with manipulated variables. To illustrate the dynamic behaviors of the control structures when the load disturbance occurs, the performance of the controlled structures were presented in the IAE value, and compared to the reference structure. The design structures were fast response and the most effective on compared with Luyben, 1998.

Detjareansri (2009) has proposed a control structures for the alkylation process using new design procedure of Wongsri. The designed control structures were evaluated the dynamic performance compared with Luyben, 2002.She has been designed for eight. Material flow and thermal disturbance were used for this case to evaluate the effectiveness of control structures. All the designed control structures have a good performance because it handle disturbances entering the process and to maintain product quality when compared by integral absolute error (IAE).

Luyben (2010) proposed the design and control of the cumene process. The purpose of this work was to use the cumene process to illustrate some interesting design optimization features and principles. The two dominant design optimization variables are reactor size and benzene recycle. Increasing either one reduces the amount of undesirable byproduct that is produced, but increasing either of these variables raises capital and energy costs. The result could be concluded that to optimize the design variables affect the capital and energy costs.The economics of the cumene process illustrate that reducing reactant losses and the production of undesirable products can produce savings that were much larger than capital or energy costs.

2.2 Heat Exchanger Networks (HENs)

Linhoff and Hindmarsh (1983) presented a novel method for the design of HEN. The method is the first to combine sufficient simplicity to be used by hand with near certainty to identify best designs, even for large problems. Best design features the highest degree of energy recovery possible with a given number of capital items. Moreover, they feature network patterns required for good controllability, plant layout, intrinsic safety, etc. Hen designed to improve the work related to the position of the process and utility heat exchangers for heating and cooling of the manufacturing process specified in the target temperature.

Linnhoff and Kotjabasakis (1984) developed a design procedure for operable HENs by inspection and using the concept of downstream paths, i.e. the paths that connect the disturbed variables downstream to the controlled variables. They generated HEN design alternatives by the pinch method for the nominal operating condition. Then, the alternative designs are inspected for the effects of disturbances on the controlled variables and they are removed by breaking the troublesome downstream paths. Path breaking can be done by relocating and/or removing exchangers. If this procedure is not feasible, control action is inserted into the structure.

Saboo and Morari (1984) classified flexible HENs into two classes according to the kind and magnitude of disturbances that effect the pinch location. For the temperature variation, they show that if the MER can be expressed explicitly as a function of stream supply and target conditions the problem belongs to Class I, i.e. the case that small variations in inlet temperatures do not affect the pinch temperature location. If an explicit function for the minimum utility requirement valid over the whole disturbance range does not exist, the problem is of Class II, i.e. the case that large changes in inlet temperature of flowrate variations cause the discrete changes in pinch temperature locations.

Floudas and Grossmann (1987) presented a synthesis procedure for resilient HENs. Their multiperiod operation transshipment model is used to find a match structure for selected design points.The design has been possible in that match.If it is not feasible, the critical point is added as an additional operating point and the problem is reformulated and solved. If the match network is feasible then the multiperiod superstructure is derived and formulated as an NLP problem to find a minimum unit solution.

Calandranis and Stephanopoulos (1988) proposed a new approach to solve the following problem: design the configuration of control loops in a network of heat exchangers and sequence the control action of the loops, to accommodate set point changes and reject load disturbances. The approach proposed exploits the structure characteristics of a HEN by identifying routes through the HEN structure that can allocate load to available sinks. They also discussed several design issues such as the placement of bypass lines and the restrictions imposed by the existence of a process pinch. An online, real-time planning of control actions was the essence of implementation strategies generated by an expert controller, which selects path through the HEN was to be used for each entering disturbance to carry the associated load to a utility unit.

Wongsri (1990) studied a resilient HENs design. He presented a simple but effective systematic synthesis procedure for the design of resilient HEN. Design of the heuristic was used in the design or synthesizes HENs with pre-specified resiliency. The design must not only feature minimum cost, but must also be able cope with fluctuation or changers in operating conditions. A resilient HEN synthesis procedure was developed based on the match pattern design and a physical understanding of the disturbances propagation concept. The disturbance load propagation technique was developed from the shift approach and was used in a systematic synthesis method.

CHAPTER III

THOERY OF PLANWIDE CONTROL

This chapter is aimed to summarize heuristic approach from the previous researches and this approach in heat pathway view point which was developed by Wongsri and Hermawan (2005). Furthermore, we propose the plantwide control involving the system and strategies required to control entire plant consisting of many interconnected unit operations.

3.1 Basic Knowledge for Pinch Technology

3.1.1 Pince Technology

Pinch technology has been developed for more than two decades and now provided a systematic methodology for analysis chemical processes and surrounding utility system. The concept was first developed by two independent research groups (Flower and Linnhoff, 1978; Umeda et al., 1979), based on an applied thermodynamics point of view.

3.1.2 Basic Pinch Analysis Concept

The pinch analysis concept is originated to design the heat recovery in network for a specified design task. Starting with do calculate heat and material balance of the process obtained after the core process has been design. By using thermal data from the process, we can set the target for energy saving prior to the design of the heat exchanger networks. The necessary thermal data is source, target temperature and heat capacity flowrate for each stream as show in Table 3.1.

Table 3.1 Thermal data for process streams (Linnhoff and hindmarsh, 1983)

Stream No	Stream type	Start Temperture (Ts), °C	Target Temperture (Tt), °C	Heat capacity flowrate (CP), kW/°C
1	Hot	150	60	2
2	Hot	90	60	8
3	Cold	20	125	2.5
4	Cold	25	100	3

Here the hot streams are referred to the stream that required cooling, i.e. the source temperature is higher than that of the target. While the cold stream are referred to those required heating, i.e. the target temperature is higher than the supply. Heat capacity flowrate is defined as the multiple between specific heat capacity and mass flowrate as show below:

$$CP = C_p*F \tag{3.1}$$

where:

CP = heat capacity flowrate (kW/C)

C_p = specific heat capacity of the stream (kJ/C.kg)

F = mass flowrate of stream (kg/s)

The data used here is base on the assumption that the heat capacity flowrate is constant. In practice, this assumption is valid because every stream with or without phase change can easily be described in terms of linearization temperature-enthalpy data (i.e. CP is constant). The location of pinch and the minimum utility requirement can be calculated by using the problem table algorithm (Linnhoff and Flower, 1979) for a specified minimum temperature different, ΔT_{min}. In the case of ΔT_{min} = 20 °C, the results obtained from this method are show in table 3.2.

Table 3.2 the problem table for data given in Table 3.1

W				T hot (°C)	T cold (°C)	∑W (kW/°C)	ΔT	Required Heat (kW)	Interval (kW)	Cascade Heat (kW)	Sum Interval (kW)
H1	H2	C1	C2								
0	0	0	0	150	130	0		Qh		-105	
2	0	0	0	145	125	2	5	108	10	2.5	10
2	0	2.5	0	120	100	-0.5	25	118	-12.5	12.5	-2.5
2	0	2.5	3	90	70	-3.5	30	105	-105	0	-108
2	8	2.5	3	60	40	4.5	30	0	135	-105	27.5
0	0	2.5	3	45	25	-5.5	15	135	-82.5	30	-55
0	0	2.5	0	40	20	-2.5	5	52.5	-12.5	-52	-67.5
										Qc	

The pinch separates the problem into 2 thermodynamic regions, namely, hot end and cold end. The hot end is the region comprising all streams or part of stream above the pinch temperature. Only hot utility is required in this region but not cold utility. In contrast to the hot end, the cold end is the region comprising all streams or part of stream below the pinch temperature and only cold utility is instead desired regardless the hot utility. It is important to note that there is no heat transfer across the pinch therefore the minimum utility requirement is achieved.

Additionally, Saboo and Morari (1984) classified flexible HENs into two classes according to the kind and magnitude of disturbances that affect the pinch location. For the temperature variation, they show that if the MER can be expressed explicitly as a function of the stream supply and target conditions the problem belongs to Class I, i.e. the case where small variations in inlet temperatures do not affect the pinch temperature location. If the explicit function for the minimum utility requirement valid over the whole disturbance range dose not exists, the problem is of Class II, i.e. the case where large changes in inlet temperatures or flowrate variations cause the discrete changes in pinch temperature locations.

3.2 Heat Exchanger Networks

It is generally accepted that an optimal network must feature a minimum number of units that reflects on a capital cost and minimum utility consumption that reflects on operating costs. A good engineering design must exhibit minimum capital and operating costs. For Heat Exchanger Network (HEN) synthesis, other features that are usually considered in design are operability, reliability, safety, etc. in recent years the attention in HEN synthesis has been focused on the operability features of a HEN, e.g. the ability of a HEN to tolerate unwanted changes in operating conditions. It has been learned that considering only a cost objective in synthesis may lead to a worse network, i.e. a minimum cost network may not be operable at some neighboring operating conditions. The design must not only feature minimum cost, but also be able cope with a fluctuation or changes in operating conditions. The ability of a HEN to tolerate unwanted changes is called resiliency. It should be note that the ability of a HEN to tolerate wanted changes is called *flexibility*.

3.2.2 Heuristics for HEN Synthesis

Several HEN matching rules with minimum energy and investment costs have been presented (Masso and Rudd, 1969, Ponton and Donalson, 1974 Rathore and Powers, 1975, Linnhoff and Hindmarsh, 1983, Jezowski and Hahne1986, Huang, Metha and fan, 1988, etc.),

The following are heuristics from the literature classified according to the design criteria

The heuristics to minimize the capital cost (the number of heat exchangers):

Heuristic C1. To generate a network featuring the minimum number of heat exchanger units, let each match eliminate at least one of the two streams; a tick-off rule (Hohmann, 1971).

Heuristic C2. Prefer the matches that will leave a residual stream at its cold end for a heating problem, or its hot end for a cooling problem. A match of this type will feature the maximum temperature difference.

Heuristic C3. Prefer matching large heat load streams together. The significance of this rule is that the control problem (a capital cost) of a mach of this type(whether it is implemented by one or many heat exchangers) should be less than that of heating or cooling a large stream with many small streams.

The heuristics to minimize the energy cost (the minimum utility requirement):

Heuristic E1. Divide the problem at the pinch into subproblems, one a heat sink (heating subproblem or hot end problem) and the other a heat source (cooling subproblem or cold end problem), and solve them separately (Linnhoff and Hindmarsh, 1983).

Heuristic E2. Do not transfer heat across the pinch.

Heuristic E3. Do not cool above the pinch.

Heuristic E4. Do not heat below the pinch.

The laws of thermodynamics:

Heuristic T1. In a heating problem, if a supply temperature of a cold stream is less than a target temperature of a hot stream by T_{min} or more and the heat capacity flowrate of a hot stream is less than or equal to the heat capacity flowrate of a cold stream, the match between these two streams is feasible. (Immediately above the pinch temperature, the heat capacity flowrate of a cold stream must be greater than or equal to that of a hot stream.)

Heuristic T2. In a cooling problem, if a supply temperature of a hot stream is greater than a target temperature of a cold stream by T_{min} or more and the heat capacity flowrate of a hot stream is greater than equal to the heat capacity flowrate of a cold stream, the match between these two streams is feasible. (Immediately below the pinch temperature, the heat capacity flowrate of a hot stream must be greater than or equal to that of a cold stream.)

Heuristic T3. For a situation different from the above rule, match feasibility must be determined by checking whether the minimum temperature difference of a match violates the minimum approach, T_{min}, specific by the design.

3.2.3 Math Classification

In order to make use of the heuristics we must classify matches. The following criteria are considered important in this research:

1. Position of a Match. One heuristic prefers a match at the cold end and another prefers a match at the hot end. Pinch heuristics prefers a match at the cold end in a heating subproblem and a match at the hot end in a cooling subproblem. However, there are other possibilities. By using the tick-off heuristic, there are four ways that two streams can match. This leads to the basic four match patterns (Wongsri, 1990).

2. Heat capacity flowrate (between hot and cold stream). See Heuristic T.1 and T.2.

3. Heat Load (between hot and cold streams). The heuristic that concerns heat load state that one must match large heat load hot and cold streams first. This leads to two additional heuristic:

Heuristic N1. For a heating subproblem, a match where the heat load of a cold stream is greater than that of a hot stream should be given higher priority than the other .The reason is that the net heat load heating subproblem is in deficit. The sum of heat loads of cold streams is greater than of hot streams. The purposed match will likely be part of a solution (Wongsri, 1990).

Heuristic N2. Conversely, we prefer a mach where the heat load of a hot stream is greater than that of a cold in a cooling subproblem (Wongsri, 1990).

4. Residual Heat Load. No heuristics for this quantity have thus far appeared in the literature. Two new heuristics are introduced.

For a match in a heating subproblem that satisfies the heat load preference heuristics N.1;

Heuristic N3. We prefer a match where the residual heat load is less than or equal to the minimum heating requirement (Wongsri, 1990).

For a match in a cooling subproblem that satisfies the heat load preference or heuristics N.2:

Heuristic N4. We prefer a match where the residual heat load is less than or equal to the minimum cooling requirement, (Wongsri, 1990).

The reason behind the above two heuristics N3 and N4 is that the residual may be matched to a utility stream. One has the possibility of eliminating two streams at once.

3.2.4 Match Patterns

HEN synthesis is usually considered as a combinatorial matching problem. For a HEN in which a design property is regarded as a network property, or a structural property, we need to look beyond the match level to a higher level where such a property exists, e.g. to a match structure or match pattern. Match patterns are the descriptions of the match configuration of two, and possibly more, process streams and their properties that are thermally connected with heat exchangers. Not only the match description, e.g. heat duty of an exchanger and inlet and outlet temperatures is required but also the position of a match, e.g. upstream or downstream, the magnitude of the residual heat load and the heat capacity flowrates between a pair of matched streams.

By using the tick off rule there are four match patterns for a pair of hot and cold streams according to the match position and the length (heat load) of streams. The four patterns are considered to the basic match pattern classes. The members of these classes are the patterns where other configurations and properties are specified. The four match pattern classes are simply called A, B, C and D and are shown in Figure3.4, 3.5, 3.6 and 3.7 respectively. Any eligible match must belong to one of the four match pattern classes.

Definition 3.1 Class A Match Pattern: The heat load of a cold stream is greater than the heat load of a hot stream in a pattern, i.e. the hot stream is totally serviced. The match is positioned at the cold end of the cold stream. The residual heat load is on the hot portion of the cold stream. (See Figure 3.1)

A match of this class is a first type match at cold end position and the heat load of the cold stream is greater than that of the hot stream. This is a upstream match. For a heating subproblem, a Class A match is favored, because it leaves a cold process stream at the hot end (Heuristic N1) and follows the pinch heuristics. (See Table3.3)

Definition 3.2 Class B Match Pattern: The heat load of a hot stream is greater than the heat load of a cold stream in a pattern, i.e. the cold stream is totally serviced. The match is positioned at the hot end of the hot stream. The residual heat load is on the cold portion of the hot stream. (See Figure 3.2)

A match of this class is a second type match; a hot end match and the heat load of the hot stream greater than that of the cold stream. This is an upstream match. For a cooling subproblem, a Class B match is favored, because it leaves a hot process stream at the cold end (Heuristic N2) and also follows the pinch heuristics. (See Table 3.3)

Definition 3.3 Class C Match Pattern: The heat load of a hot stream is greater than the heat load of a cold stream in a pattern, i.e. the cold stream is totally serviced. The match is positioned at the cold end of the hot stream. The residual heat load is on the hot portion of the hot stream. (See Figure 3.3)

A match of this class is a first type match; a cold end match and the heat load of the hot stream greater than that of the cold stream. This is a downstream match. (See Table 3.4)

Definition 3.4 Class D Match Pattern: The heat load of a cold stream is greater than the heat load of a hot stream in a pattern, i.e. the hot stream is totally serviced. The match is positioned at the hot end of the cold stream. The residual heat load is on the cold portion of the cold stream. (See Figure 3.4)

A match of this class is a second type match; a hot end match and the heat load of the cold stream greater than that of the hot stream. This is a downstream match. (See Table 3.4)

When the residual heat load in a match pattern is matched to a utility stream, it is closed or completed pattern. Otherwise, it is an open or incomplete pattern. It can be seen that if the heat load of the residual stream is less than the minimum heating or cooling requirement then the chances that the match pattern will be matched to a utility stream is high. So we give a match pattern which it s residual less than the minimum heating or cooling requirement a high priority in match pattern.

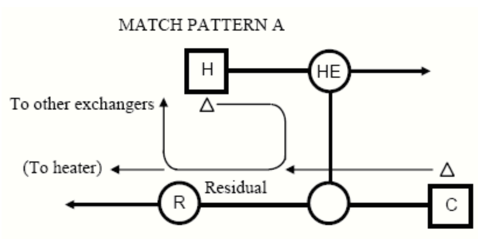


Figure 3.1 Class A Match Pattern

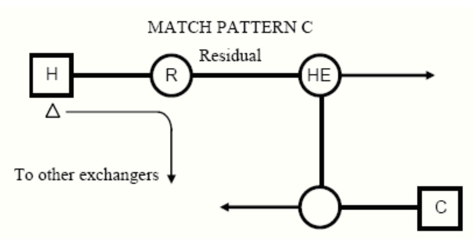


Figure 3.3 Class C Match Pattern

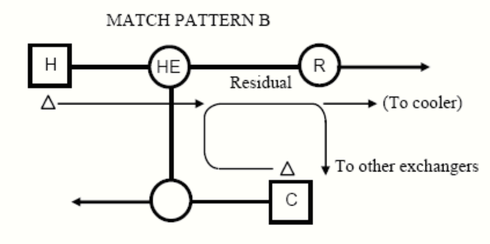


Figure 3.2 Class B Match Pattern

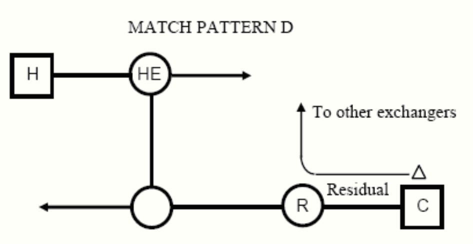


Figure 3.4 Class D Match Pattern

A match of Class A or Class C will leave a residual at the hot end, while a match of Class B or D will leave a residual at the cold end. Heuristics N.3 and N.4 will be use heuristics to further subclassify matches of Class A and B into matches of high priority.

We will make use of Heuristic N.3 and N.4 to further subclassify matches of Class A and B and give the following subclass matches high priorities.

Subclass AH. A match of this subclass is a member of Class A, a heating problem where the residual is less than or equal to the minimum heating requirement. (A letter H in the subclass name denotes that the residual is matched to a heating utility.)

Subclass BK. A match of this subclass is a member of Class B, a cooling problem where the residual is less than or equal to the minimum cooling requirement. (A letter K in the subclass name denotes that the residual is matched to a cooling utility.)

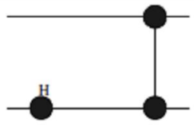
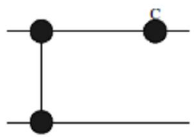
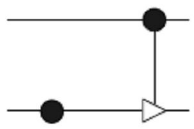
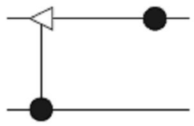
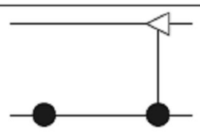
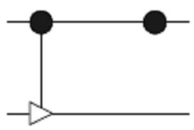
As it might be expected, we give a match of subclasses AH in a heating subproblem and BK in a cooling subproblem the highest priorities. See Table 3.4.

We further discriminate match patterns according to heat capacity flowrate. By following pinch heuristics, in a heating problem, we prefer a match where the heat capacity flowrate of a cold stream is greater than or equal to that of a hot stream. For example, A[H]H is a match in which the heat capacity flowrate of the cold stream is greater than that of the hot stream and the residual of the cold stream is matched to the heating utility.

Similarly in a cooling problem, we prefer a match where the heat capacity flowrate of the hot stream is greater than or equal to that of the cold stream. For example, B[C]K is a match in which the heat capacity flowrate of the hot stream is greater than that of the cold stream and the residual of the hot stream is matched to the cooling utility.

In summary, the rankings of the match patterns in a heating problem are AH, A[H], B[C], A[C], B[H], C[H], D[C], C[C] and D[H]. For a cooling problem, BK, B[C], A[H], B[H], A[C], D[C], C[H], D[H] and C[C].

Table 3.3 Match Pattern Operators of Class A and B

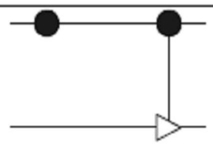
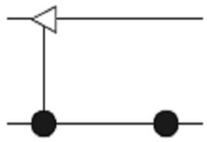
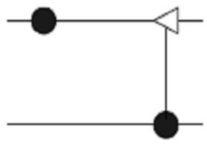
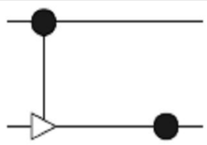
Match Operators	Conditions	Actions
 Pattern AH	$T_H^s\,{}^{*} \ge T_C^t\,{}^{**}$ $L_H \le L_C$ $T_H^s \ge T_C^s + L_H\,W_C^{-1}$ $L_C - L_H \le Q_{\min}^{heating}$	Match H and C Status of H ⇐ Matched*** $T_C^s \Leftarrow T_C^s + L_H\,W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 Pattern BK	$T_H^s \ge T_C^t$ $L_C \le L_H$ $T_C^s \le T_H^s - L_C\,W_H^{-1}$ $L_H - L_C \le Q_{\min}^{cooling}$	Match H and C Status of C ⇐ Matched $T_H^s \Leftarrow T_H^s - L_C\,W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 Pattern A[H]	$T_H^t \ge T_C^s$ $L_H \le L_C$ $W_C \ge W_H$	Match H and C Status of H ⇐ Matched $T_C^s \Leftarrow T_C^s + L_H\,W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 Pattern B[C]	$T_H^s \ge T_C^t$ $L_C \le L_H$ $W_C \le W_H$	Match H and C Status of C ⇐ Matched $T_H^s \Leftarrow T_H^s - L_C\,W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 Pattern A[C]	$T_H^t \ge T_C^s$ $L_H \le L_C$ $W_C < W_H$ $T_H^s \ge T_C^s + L_H\,W_C^{-1}$	Match H and C Status of H ⇐ Matched $T_C^s \Leftarrow T_C^s + L_H\,W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 Pattern B[H]	$T_H^s \ge T_C^t$ $L_C \le L_H$ $W_H < W_C$ $T_C^s \le T_H^s - L_C\,W_H^{-1}$	Match H and C Status of C ⇐ Matched $T_H^s \Leftarrow T_H^s - L_C\,W_H^{-1}$ $L_H \Leftarrow L_H - L_C$

* Tt=target temp, Ts=supply temp, W=heat capacity flowrate, L, Q=heat load.

** Cold stream temperatures are shifted up by Tmin

*** There are two statuses of process streams, active and matched. This will exclude this stream from a set of process streams to be selected next.

Table 3.4 Match Pattern Operators of Class C and D

Match Operators	Conditions	Actions
 Pattern C[H]	$T_H^t \ge T_C^s$ $L_H > L_C$ $W_H \le W_C$	Match H and C Status of C $\Leftarrow$ Matched $T_H^t \Leftarrow T_H^t - L_C W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 Pattern D[C]	$T_H^s \ge T_C^t$ $L_H < L_C$ $W_H \ge W_C$	Match H and C Status of H $\Leftarrow$ Matched $T_C^t \Leftarrow T_C^t + L_H W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 Pattern C[C]	$T_H^t \ge T_C^s$ $L_H > L_C$ $W_C < W_H$ $T_C^t \le T_H^t + L_C W_H^{-1}$	Match H and C Status of C $\Leftarrow$ Matched $T_H^t \Leftarrow T_H^t - L_C W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 Pattern D[H]	$T_H^s \ge T_C^t$ $L_H \le L_C$ $W_H < W_C$ $T_H^t \ge T_C^t - L_H W_C^{-1}$	Match H and C Status of H $\Leftarrow$ Matched $T_C^t \Leftarrow T_C^t + L_H W_C^{-1}$ $L_C \Leftarrow L_C - L_H$

* Tt=target temp, Ts=supply temp, W=heat capacity flowrate, L, Q=heat load.

** Cold stream temperatures are shifted up by Tmin

*** There are two statuses of process streams, active and matched. This will exclude this stream from a set of process streams to be selected next.

3.3 The main functions of control system

In general, the control system installed in process has three main functions.

3.3.1 To reject disturbance.

It is the main objective in installing control system. The external disturbance is uncertain so the operator cannot monitor the changing in process. As a result, the control system must be installed to follow the changing of process and manipulate the process variable to compensate for the disturbance from external factors.

3.3.2. To maintain stability.

The stability is necessary for every process. As a result the control system is set to improve the process stability for the guarantee of quality of product, safety to equipment of process and plant.

3.3.3. To keep the process performing highest efficiency.

Besides rejecting disturbance and maintaining stability, the control system can achieve the great profit because it losses less energy and raw materials during the operating. Moreover the product will meet the required specification and have high production rate.

3.4 Integrated Processes

Three basic features of integrated chemical processes lie at the root of the need to consider the entire plant's control system:

3.4.1 The effects of material recycle.

3.4.2 The effect of energy integration.

3.4.3 The need to account for chemical component inventories.

If we did not have to worry about these issues, then we would not have to deal with a complex plantwide control problem. However, there are fundamental reasons why each of these exists in virtually all real processes.

3.4.1 Material recycles

Material is recycled for six basic and important reasons.

1. Increase conversion: For chemical processes involving reversible reactions, conversion of reactants to products is limited by thermodynamic equilibrium constraints. Separation and recycle of reactants are essential if the process is to be economically viable

2. Improve economics: A reactor followed by a stripping column with recycle is cheaper than one large reactor or three reactors in series.

3. Improve yields: In reaction systems such as A→ B→ C, where B is the desired product, the per-pass conversion of A must be kept low to avoid producing too much of the undesirable product C. Therefore the concentration of B is kept fairly low in the reactor and a large recycle of A is required.

4. Provide thermal sink: In adiabatic reactors and in reactors where cooling is difficult and exothermic heat effects are large, it is often necessary to feed excess material to the reactor so that the reactor temperature increase will not be too large, so the temperatures of the excess materials in the stream flow through the reactor.

5. Prevent side reactions: A large excess of one the reactant is often used so that the concentration of the other reactant is kept low. If this limiting reactant is not kept in low concentration, it could react to produce undesirable products. Therefore the reactant that is in excess must be separated from the product components in the reactor effluent stream and recycled back to the reactor.

6. Control properties: In many polymerization reactors, conversion of monomer is limited to achieve the desired polymer properties. Another reason for limiting conversion to polymer is to control the increase in viscosity that is typical of polymer solutions. This facilitates reactor agitation and heat removal and allows the material to be further processed.

3.4.2 Energy integration

The fundamental reason for the use of energy integration is to improve the thermodynamic efficiency of the process. This translates into a reduction in utility cost. For energy-intensive processes, the savings can be quite significant.

3.4.3 Chemical component inventories

We can characterize a plant's chemical species into three types: reactants, products and inerts. However, the real problem usually arises when we consider reactants (because of recycle) and account for their inventories within the entire process. Because of their value, we want to minimize the loss of reactants exiting the process, so we prevent reactants form leaving. This means we must ensure that every mole of reactant fed to the process is consumed by the reactions.

This is an important concept and is generic to many chemical processes. From the viewpoint of individual units, chemical component balancing is not a problem because exit streams form the units automatically adjust their flows and compositions. However, when we connect units together with recycle streams, the entire system behaves almost like a pure integrator in terms of the reactants. If additional reactant is fed into the system without changing reactor conditions to consume the reactant, this component will build up gradually within the plant because it has no place to leave the system.

3.5 Plantwide control Problem

3.5.1. Units in Series

If process units are arranged in a purely series configuration, where the products of each unit feed downstream units and there is no recycle of material or energy, the plantwide control problem is greatly simplified. It is not had to worry about the issues discussed in the previous section and it can be simply configure the control scheme on each individual unit operation to handle load disturbances.

If production rate is set at the front end of the process, each unit will only see load disturbances coming from is upstream neighbor. If the plant is set up for "on-demand" production, changes in throughput will propagate back through the process. So any individual unit will see load disturbances coming from both its downstream neighbor (flowrate changes to achieve different throughputs) and its upstream neighbor (composition changes as the upstream units adjust to the load changes they see).

Figure 3.5 (a) shows the situation where the fresh feed stream is flow-controlled into the process. The inventory loops (liquid levels) in each unit are controlled by manipulating flows leaving that unit. All disturbances propagate from unit to unit down the series configuration. The only disturbances that each unit sees are changes in its feed conditions.

Figure 3.5 (b) shows the on-demand situation where the flowrate of product C leaving the bottom of the second column is set by the requirements of a downstream unit. Now some of the inventory loops (the base of both columns) are controlled by manipulating the feed into each column

(a)

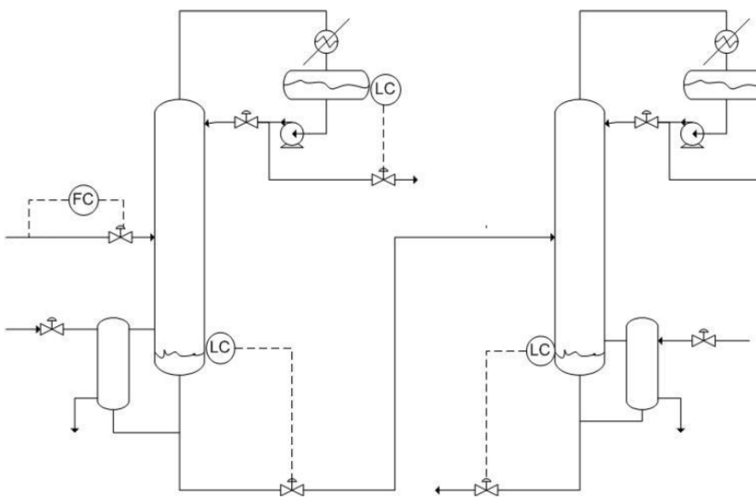


Figure 3.5 Unit in series. (a) Level controls in direction of flow

(b)

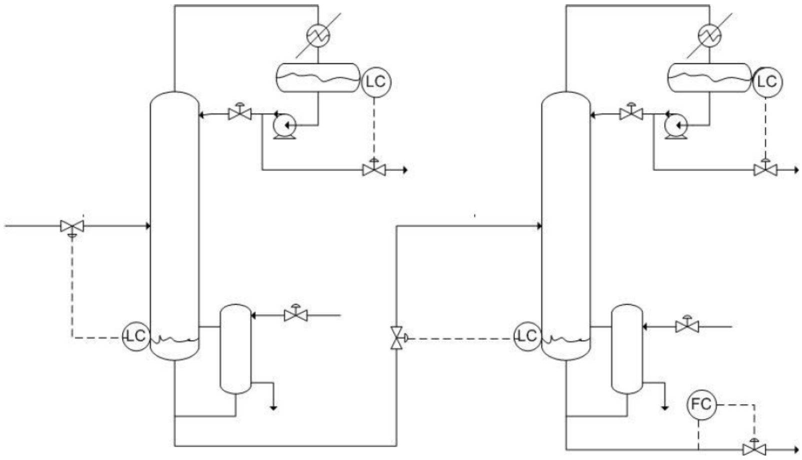


Figure 3.5 Unit in series. (b) Level control in opposite flow.

3.5.2 Effects of Recycle

Most real processes contain recycle streams. In this case the plantwide control problem becomes much more complex and its solution is not intuitively obvious.
The presence of recycle streams profoundly alters the plant's dynamic and steady-state behavior.

Two basic effects of recycle:

1. Recycle has an impact on the dynamics of the process. The overall time constant can be much different than the sum of the time constants of the individual units.
2. Recycle leads to the "snowball" effect.

3.5.3 Snowball effects

High sensitivity of the recycle flowrate to small disturbances is called the snowball effect. It is important to note that this is not a dynamic effect; it is a steady-state phenomenon. But it does have dynamic implications for disturbance propagation and for inventory control. It has nothing to do with closed-loop stability. However, this does not imply that it is independent of the plant's control structure. On the contrary, the extent of the snowball effect is very strongly dependent upon the control structure used.

The large swings in recycle flowrates are undesirable in a plant because they can overload the capacity of the separation section or move the separation section into a flow region below its minimum turndown. Therefore it is important to select a plantwide control structure that avoids this effect.

3.5.4 Reaction/Separation Section Interaction

The plantwide control implication of this idea is that production rate changes should preferentially be achieved by modifying the setpoint of a partial control loop in the reaction section. This means that the separation section will not be significantly disturbed. This means that the column is not disturbed as much as with the alternative control scheme.

Hence a goal of the plantwide control strategy is to handle variability in production rate and in fresh reactant feed compositions while minimizing changes in the feed stream to the separation section. This may not be physically possible or economically feasible. But if it is, the separation section will perform better to accommodate these changes and to maintain product quality, which is one of the vital objectives for plant operation. Reactor temperature, pressure, catalyst/initiator activity, and holdup are preferred dominant variables to control compared to direct or indirect manipulation of the recycle flows, which of course affect the separation section.

3.6 Basic Concepts of Plantwide Control

3.6.1. Buckley basics

Page Buckley (1964), a true pioneer with DuPont in the field of process control, was the first to suggest the idea of separating the plantwide control problem into two parts: material balance control and product quality control.

He then proposed establishing the product-quality control loops by choosing appropriate manipulated variables. The time constants of the closed-loop product-quality loops are estimated.

Then the inventory loops are revisited. The liquid holdups in surge volumes are calculated so that the time constants of the liquid level loops are a factor of 10 large than the product-quality time constants. Note that most level controllers should be proportional-only (P) to achieve flow smoothing.

3.6.2 Douglas doctrines

Jim Douglas (1988) of the University of Massachusetts has devised a hierarchical approach to the conceptual design of process flowsheets. Although he primarily considers the steady-state aspects of process design, he has developed several useful concepts that have control structure implications.

Douglas points out that in the typical chemical plant the costs of raw materials and the value of the products are usually much greater than the costs of capital and energy. This leads to the two Douglas doctrines:

1. Minimize losses of reactants and products.
2. Maximize flowrates through gas recycle system.

The first idea implies that we need tight control of stream compositions exiting the process to avoid losses of reactants and product. The second rests on the principle that yield is worth more than energy. Recycles are used to improve yields in many processes.

3.6.3 Downs drill

Jim Downs (1992) of Eastman Chemical Company has insightfully pointed out the importance of looking at the chemical component balances around the entire plant and checking to see that the control structure handles these component balances effectively.

We must ensure that all components (reactants product, and inerts) have a way to leave or be consumed within the process. The consideration of inerts is seldom overlooked. Heavy inerts can leave the system in the bottoms product from a distillation column. Light inerts can be purged form a gas recycle stream or form a partial condenser on a column. Intermediate inerts must also be removed in some way, for example in sidestream purges or separate distillation columns.

Most of the problems occur in the consideration of reactants, particularly when several chemical species are involved. All of the reactants fed into the system must either be consumed via reaction or leave the plant as impurities in the exiting streams. Since we usually want to minimize raw material costs and maintain high-purity products, most of the reactants fed into the process must be chewed up in the reactions. And the stoichiometry must be satisfied down to the last molecule.

3.6.4 Luyben laws

Three laws have been developed as a result of a number of case studies of many types of systems:

1. A stream somewhere in all recycle loops should be flow controlled. This is to prevent the snowball effect.

2. A fresh reactant feed stream cannot be flow-controlled unless there is essentially complete one-pass conversion of one of the reactants. This law applies to systems with reaction type such as $A+B \rightarrow$products. In systems with consecutive reactions such as $A+B \rightarrow M+C$ and $M+B \rightarrow D+C$., the fresh feeds can be flow-controlled into the system because any imbalance in the ratios of reactants is accommodated by a shift in the amounts of the two products (M and D) that are generated.

3. If the final product from a process comes out the top of a distillation column, the column feed should be liquid. If the final product comes out the bottom of a column, the feed to the column should be vapor.

3.6.5 Richardson rule

Bob Richardson of Union Carbide suggested the heuristic that the largest stream be selected to control the liquid level in a vessel. This makes good sense because it provides more muscle to achieve the desired control objective. The point is that the bigger the handle you have to affect a process, the better you can control it. This is why there are often fundamental conflicts between steady-state design and dynamic controllability.

3.6.6 Shinskey schemes

Greg Shinskey (1988), over the course of a long and productive career at Foxboro, has proposed a number of "advanced control" structures that permit improvements in dynamic performance. These schemes are not only effective, but they are simple to implement in basic control instrumentation. Liberal use should be made of ratio control, cascade control, override control, and valve-position (optimizing) control. These strategies are covered in most basic process control textbooks.

3.6.7 Tyreus tuning

One of the vital steps in developing a plantwide control system, once both the process and the control structure have been specified, is to determine the algorithm to be used for each controller (P, PI, or PID) and to tune each controller. We strongly recommend the use of P-only controllers for liquid levels. (Controller gain equal to 1.67). This will have the valve wide open when the level is at 80 percent and the valve shut when the level is at 20 percent.

For other control loops, we suggest the use of PI controllers. The relay-feedback test is a simple and fast way to obtain the ultimate gain (K_u) and ultimate period (P_u). Then either the Ziegler-Nichols settings (for very tight control with a closed-loop damping coefficient of about 0.1) or the Tyreus-Luyben (1992) settings (for more conservative loops where a closed-loop damping coefficient of 0.4 is more appropriate) can be used:

$$K_{ZN}=K_u/2.2 \qquad T_{ZN}=P_u/1.2$$

$$K_{TL}=K_u/3.2 \qquad T_{TL}=2.2P_u$$

The use of PID controllers should be restricted to those loops where two criteria are both satisfied: the controlled variable should have a very large signal-to-noise ratio and tight dynamic control is really essential from a feedback control stability perspective. The classical example of the latter is temperature control in an irreversible exothermic chemical.

3.7 Step of Plantwide Process control Design Procedure

The nine steps of the design procedure center around the fundamental principles of plantwide control: energy management; production rate; product quality; operational, environmental, and safety constraints; liquid level and gas pressure inventories; makeup of reactants; component balances; and economic or process optimization.

Step 1: Establish control objectives.

Assess the steady-state design and dynamic control objectives for the process.

This is probably the most important aspect of the problem because different control objectives lead to different control structures.

These objectives include reactor and separator and separation yields, product quality specifications, product grades and demand determination, environmental restrictions, and the range of safe operation conditions.

Step 2: Determine control degrees of freedom

Count the number of control valves available.

This is the number of degrees of freedom for control that is the number of variables that can be controlled to setpoint. The valves must be legitimate (flow through a liquid-filled line can be regulated by only one control valve **Step3:**

Establish energy management system

Make sure that energy disturbances do not propagate throughout the process by transferring the variability to the plant utility system.

The term energy management is used to describe two functions:

1. To provide a control system that removes exothermic heats of reaction form the process. If heat is not removed to utilities directly at the reactor, then it can be used elsewhere in the process by other unit operations. This heat however must ultimately be dissipated to utilities.

2. To provide a control system that prevents the propagation of thermal disturbances and ensures the exothermic reactor heat is dissipated and not recycled.

Step 4: set production rate

Establish the variables that dominate the productivity of the reactor and determine the most appropriate manipulator to control production rate.

Often design constraints require that production be set at a certain point. An upstream process may establish the feed flow sent to the plant. A downstream process may require on demand production, with fixes the product flow rate from the plant. If no constraint applies, then we select the valve that provides smooth and stable production-rate transitions and rejects disturbances. We often want to select the variable that has the least effect on the separation section, but also has a rapid and direct effect on reaction rate in the reactor without heating an operational constraint.

Step 5: Control product quality and handle safety, operational, and environmental constraints.

Select the "best" valves to control each of the product-quality, safety, and environmental variables.

We want tight control of these quantities for economic and operational reasons. Hence we should select manipulated variables such that the dynamic relationships between controlled and manipulated variables feature small time constants and dead times and large steady-state gains.

Step 6: Control Inventories (Pressures and Levels) and Fix a Flow in Every Recycle Loop.

Determine the valve to control each inventory variable. These variables include all liquid levels and gas pressures. An inventory variable should typically be controlled with the manipulated variable that has the largest effect on it within that unit.

Proportional-only control should be used in nonreactive level loops for cascaded units in series. Even in reactor-level control, proportional control should be considered to help filter flow-rate disturbances to the downstream separation system. There is nothing necessarily sacred about holding reactor level constant.

In most processes a flow controller should be present in all liquid recycle loops. This is a simple and effective way to prevent potentially large changes in recycle flows that can occur if all flows in recycle loops are controlled by levels. Two benefits result from this flow-control strategy. First, the plant's separation section is not subjected to large load disturbances. Second, consideration must be given to alternative fresh reactant makeup control strategies rather than flow control. In dynamic sense, level controlling all flows in recycle loop is a case of recycling of disturbances and should be avoided.

Step 7: Check Component Balances.

Identify how chemical components enter, leave, and are generated or consumed in the process.

Ensure that the overall component balances for each chemical species can be satisfied either through reaction or exit streams by accounting for the component's composition or inventory at some point in the process. Reactant must be consumed in the reaction section or leave as impurities in the product streams. Fresh reactant makeup feed streams can be manipulated to control reactor feed composition or a recycle stream composition (or to hold pressure or level as noted in the previous step). Purge streams can also be used to control the amount of high- or low-boiling impurities in a recycle stream.

Step 8: Control Individual Unit Operations

Establish the control loops necessary to operate each of the individual unit operations.

For example, a tubular reactor usually requires control of inlet temperature. High-temperature endothermic reactions typically have a control system to adjust fuel flowrate to a furnace supplying energy to the reactor. Crystallizers require manipulation in the stack gas from a furnace is controlled to prevent excess fuel usage. Liquid solvent feed flow to an absorber is controlled as some ratio.

Step 9: Optimize Economics or Improve Dynamic Controllability

Establish the best way to use remaining control degrees of freedom.

After satisfying all of the basic regulatory requirements, additional degrees of freedom involve control valves that have not been used and setpoints in some controllers that can be adjusted. These can be used either to optimize steady-state economic process performance or to improve dynamic response.

3.8 Plantwide process control

Buckley (1964) proposed a control design procedure for the plantwide control problem that consisted of two stages. The first stage determined the material balance control structure to handle vessel inventories for low-frequency disturbances. The second established the product quality control structure to regulate high-frequency disturbances. This procedure has been widely and effectively utilized. It has served as the conceptual framework in many subsequent ideas for developing control systems for complete plants. However, the two-stage Buckley procedure provides little guidance concerning three important aspects of plantwide control strategy. First, it does not explicitly discuss energy management. Second, it does not address the specific issues of recycle systems. Third, it does not deal with component balance in the context of inventory control. By placing the priority on material balance over product quality control, the procedure can significantly limit the flexibility in choosing the latter.

The goals for an effective plantwide process control system include.

1. Safe and smooth process operation.
2. Tight control of product quality in the face of disturbances.
3. Avoidance of unsafe process condition.
4. A control system runs in automatic, not manual, requiring minimal operator attention.
5. Rapid rate and product quality transitions.
6. Zero unexpected environmental releases.

3.9 New Plantwide Control Structure Design Procedure

The book by Luyben et al. (1998) outlines the control structure design procedure based on heuristic and their process engineering insight. Several case studies are given in the book. Luyben's procedure is widely studied and used the plantwide process control community. However, the structural design procedure is not given explicitly. Their case study designs followed the outline and collected heuristic law but need the designer's process engineering insight to pair CVs and MVs. Skogestad (2004) presented the new design procedure mainly based on the mathematical analysis. First, the dynamic and steady state degree of freedom are identified. Then the set of primary controlled variables are determined. They basically are the active constraints and the variables that must be maintained to achieve minimal economic loss when disturbances occur. Then the control variable determining the production rate is selected based on the optimization resulted in the previous step. The pairings of the selected sets of MVs and CVs are done hierarchically: regulatory control, supervisory control (loop enhancement for SISO or constraint handling for MPC), and optimization layers. Several analysis tools are utilized in these steps, e.g. pole vector analysis, RGA, CLDG, linear and nonlinear optimization. However, he did not discuss which controlled variables should take precedence? Normally the plant would have a large number of variables; the precedence of the control variables must be established to assure the optimality of the designs and to avoid ambiguity in the design decision. Second, how to handle the disturbances is not discussed.

New design procedure of Wongsri (2009) presented plantwide control structure design procedure based on heuristics and mathematical analysis. In this procedure, the precedence of control variables is established The major disturbances are directed or managed explicitly to achieve the minimal interaction between loops by using the extended (thermal) disturbance propagation method (Wongsri, 1990) to cover the material disturbances. The purposed plantwide control structure design procedure for selection the best set of control structure is intuitive, simple and straightforward.

Normally, plantwide control design procedures consider decision about plant control structures in perspective. The plantwide control structure design is complex: hierarchical, structural, having mixed objectives, containing many units and layers, and therefore confusing. One easy way to deal with this complexity is compartmentalizing it. However, the plant is not merely the units combined; it has its own properties. The whole is greater than the sun of its parts. There properties (or behavior) of a system as a whole emerge out of the interaction and the relationship of the components comprising the system. Therefore, a designer must deal with both parts and system.

New design procedures of Wongsri (2009) are:

Step1. Establishment of control objective.

Step2. Selection of controlled variables to maintain product quality and to satisfy safety operational and environmental constrains and to setting the production rate. The selected CVs are ranked using the Fixture Point theorem.

Step3. Selection of manipulated variables and measurements via DOF analysis.

Step4. Energy management via heat exchanger networks.

Step5. Selection of control configuration using various tools available.

Step6. Completing control structure design by checking the component balance.

Step7. Selection of controller type: single loop or MPC.

Step8. Validation via rigorous dynamic simulation.

Fixture point theorem analysis:

1. The process is considered at dynamic mode (we run the process until the process responses are at steady state).
2. Controlled variable (CV) can be arranged to follow the most sensibility of the process variable by step change of the MV in open loop control (change only one MV, the other should be fixed than alternate to other until complete).
3. Study the magnitude of integral absolute error (IAE) of all process variables that deviates from steady state.
4. Select CV by considering CV that gave the most deviation from steady state (high value score.

3.10 Control of process-to-process exchanger

Process-to-process (P/P) exchangers are used for heat recover within a process. We can control the two exit temperatures provided we can independently manipulate the two inlet flowrates. However, these flowrates are normally unavailable for us to manipulate and we therefore give up two degrees of freedom fairly easily.

It is possible to oversize the P/P exchanger and provides a controlled bypass around it as in Figure 3.8.a. It is possible to combine the P/P exchanger with a utility exchanger as in Figure 3.8.b.

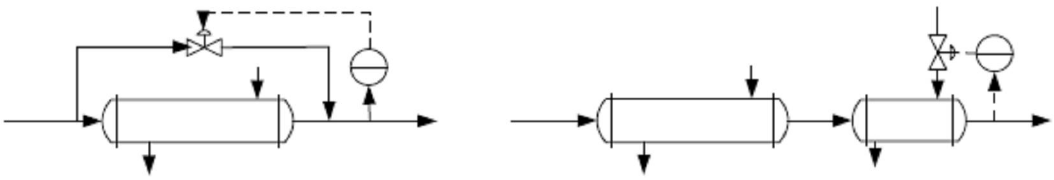


Figure 3.8 Control of P/P heat exchangers: (a) use of bypass; (b) use of auxiliary utility exchanger.

3.10.1 Use of Bypass Control

When the bypass method is used for unit operation control, we have several choices about the bypass location and the control point. Figure 3.9(b) shows the most common alternatives. For choosing the best option, it depends on how we define the best. Design consideration might suggest, we measure and bypass on the cold side since it is typically less expensive to install a measurement device and a control valve for cold service than it is for high-temperature service. Cost consideration would also suggest a small bypass flow to minimize the exchanger and control valve sizes.

From a control standpoint, we should measure the most important stream, regardless of temperature, and bypass on the same side as well we control (see Figure 3.9 a and c). This minimizes the effects of exchanger dynamics in the loop. We should also want to bypass a large fraction of the controlled stream since it improves the control range. This requires a large heat exchanger. There are several general heuristic guidelines for heat exchanger bypass streams. We typically want to bypass the flow of the stream whose temperature we want to control. The bypass should be about 5 to 10 percent of the flow to be able to handle disturbances. Finally, we must carefully consider the fluid mechanics of the bypass design for the pressure drops through the control valves and heat exchanger.

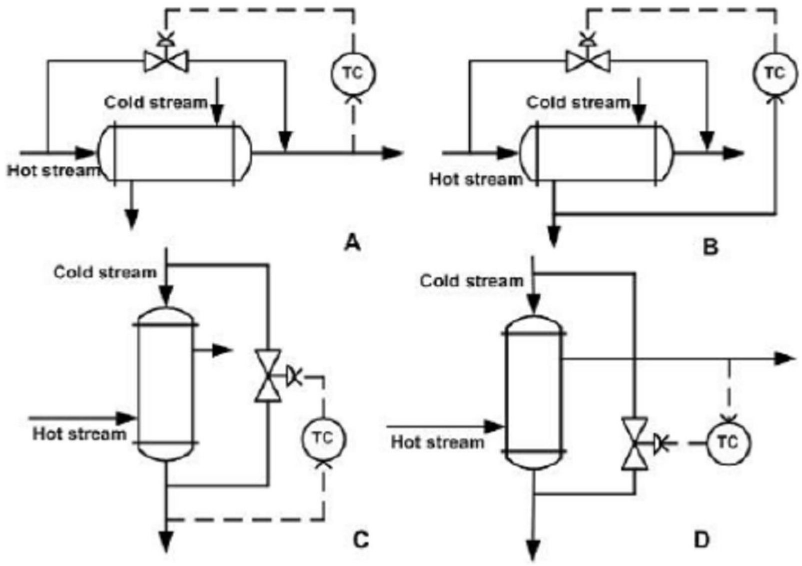


Figure 3.9 Bypass control of process-to process heat exchangers. (a) Controlling and bypassing hot stream; (b) controlling cold stream and bypassing hot stream; (c) controlling and bypassing cold stream; (d) controlling hot stream and bypassing hot stream.

CHAPTER IV

CUMENE PROCESS

4.1 Introduction

The structural control design structures of cumene process are complex, which features several unit operations. There are two recycles stream and many control loops. In addition, cumene process consists of reaction section, separation section and energy recovery in order to take full advantage in the production. There is one reactant feeds split among the plug flow reactor. The other reactant is kept in excess by a large recycle stream. The reactor operates at high pressure. And refrigeration is required to remove the exothermic heat of reaction. Reactor cooling is achieved by auto refrigeration in the HYSYS simulation. The separation section consists of two distillation columns, to separate products from reactants.

4.2 Process Description

The cumene process features the desired reaction of benzene with propylene to form cumene and the undesirable reaction of cumene with propylene to form p-diisopropylbenzene.

The main reaction for production of cumene involves the reaction of benzene with propylene in a high-temperature, high-pressure gas-phase reactor.

$$C_6H_6 + C_3H_6 \longrightarrow C_9H_{12} \qquad (4.1)$$

There is also a sequential reaction of cumeme and propylene to form p-diisopropylbenzene (PDIB).

$$C_9H_{12} + C_3H_6 \longrightarrow C_{12}H_{18} \qquad (4.2)$$

The kinetic data is taken from a case study given by Turton et al. (2003). These exothermic reactions are irreversible and occur in the vapor phase.

The kinetic expressions assumed to be valid for the system are:

$$R_1 = 2.8x10^7 \exp\left(\frac{-104174}{RT}\right) C_P C_B \qquad (4.3)$$

$$R_2 = 2.32x10^9 \exp\left(\frac{-146742}{RT}\right) C_P C_B \qquad (4.4)$$

where the unit of reaction rates, concentration and activation energies are kmol s^{-1} m^3, molarity and kJ/kmol, respectively.

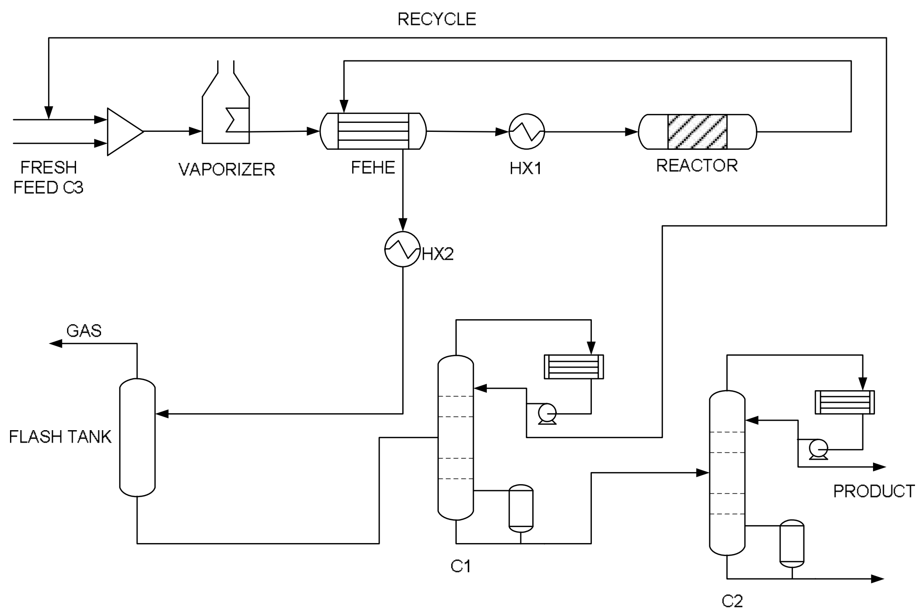


Figure 4.1 Conceptual flowsheet for manufacturing cumene

Figure 4.1 show the flowsheet of the cumene process. The fresh feed streams of benzene and mixed C3 (propylene and propane) enter the process as liquids. The fresh feed flow rate of the C3 feed is set at101.93 kmol/h. The composition of this feed is 95 mol% propylene and 5 mol% propane. Propane must leave the process somewhere. It is vented in a gas stream from the flash drum.

The fresh feed of benzene is98.78 kmol/h. The liquid fresh feeds are combined with a liquid recycle stream (benzene recycle, the distillate from the first column C1) and fed into a vaporizer. The total benzene fed to the reactor (sum of fresh plus recycle) is180.2 kmol/h. The saturated gas leaves the vaporizer at 210 °C and 25 bar. It is preheated in two heat exchangers. The first recovers heat from the hot reactor effluent at 358.5 °C. The second adds additional heat to bring the reactor inlet temperature up to 358 °C.

Reaction Section, the reactor is a cooled tubular reactor that generates high-pressure steam from the exothermic reactions. There are 1500 tubes, 0.0763 m in diameter, and 6 m in length. There are solid catalyst with a void fraction of 0.5 and a solid density of 2000 kg/m3. The temperature on the steam side of the reactor is 358 °C. An overall heat-transfer coefficient of 0.065 kW m^{-2} K^{-1} is used. Reactor effluent leaves at 358.5 °C, is cooled to 281.7 °C in the feed-effluent heat exchanger (FEHE), and sent to a condenser in which it is cooled to 90 °C using cooling water. The two-phase stream from the heat exchanger is fed to a flash tank. The gas from the tank is used as fuel. The liquid is fed into the first distillation column C1.

Benzene Recycle Column C1, The column has 15 stages and is fed on stage 6, which is the optimum feed stage to minimize reboiler heat input. The operating pressure is 1.75 bars, which gives a reflux-drum temperature of 60 °C, so cooling water can be used in the condenser. The reflux ratio is small (RR= 0.44). The distillate is mostly benzene and is recycled back to the reactor. The composition is 92.86 mol% of benzene with small amounts of propylene and propane that are in the liquid from the flash drum.

The design specification is to keep benzene from dropping out of the bottom and affecting the purity of the cumene product leaving in the distillate of the downstream column. Since the specified cumene purity is 99.9 mol %, a very small benzene composition in the bottoms (0.001 mol %) is required.

Cumene Product Column C2, The column has 20 stages and is fed on stage 12. The operating pressure is 1 bar psia, which gives a reflux-drum temperature of 152 °C. The reflux ratio is low (RR=0.63).

The design specification is to attain high-purity cumene in the distillate and minimize the loss of cumene in the bottoms. The bottoms composition is set at 0.1 mol % cumene. The distillate composition is 99.9 mol % cumene using the 0.63 reflux ratio. Figure 4.2 show the flowsheet of the cumene process by using HYSYS simulations with equipment size and condition used by Luyben (2010).

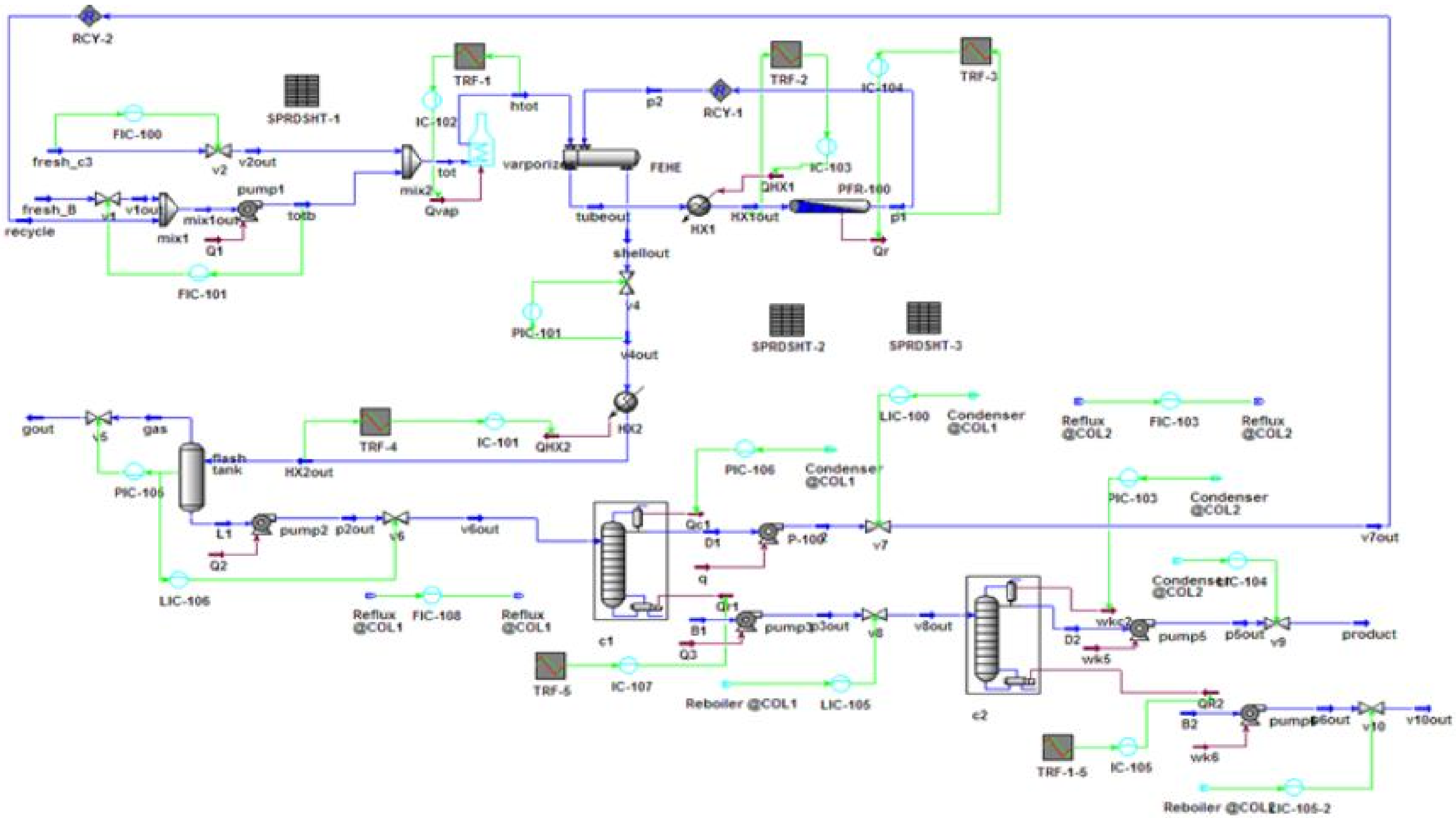


Figure 4.2 The flowsheet of the cumene process by using HYSYS simulation with equipment size and condition used by Luyben(2010).

4.3 Design of Heat Exchanger Networks

At this point, the heat exchanger network design method provided by Wongsri (1990) is used to design the resilient heat exchanger network for cumene process. The design procedures and definitions from previous chapters will be an accessory to design in conceptual design. The information for used in design is shown in Table 4.1.

Table 4.1 The information of cumene process.

Stream Name	Tin (°C)	Tout (°C)	W (MJ/hr °C)	duty(MJ/hr)
H1: Reactor Product Stream	358.50	80.00	64.90	18074.39
H2: Benzene Column Condenser	95.95	49.68	122.11	5650.18
H2: Cumene Column Condenser	151.80	151.60	28142.17	5628.43
C1: Reactor feed Stream	35.50	358.00	61.67	19889.92
C2: Benzene Column Reboiler	176.10	177.10	7583.42	7583.42
C3: Cumene Column Reboiler	212.50	213.00	10054.32	5027.16

4.3.1 HEN Base case

According to Table 4.1, it can be simply translated to a heat exchanger network for cumene process (Base case) in Figure 4.3.

There are two streams in the network. We do not fine Pinch temperature using Problem table method.

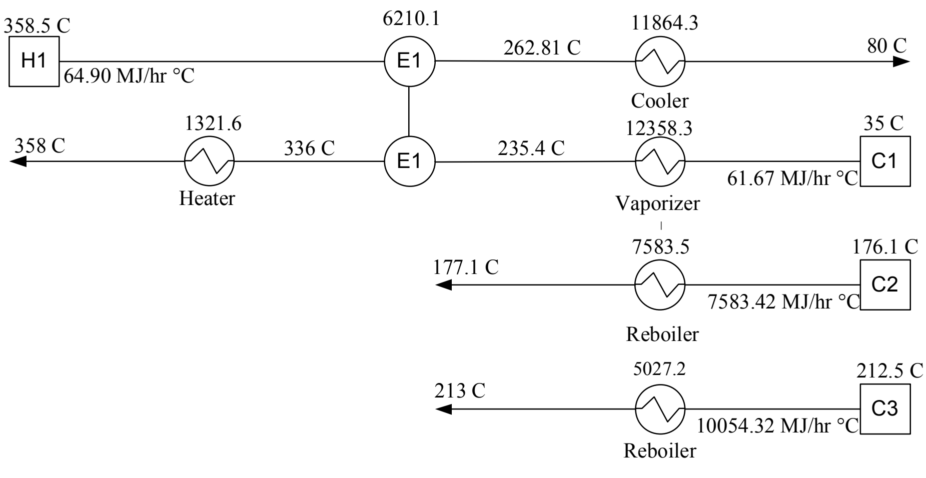


Figure 4.3 The heat exchanger network, base case for cumene process.

4.3.2 HEN design for cumene process

There are three hot streams and three cold streams in the network. We can fine Pinch temperature using Problem table method as show in table 4.2. At the minimum heat load condition, the pinch temperature occurs at 186.1/176.1 °C. The minimum utility requirement has been predicted 1.2631×10^4MJ/hr of hot utilities and 9779.64 MJ/hr of cold utilities**.** Figure 4.4 shows a design of heat exchanger network for cumene process.

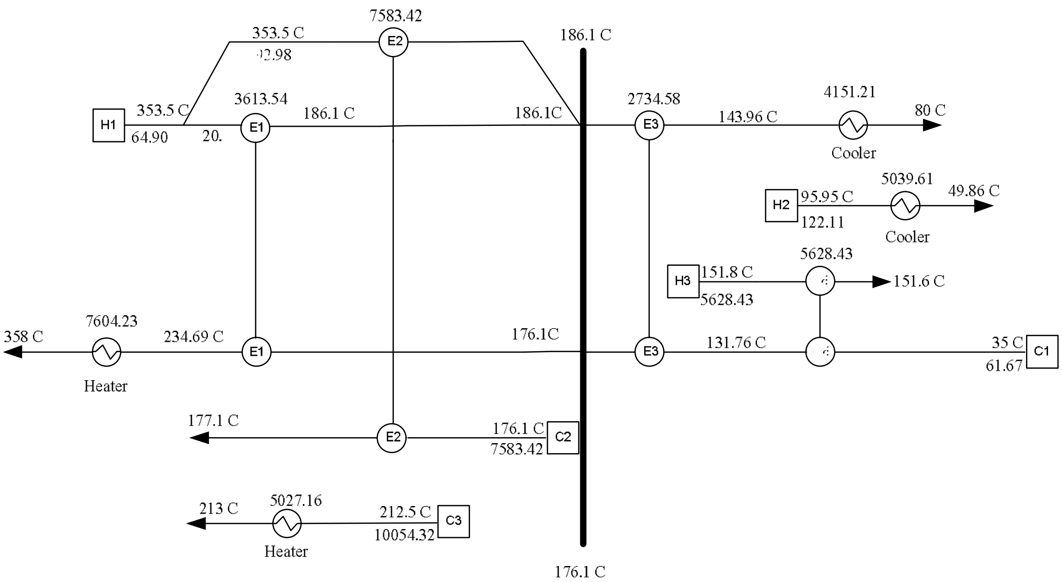


Figure 4.4 The heat exchanger network for cumene process

Table 4.2 Problem table for HEN design of cumene process

W						T hot	T cold	del T	Sum W	Require	Interval (H)	Cascade	Sum Interval
H1	H2	H3	C1	C2	C3								
0	0	0	0	0	0	368	358			Qh			
0	0	0	62	0	0	359	349	10	-62	12641	-586	12055	-586
65	0	0	62	0	0	223	213	136	3	12055	437	12492	-149
65	0	0	62	0	10054	223	213	1	-10051	12492	-5026	7466	-5174
65	0	0	62	0	0	187	177	35	3	7466	114	7580	-5060
65	0	0	62	7583	0	186	176	1	-7580	7580	-7580	0	-12641
65	0	0	62	0	0	152	142	34	3	0	111	111	-12530
65	0	28142	62	0	0	152	142	0	28145	111	5629	5740	-6901
65	0	0	62	0	0	96	86	56	3	5740	179	5919	-6721
65	122	0	62	0	0	80	70	16	125	5919	1999	7918	-4722
0	122	0	62	0	0	50	40	30	60	7918	1833	9751	-2890
0	0	0	62	0	0	46	36	4	-62	9751	-258	9493	-3147
												Qc	

The various alternatives of heat exchanger network are design for the cumene process, the energy saved from Base case as show in Table 4.8

Table 4.4 Energy integration of Cumene process

Utilities usage (MJ/hr)	Base Case	HEN
Heater	1347.94	7604.23
Vaporizer	12358.26	-
C1 reboiler	7583.42	-
C2 reboiler	5027.16	5027.16
Hot utilities usage	26316.78	12631.40
Cold utilities usage	11864.30	4151.21
Total utilities usage	38181.08	16782.60
Energy saving %	-	56.06

4.4 New alternatives for Cumene process

Three alternative of heat exchanger networks (HEN) designs of cumene process are proposed to save energy from the base case and use to evaluate performance of control structures are design both simply energy-integrated process and complex energy-integrated process.

In Figure 4.5 show the base case of cumene process with simply energy-integration, we used a feed-effluent of heat exchanger (FEHE) to reduce the energy of heater (HX1). The heat added the heater (HX1) and heat of reaction are therefore removed in the condenser (HX2).

In Figure 4.6 show heat exchanger network design for cumene process There are three heat exchangers for preheat the reactor feed stream and there are two heat exchangers for preheat the reboiler in the benzene recycle column (C1) driven by the reactor effluent stream. The heat added the heater (HX1) and heat of reaction are therefore removed in the condenser (HX2).

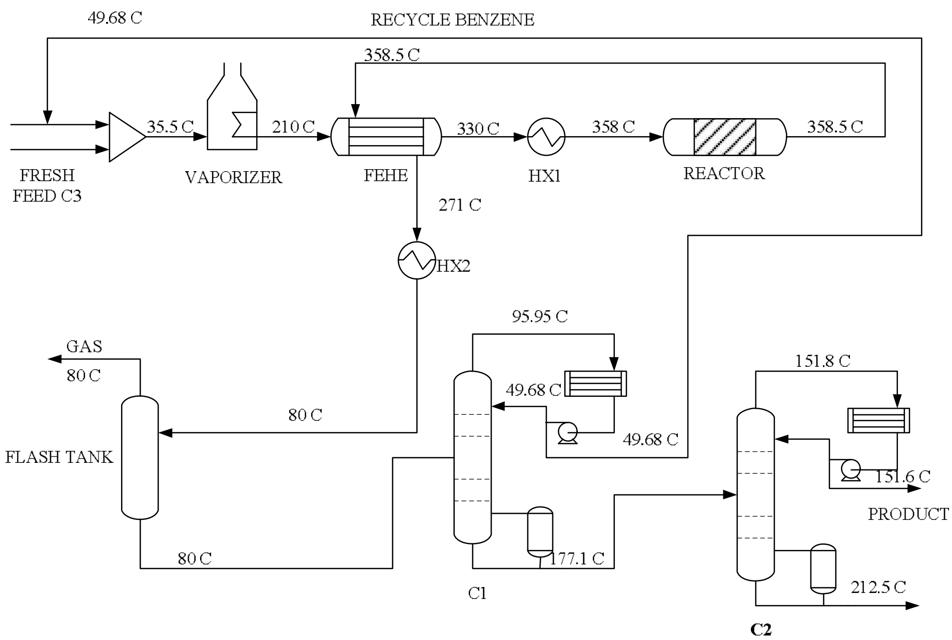


Figure 4.5 Cumene process, Base case

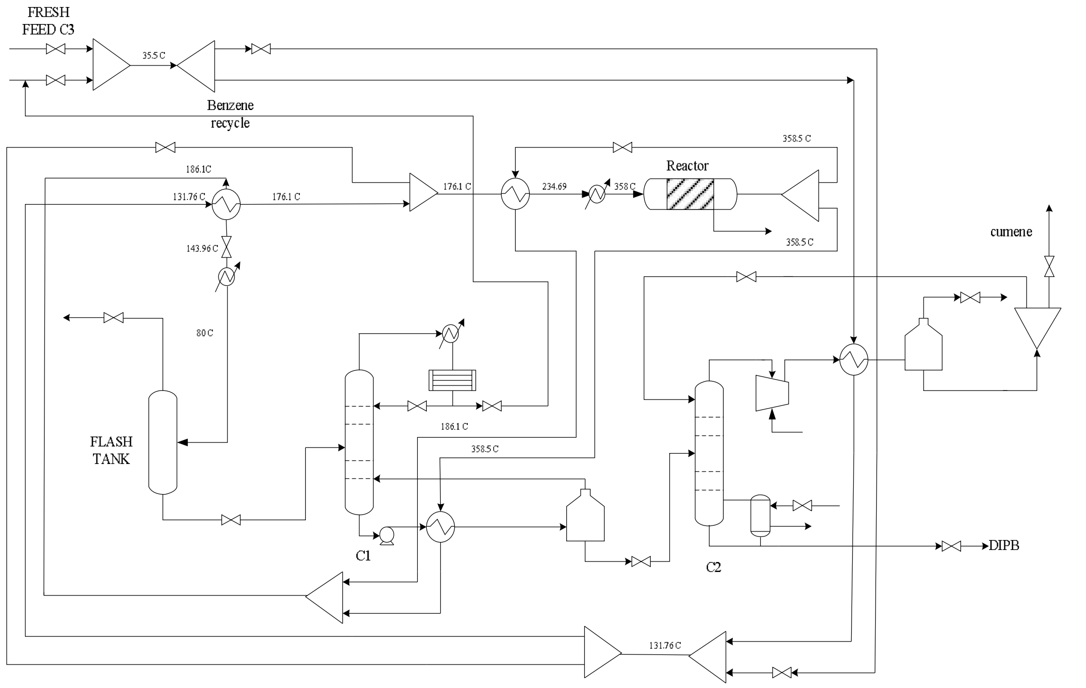


Figure 4.6 Cumene process heat exchanger network design

4.5 Steady-State Modeling of Cumene Process

First, a steady state model is built in HYSYS PLANT, using the flowsheet and equipment design information, mainly taken from Luyben (2010). Appendix A presents data and specification for the different equipment. For the simulation, the NRTL physical property package is used in the Aspen HYSYS simulations. The reaction kinetics of both reactions are modeled with standard Arrhenius kinetic expressions available in HYSIS PLANT, and the kinetic data are taken from Luyben (2010).

4.5.1 Steady State Simulation of Cumene Process (Base case)

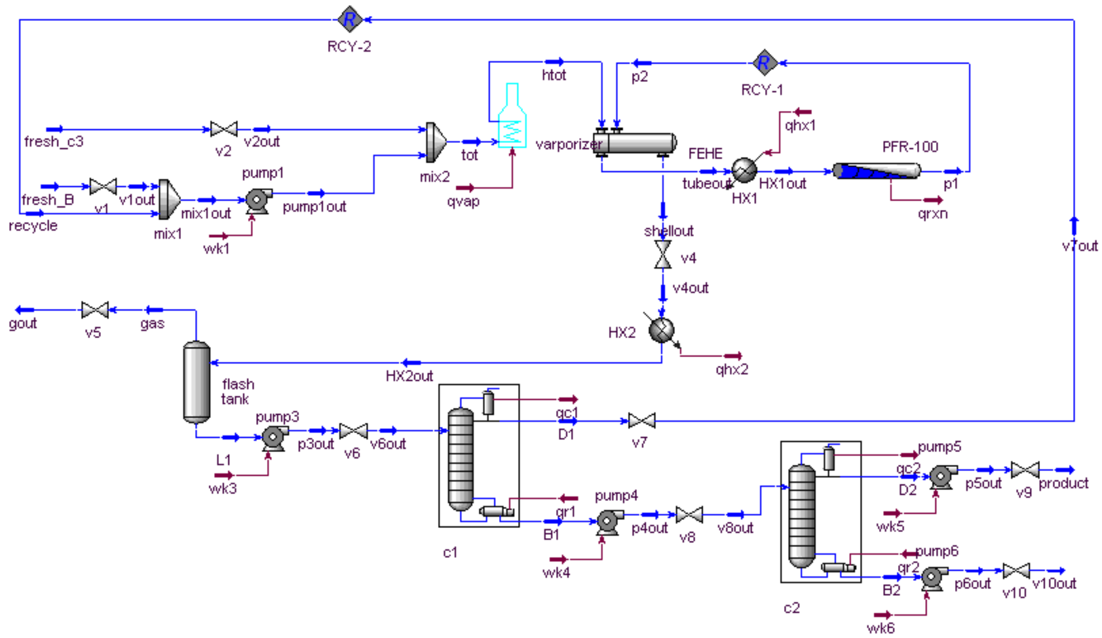


Figure 4.7 The HYSYS flowsheet of Cumene process (base case)

4.5.2 Steady State Simulation of Cumene Process for heat exchanger network

There is a heat exchanger and additionally the reboiler in the benzene recycle column (C1 column) which is driven by the reactor effluent stream. The heat exchanger is used to reboil the C1 column. Moreover, there is a heat exchanger and additionally the condenser in the cumene column (C2 column) for preheat the reactor feed stream.

The C1 column is simulated using a “refluxed absorber” which does not include a reboiler. The cumene column (C2 column) is simulated using the “distillation column” module. Since a “refluxed absorber” module is used, only one variable need to be specified for the columns with condenser. The overhead mole fraction is chosen to be specified for a “refluxed absorber” module. There is a tank to accommodate liquid the bottom of C1 column.

The C2 column is simulated using a “reboiled absorber” that is does not include a condenser. The benzene recycle column (C1 column) is simulated using the “distillation column” module. Since a “reboiled absorber” module is used, only one variable need to be specified for the columns with reboiler. The bottom mole fraction is chosen to be specified for a “reboiled absorber” module. A tank is needed to accommodate liquid the distillate of C2 column.

Figure 4.8 show the HYSYS flowsheet of Cumene process with energy integration schemes for heat exchanger network design by using disturbance propagation method.

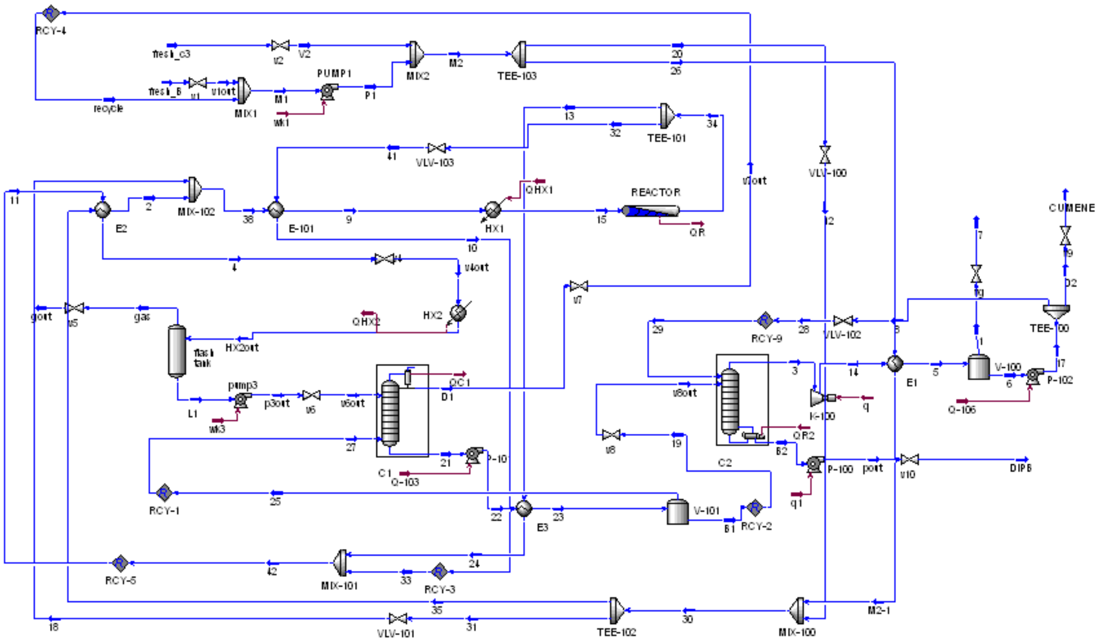


Figure 4.8 The HYSYS flowsheet of Cumene process for heat exchanger network design

CHAPTER V

CONTROL STRUCTURE DESIGN AND DYNAMIC SIMULATION

Maintaining the plant energy and mass balances are the essential task of plantwide for a complex plant consists of recycle streams and energy integration when the disturbance load come through the process. The control system is needed to reject loads and regulate an entire process into a design condition to achieve its objectives therefore our purpose of this chapter is to present the new control structures of cumene process. Moreover, the two designed control structures are also compared between base case of cumene process based on rigorous dynamic simulation by using the commercial software HYSYS version 7.0.

5.1 New Plantwide Control Strategies

The proposed plantwide control structure design procedure for selection the best set of control structure is intuitive, simple and straightforward. In this research the plantwide control structures of cumene process are designed based on the new design procedure given by Wongsri (2009) for all designed control structures and discussed below.

Step 1: Establishment of control objectives

In this research, the objectives were decomposed into two levels: Plantwide level and Unit level.

Plantwide Level: For cumene process, the control objectives maintain product purity of cumene composition at 99.9 mol%.

Unit Level: Stabilization and smooth operation.

Process constraints during operation: The reactor temperature should be around 358 °C and pressure about 25 bars. This is an optimization decision to have better reaction rate. Providing a large recycle stream to maintain the desired yield at 105 kmols/ hr with 94 mol% benzene purity to recycle back to reactor.

Step 2: Selection of controlled variables to maintain product quality and to satisfy safety operational and environmental constrains and to setting the production rate. The selected CVs are ranked using the Fixture Point theorem

Plantwide Level: Consider material recycle loop the causes of "snowball effect" to a system. Cumene process has a small recycle stream. Therefore snowball effect is relatively little impact.

Unit Level: Use the Fixture Point theorem to select appropriate controlled variables from a candidate output to maintain product quality, to satisfy safety operational, environmental constrains and to setting the production rate. The most disturbed points must be satisfactorily controlled by giving them consideration before other variables. Screening output variables for identification controlled variables by using input variables change (change five percent of manipulated variables). Table 5.1 shows the ranked CVs and the IAE summation result.

Table 5.1 Ranked CVs and IAE summation result

Rank	Variables		SUM IAE
1	flash tank	level	0.20069
2	C1_Condenser	Vessel pressure	0.08036
3	flash tank	Vessel pressure	0.07976
4	feed stream to HX2	pressure	0.07139
5	feed stream to Benzene column(C1)	molar flow rate	0.06390
6	C1_Reboiler	level	0.04038
7	Benzene Fresh feed stream	molar flow rate	0.03214
8	feed stream to cumene column(C2)	molar flow rate	0.02986
9	flash tank	inlet temperature	0.02756
10	C2_Condenser	Vessel pressure	0.02445
11	C2_Reboiler	level	0.01977
12	Reactor inlet stream	temperature	0.01475

Table 5.1 (Continued) Ranked CVs and IAE summation result

Rank	Variables		SUM IAE
13	Reactor outlet stream	temperature	0.01289
14	Total Benzene stream	molar flow rate	0.01147
15	Propylene Fresh feed stream	molar flow rate	0.00655
16	C1_Condenser	level	0.00617
17	C2_Condenser	level	0.00524

The control scheme for the benzene column (C1) and cumene column (C2) are standard stand-alone schemes. The controlled variables are Tray temperature, pressure condenser, condenser level and reboiler level.

Table 5.2 and Figure 5.1 show the IAE summation result of Tray temperature for benzene column from Fixture Point method to select the appropriate controlled variables of Tray temperature for benzene column from a candidate output deviation. The temperature on Tray 10 for benzene column is the appropriate controlled variables for all designed control structures because it is the most sensitive.

Figure 5.2 shows the temperature gradient for benzene column (this value is the slope value of Tray temperature for benzene column from the steady state value). The appropriate controlled variables of Tray temperature for benzene column from Fixture Point method is similar when compare with slope value of Tray temperature for benzene column from the steady state value. Figure 5.1 and Figure 5.2 show the similar appropriate controlled variables of Tray temperature for benzene column.

Table 5.2 IAE summations result of tray temperature for Benzene column

Rank	Variable		SUM IAE
1	Tray-1	temperature	0.02015
2	Tray-2	temperature	0.02365
3	Tray-3	temperature	0.02687
4	Tray-4	temperature	0.02872
5	Tray-5	temperature	0.02801
6	Tray-6	temperature	0.02548
7	Tray-7	temperature	0.03547
8	Tray-8	temperature	0.05690
9	Tray-9	temperature	0.08525
10	Tray-10	temperature	0.09403
11	Tray-11	temperature	0.06924
12	Tray-12	temperature	0.03838
13	Tray-13	temperature	0.02359

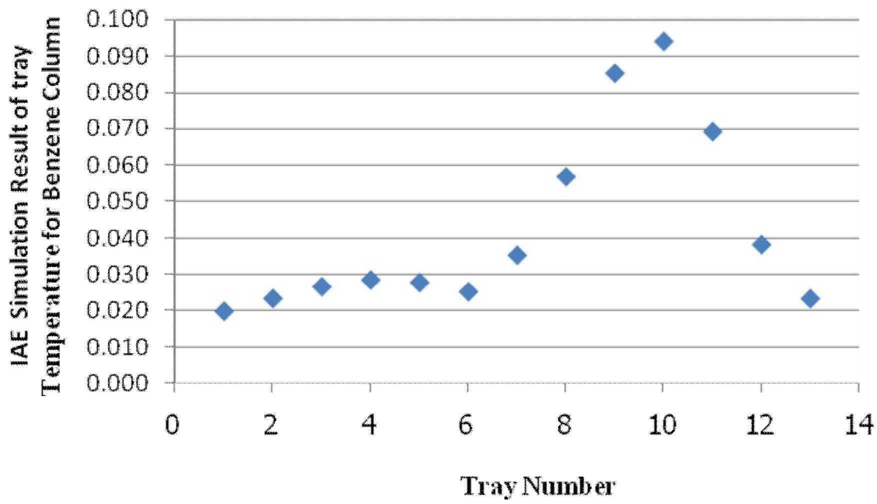


Figure 5.1 IAE summation result of Tray temperature for Benzene column

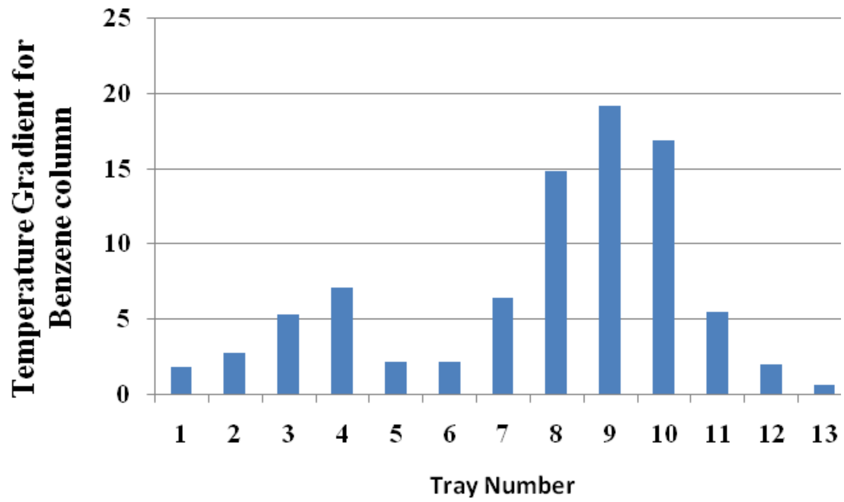


Figure 5.2 Temperature Gradient for Benzene column

Table 5.3 and Figure 5.3 show the IAE summation result of Tray temperature for cumene column from Fixture Point method to select the appropriate controlled variables of Tray temperature for benzene column from a candidate output deviation. The temperature on Tray 15 for cumene column is the appropriate controlled variables for all designed control structures because it is the most sensitive.

Figure 5.4 shows the temperature gradient for cumene column (this value is the slope value of Tray temperature for benzene column from the steady state value). The appropriate controlled variables of Tray temperature for cumene column from Fixture Point method is similar when compare with slope value of Tray temperature for cumene column from the steady state value. Figure 5.3 and Figure 5.4 show the similar appropriate controlled variables of Tray temperature for benzene column

Table 5.3 IAE summations result of tray temperature for Cumene column

Rank	Variable		SUM IAE
1	Tray-1	temperature	0.009
2	Tray-2	temperature	0.009
3	Tray-3	temperature	0.009
4	Tray-4	temperature	0.009
5	Tray-5	temperature	0.009
6	Tray-6	temperature	0.009
7	Tray-7	temperature	0.009
8	Tray-8	temperature	0.010
9	Tray-9	temperature	0.010
10	Tray-10	temperature	0.011
11	Tray-11	temperature	0.011
12	Tray-12	temperature	0.012
13	Tray-13	temperature	0.020
14	Tray-14	temperature	0.046
15	Tray-15	temperature	0.070
16	Tray-16	temperature	0.060
17	Tray-17	temperature	0.031
18	Tray-18	temperature	0.012

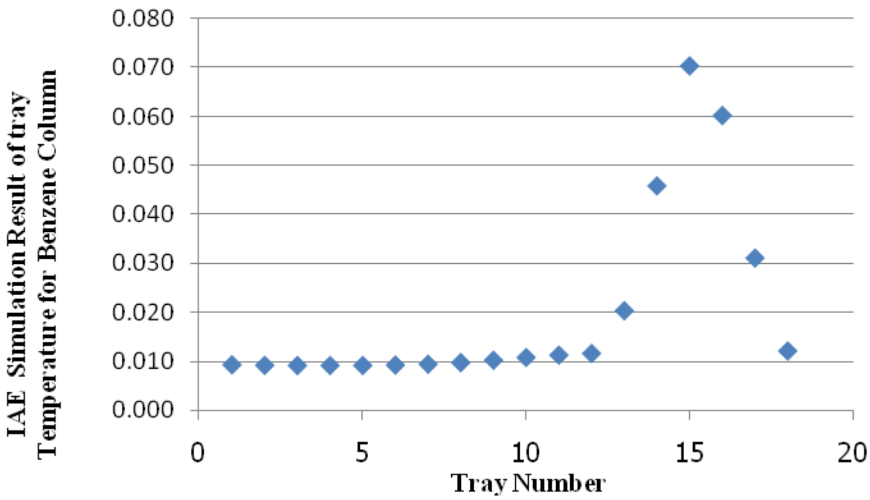


Figure 5.3 IAE summation result of Tray temperature for Cumene column

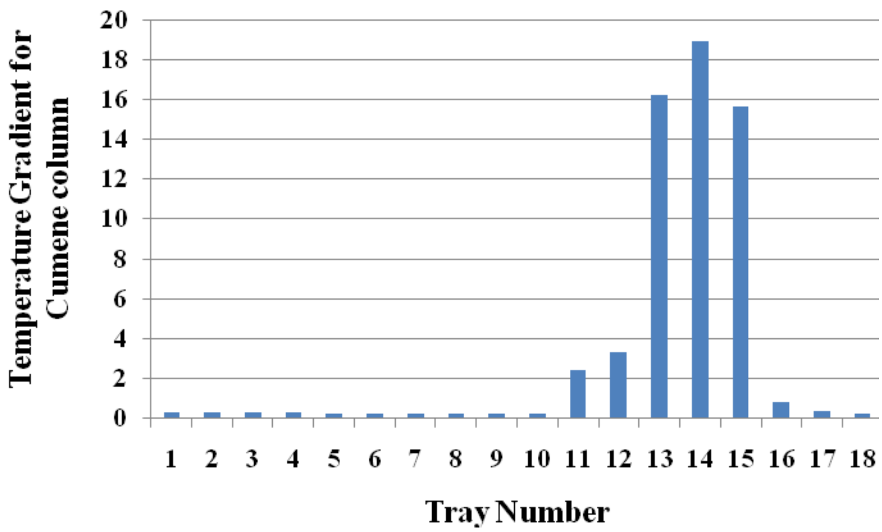


Figure 5.4 Temperature Gradient for Cumene column

Step3. Selection of manipulated variables and measurements via DOF analysis

The manipulated variables and measurements which is selected via DOF analysis has already been shown in Table 5.4.

Table 5.4 Degree of freedom of Cumene process

Unit Operation	Position of Valves	Degree of Freedom
Vaporizer	Fresh feed valve of Benzene (V1), Fresh feed valve of C3(V2) and heating duty of reactor(Qv)	3
Plug flow reactor	heating duty of HX1(Qhx1) and cooling duty of reactor (Qrxn)	2
FEHE heat exchanger	shell out valve (V4), by pass valve*	2
Flash tank	cooling duty of HX2 (Qhx2), separator base valve (V6) and purge valve (V5)	3
Benzene Recycle Column	Distillate (V7), Bottom (V8), Reflux, Condenser duty (Qc1) and Reboiler duty (Qr1)	5
Cumene column	Distillate (V9), Bottom (V10), Reflux, Condenser duty (Qc2) and Reboiler duty (Qr2)	5
Sum		20

Note: by pass valve* is used in new design control structures.

Step4. Energy management via heat exchanger networks

The Energy management via heat exchanger network is described in energy management of heat-integrated process section.

Step5. Selection of control configuration using various tools available

Selection of control configuration use heuristic knowledge. The criteria for selecting an adjustable variable include: causal relationship between the valve and controlled variable, automated valve to influence the selected flow, fast speed of response, ability to compensate for large disturbances and ability to adjust the manipulated variable rapidly and with little upset to the remainder of the process. Table 5.5 shows matching of CVs with MVs of cumene process for all control structures.

Table 5.5 Matching of CVs with MVs

CVs		MVs					
		BC	HEN				
			CS1	CS2	CS3	CS4	CS5
flash tank	LC	V6	V6	V6	V6	V6	V6
C1_Condenser	PC	Qc1	Qc1	Qc1	Qc1	Qc1	Qc1
flash tank	PC	V5	V5	V5	V5	V5	V5
feed stream to HX2	PC	V4	V4	V4	V4	V4	V4
feed stream to Benzene column(C1)	FC	R1/F	R1/F	Fix R1	R1/F	R1/F	R1/F
C1_Reboiler	LC	V8	V8	V8	V8	V8	V8
Benzene Recycle stream	FC	V1	V1	V1	V1	V1	V1
feed stream to cumene column(C2)	FC	R2/F	R2/F	Fix R2	R2/F	R2/F	R2/F
flash tank	TC	Qhx2	Qhx2	Qhx2	Qhx2	Qhx2	Qhx2
C2_Condenser	PC	Qc2	Qc2	Qc2	Qc2	Qc2	Qc2
C2_Reboiler	LC	V10	V10	V10	V10	Qr2	Qr2
Reactor inlet stream	TC	Qhx1	Qhx1	Qhx1	Qhx1	Qhx1	Qhx1
Reactor outlet stream	TC	Qrxn	Qrxn	Qrxn	Qrxn	Qrxn	Qrxn
Propylene Fresh feed stream	FC	V2	V2	V2	V2	V2	V2
C1_Condenser	LC	V7	V7	V7	V7	V7	V7
C2_Condenser	LC	V9	V9	V9	V9	V9	V9
C1_Tray temperature	TC	Qr1	BP	BP	BP	BP	BP
C2_Tray temperature	TC	Qr2	Qr2	Qr2	R2/F	R2/F	V10
Total stream	TC	Qvap	V3	V3	V3	V3	V3

Note: LC refers to liquid level, PC refers to pressure, TC refers to temperature, and BP refers to Bypass valve.

For all control structures, fresh C3 feed is flow controlled by manipulated V2. Total benzene (fresh feed plus benzene recycle D1 from column C1) is ratioed to the C3 flow rate. The fresh feed of benzene is manipulated to achieve the desired flow rate of total benzene. Mixed of Fresh C3 feed and total benzene are preheated by vaporizer. Vaporizer temperature is controlled by vaporizer heat input (Q_{vap}). There is no liquid outlet stream.

For reaction section, reactor inlet temperature is controlled by heat input to HX1 (Q_{hx1}). Reactor exit temperature is controlled by manipulating the pressure (temperature) of the steam being generated on the shell side of the tubular reactor (Q_{rxn}). The back-pressure on the reactor system is controlled by manipulating the control valve located after the heat exchanger (V4). Temperature leaving the condenser HX2 is controlled by heat removal (Q_{hx2}).

For tank section, the level in the flash tank is controlled by manipulating the flow rate of liquid leaving the tank and going to column C1 (V6). The pressure in the flash tank is controlled by manipulating the gas stream (V5).

For distillation section, the control scheme is standard stand-alone. The manipulated variables are condenser duty, distillation valve, reboiler heat input and bottom product valve.

In the CS0 control structure, pressures of all columns are controlled by condenser duty. Base levels of all columns are controlled by bottoms product valve. Reflux drum levels in all columns are controlled by distillate valve. Reflux flow rates are ratioed to the column feed. The temperature on stage 10 in column C1 and the temperature on stage 15 in column C2 are controlled by manipulating the reboiler heat input in each column.

In the first control structure (CS1), 10^{th} tray temperature of benzene recycle column are controlled by bypass valve. Total stream temperature is controlled by V3. Both control variables are used in all of new design control structure. Moreover, 16^{th} tray temperature of cumene column (C2 column) is controlled by reboiler duty. Condenser pressure of cumene column (C2 column) is controlled by compressure duty

In the second control structure (CS2), fixed reflux for both columns

In the third control structure (CS3), 16^{th} tray temperature of cumene column (C2 column) is controlled by reflux.

In the fourth control structure (CS4), 16^{th} tray temperature of cumene column (C2 column) is controlled by reflux and Base level in this column is controlled by reboiler duty of cumene column.

In the fifth control structure (CS5), pressure of benzene recycle column (C1 column) is controlled by distillation valve and Reflux drum levels in this column is controlled by condenser duty of cumene column.

Step6. Completing control structure design by checking the component balance

Component balances are particularly important in process with recycle streams because of their integration effect. The specific mechanism or control loop must be identified to guarantee that there will be no uncontrollable build up of any chemical component within the process (Downs drill). In process, the reactant components should not be left in the product stream because of the yield loss and the specification of the desired product purity. Hence we are limited to use of two methods: consuming the reactants by reaction or adjusting their fresh feed flow. Table 5.6 shows the checking of all components in the process by overall mass balance equation.

Table 5.6 Component balance

component	input	generation	output	consumption	accumulation (= inventory) controlled by
Propylene	fresh feed	0	0	eq. 4.2	propylene feed flow control
Propane	fresh feed	0	Propane	0	tank pressure control
Benzene	fresh feed	0	tot/ben	0	1^{st} column temperature control
Cumene	0	eq. 4.1	cumene	eq. 4.2	2^{nd} column temperature control
DIPB	0	eq. 4.2	DIPB	0	

Cumene process has five components to be accounted for: There are propylene, propane, benzene, cumene and diisopropylbenzene. The compositions of two fresh feeds are propylene and benzene. The fresh propylene feed stream contains some propane impurity, and leave out of the process in flash tank. The product cumene leave in distillate stream of C2 column. By-product (diisopropylbenzene) leaved in bottom stream of C2 column. Unreact benzene is combined with the distillate stream of C1 column and recycle back to the reaction section.

Step7. Selection of controller type: single loops or MPC

In this research, all controller type is single-input-single-output loop. There are temperature controller, pressure controller, flow controller and level controller. Temperature controllers are PIDs which are tuned using relay feedback. Pressure controllers and flow controllers are PIs and their parameters are heuristics valves.

Step8. Validation via rigorous dynamic simulation

Using software HYSYS to evaluate performance for Cumene process of all designed control structures at dynamic simulation

5.2 Design of plantwide control structure

In this research, we apply the new design procedure by Wongsri (2009) to all design control structures for cumene process. In all of control structures (CS1, CS2 and CS3) the same loops are used as follows:

Plantwide level

Valve V1 is manipulated to control the recycle flowrate.

Unit level

Plug flow reactor unit

Heating duty of HX1 (Q_{hx1}) is manipulated to control the reactor inlet temperature.

Cooling duty (Q_{rxn}) is manipulated to control the reactor outlet temperature.

Flash tank unit

Cooling duty (Q_{hx2}) is manipulated to control the vessel inlet temperature.

Valve V6 is manipulated to control the liquid level in the vessel.

Valve V5 is manipulated to control the vessel pressure.

Benzene recycle column unit (C1)

Condenser duty (Qc1) is manipulated to control the condenser pressure

Valve V8 is manipulated to control the Base level.

Bypass valve is manipulated to control tray 10^{th} temperature.

Distillation valve of benzene recycle column (V7) is manipulated to control Reflux drum level of benzene recycle column.

Cumene column unit (C2)

Condenser duty (Qc2) is manipulated to control the condenser pressure.

Distillation valve of cumene column (V9) is manipulated to control Reflux drum level of cumene column

Valve V10 is manipulated to control the Base level.

Reboiler duty (Qr2) is manipulated to control tray 15^{th} temperature.

In all of these control structures (CS1, CS2, CS3, CS4 and CS5) the difference loops are use as follows:

5.2.1 Design control structure (CS1)

CS1 and CS0 are having the same control structures

5.2.2 Design of control structure I (CS2)

For benzene recycle column and cumene column, there are fixed reflux for both columns.

5.2.3 Design of control structure II (CS3)

For cumene column, the 16^{th} tray temperature is controlled by manipulating the reflux of cumene column

5.2.4 Design of control structure III (CS4)

There is not much difference between control structures II (CS3) and control structures III (CS4).For cumene column, the 16^{th} tray temperature is controlled by manipulating the reflux of cumene column. Moreover, Base level in this column is controlled by reboiler duty of cumene column.

5.2.5 Design of control structure IV (CS5)

In the fourth control structure IV (CS5), the 16^{th} tray temperature is controlled by manipulating the bottom valve of cumene production. Base level in this column is controlled by reboiler duty of cumene column.

For all of control structures, we apply them to the heat exchanger networks for cumene process as show in Figure 5.5 to Figure 5.10

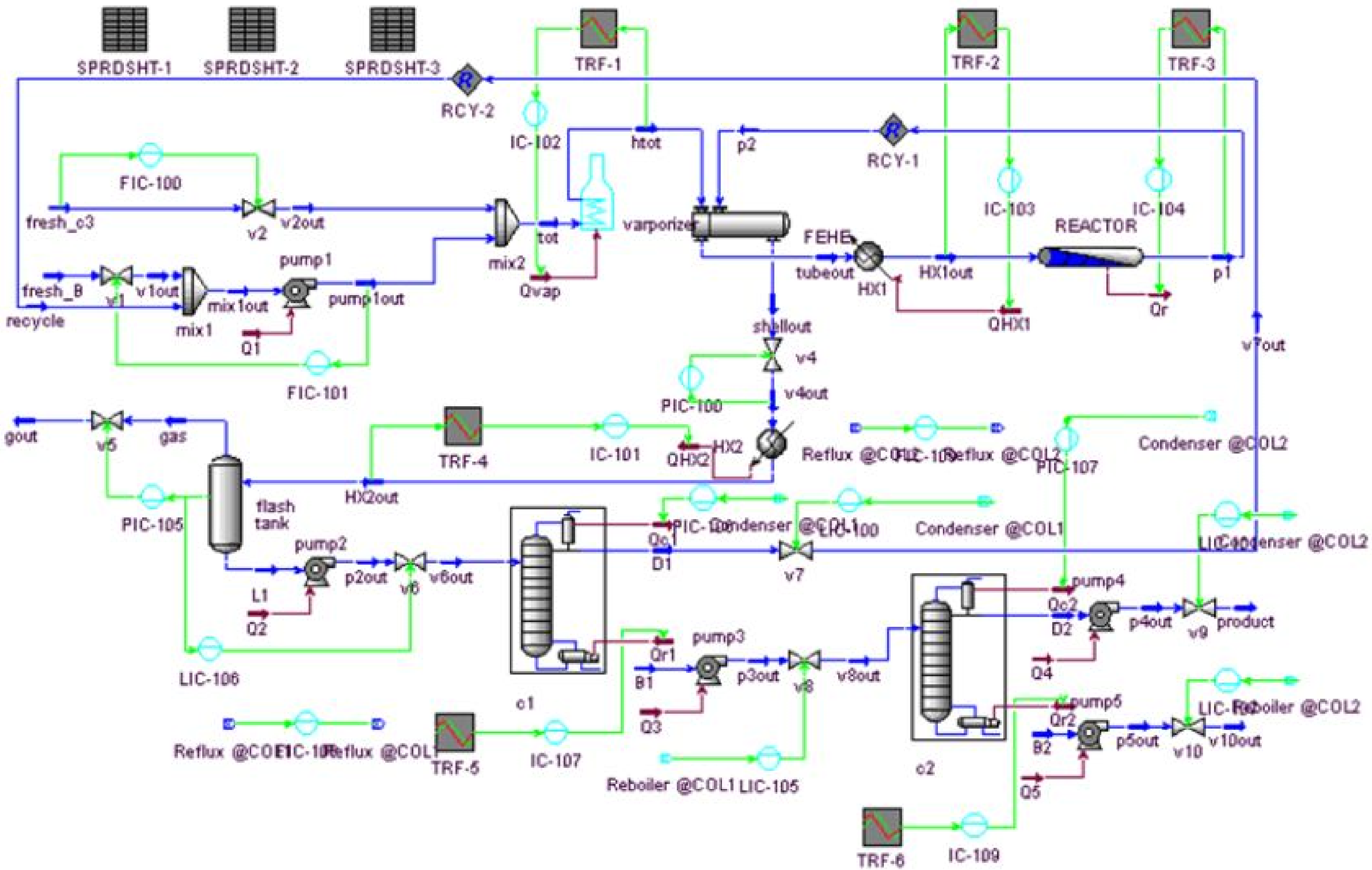


Figure 5.5 Application of control structure (CS0) for cumene process

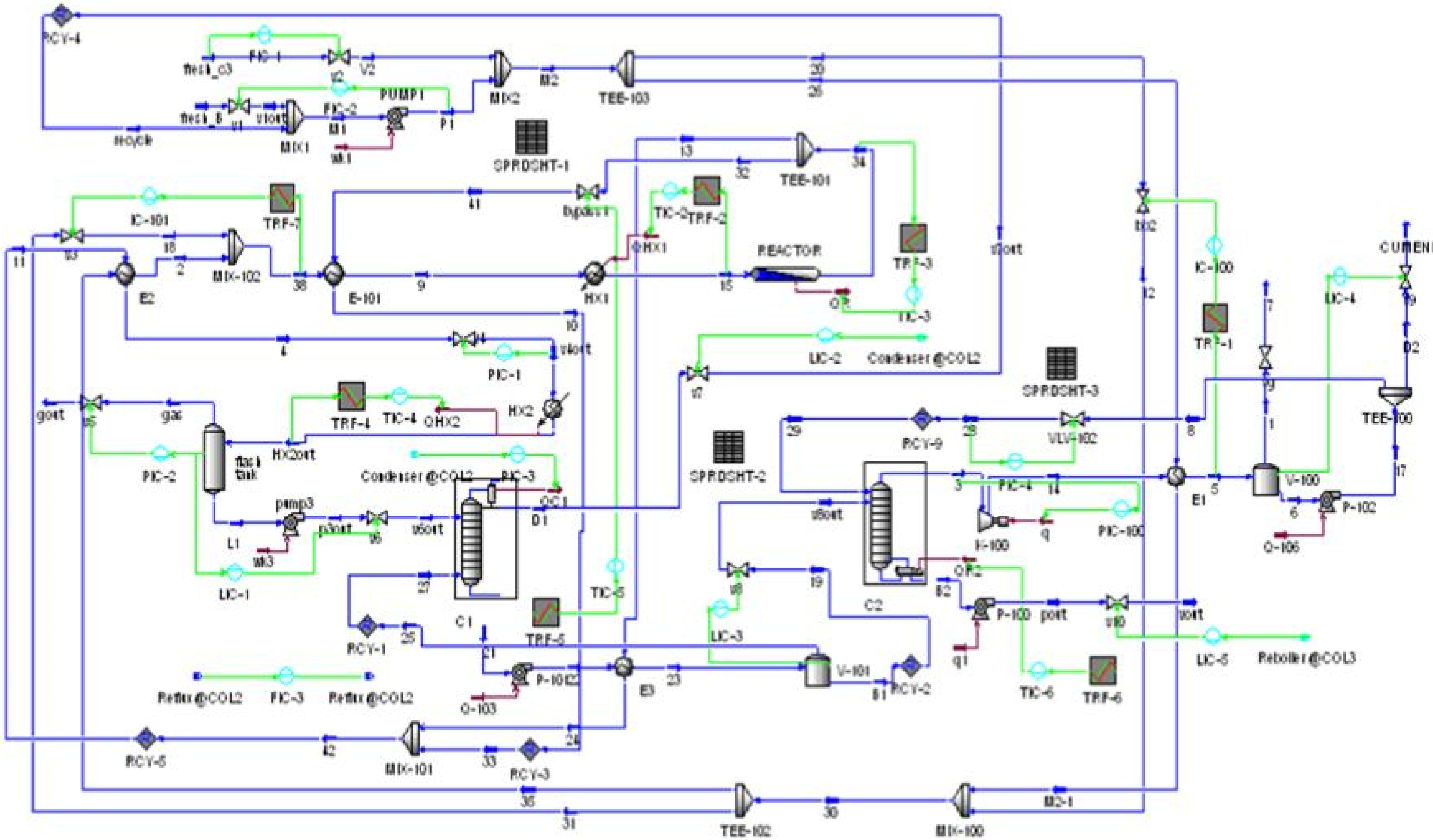


Figure 5.6 Application of control structure (CS1) for cumene process

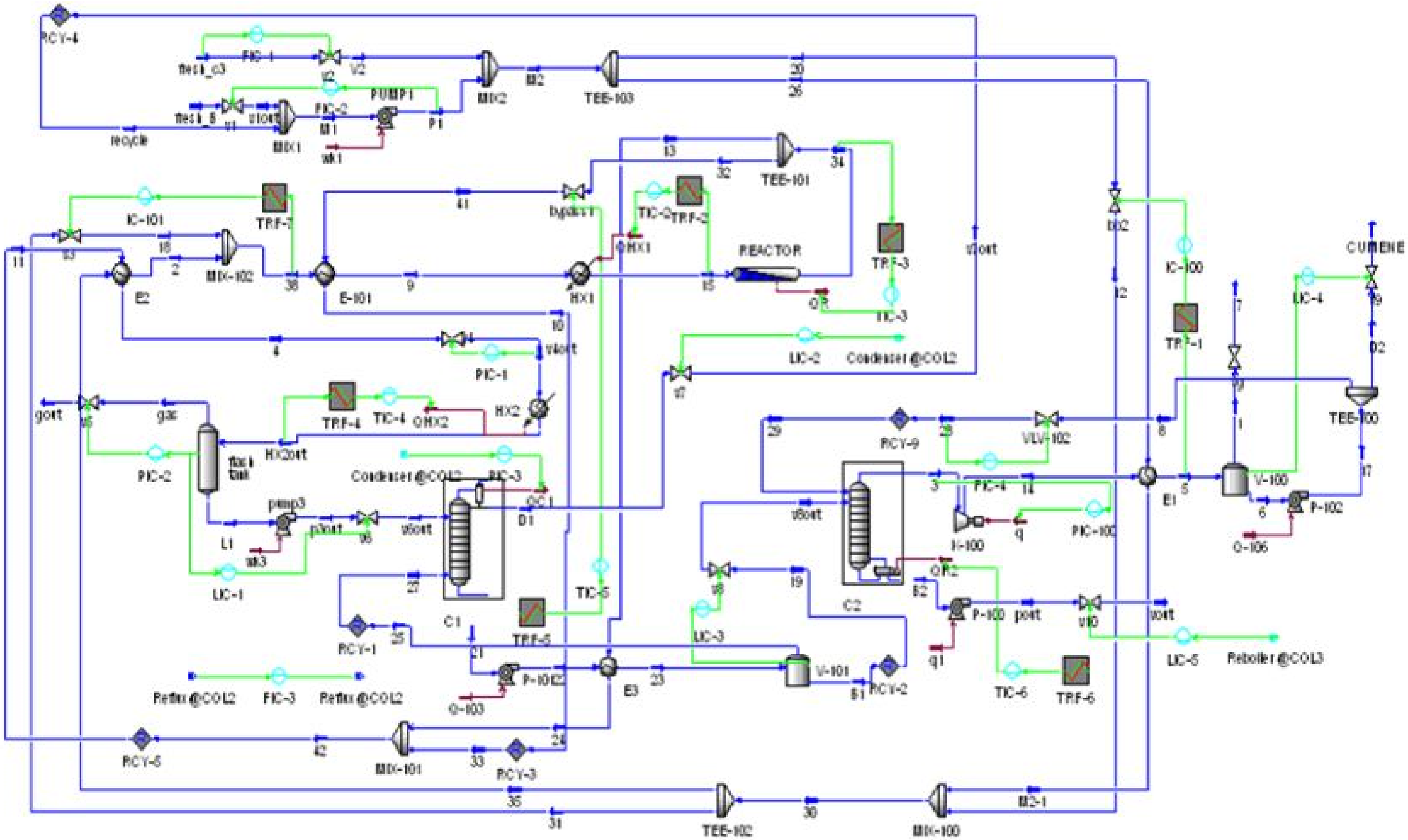


Figure 5.7 Application of control structure (CS2) for cumene process

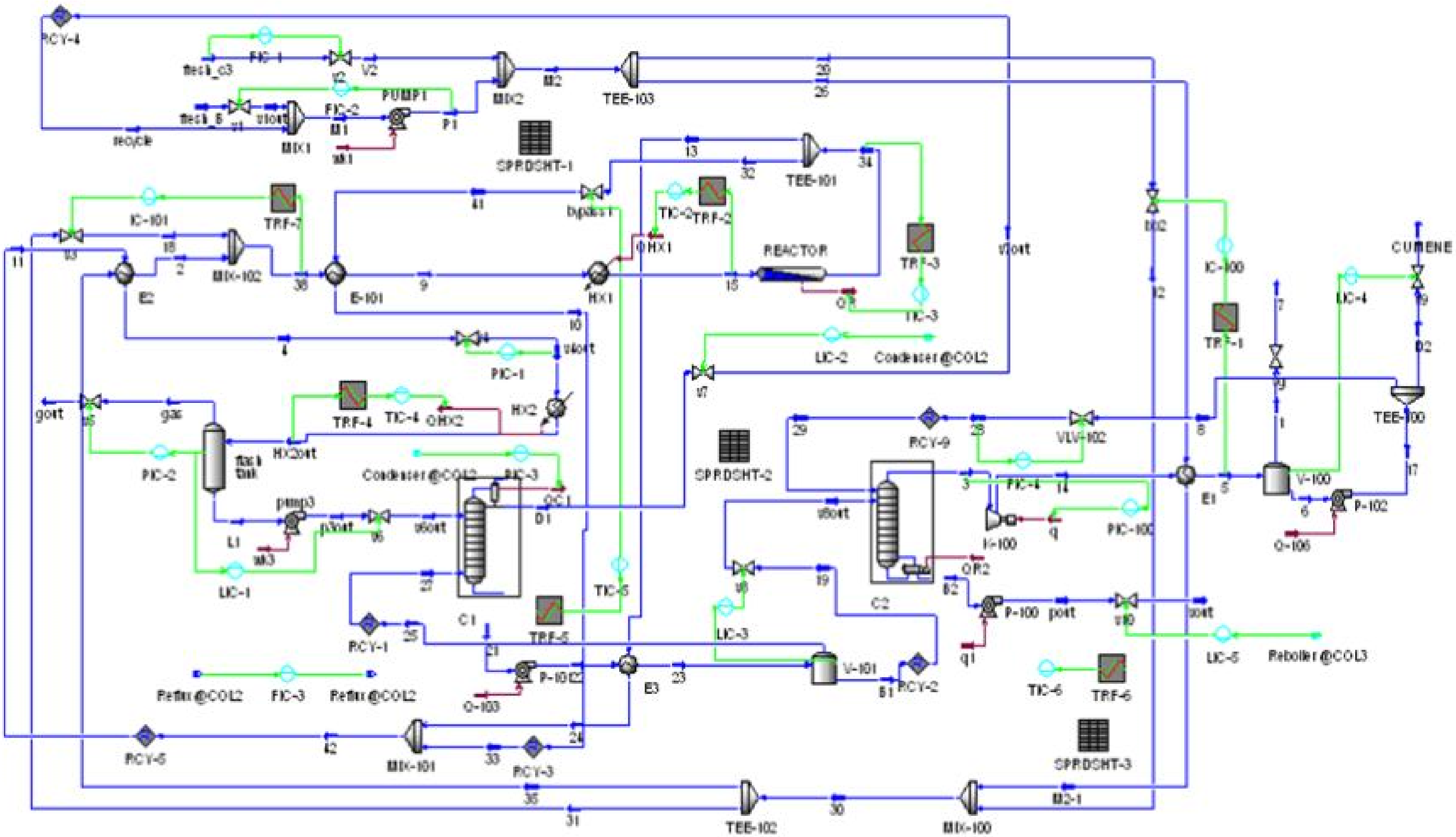


Figure 5.8 Application of control structure (CS3) for cumene process

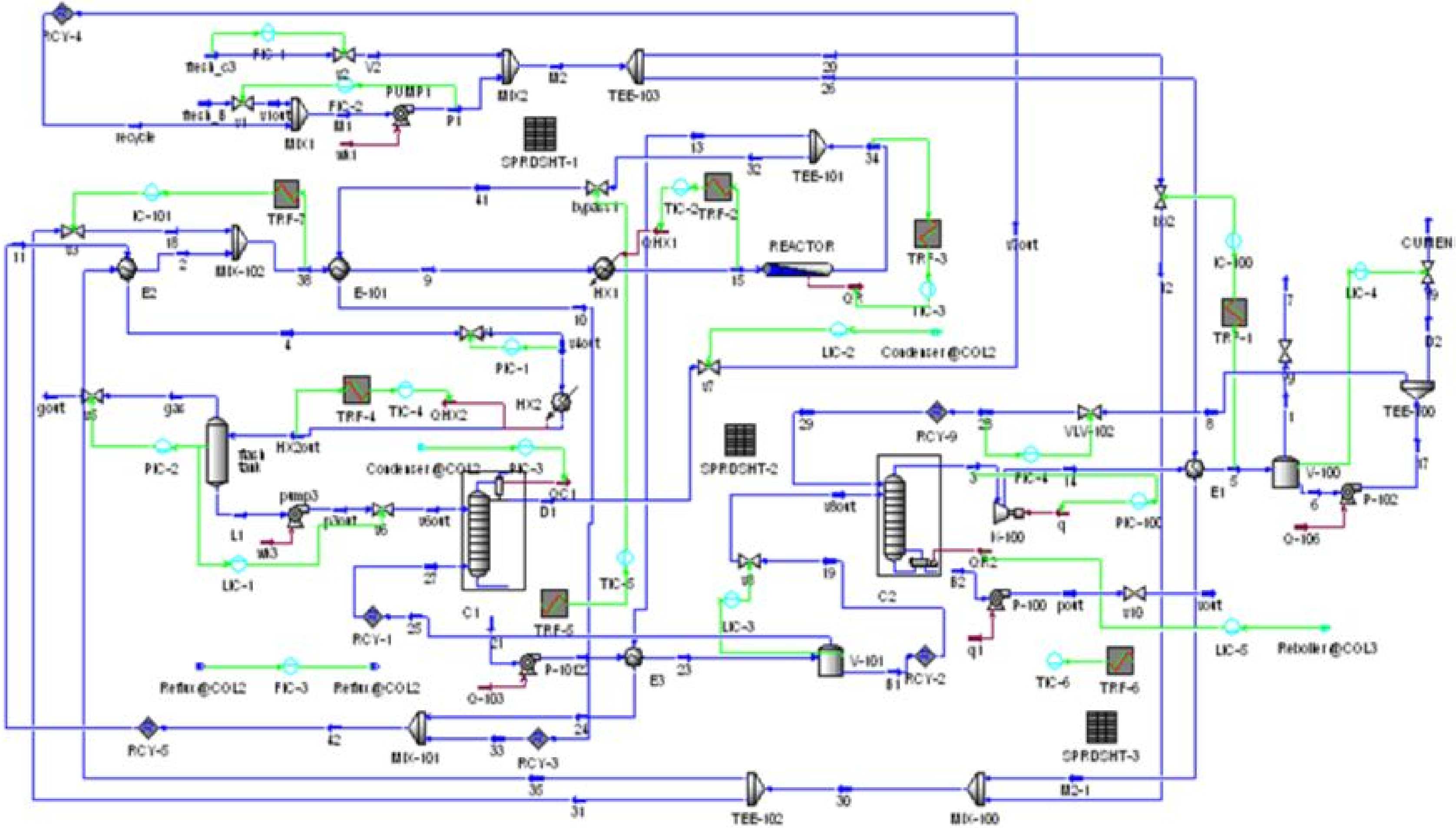


Figure 5.9 Application of control structure (CS4) for cumene process

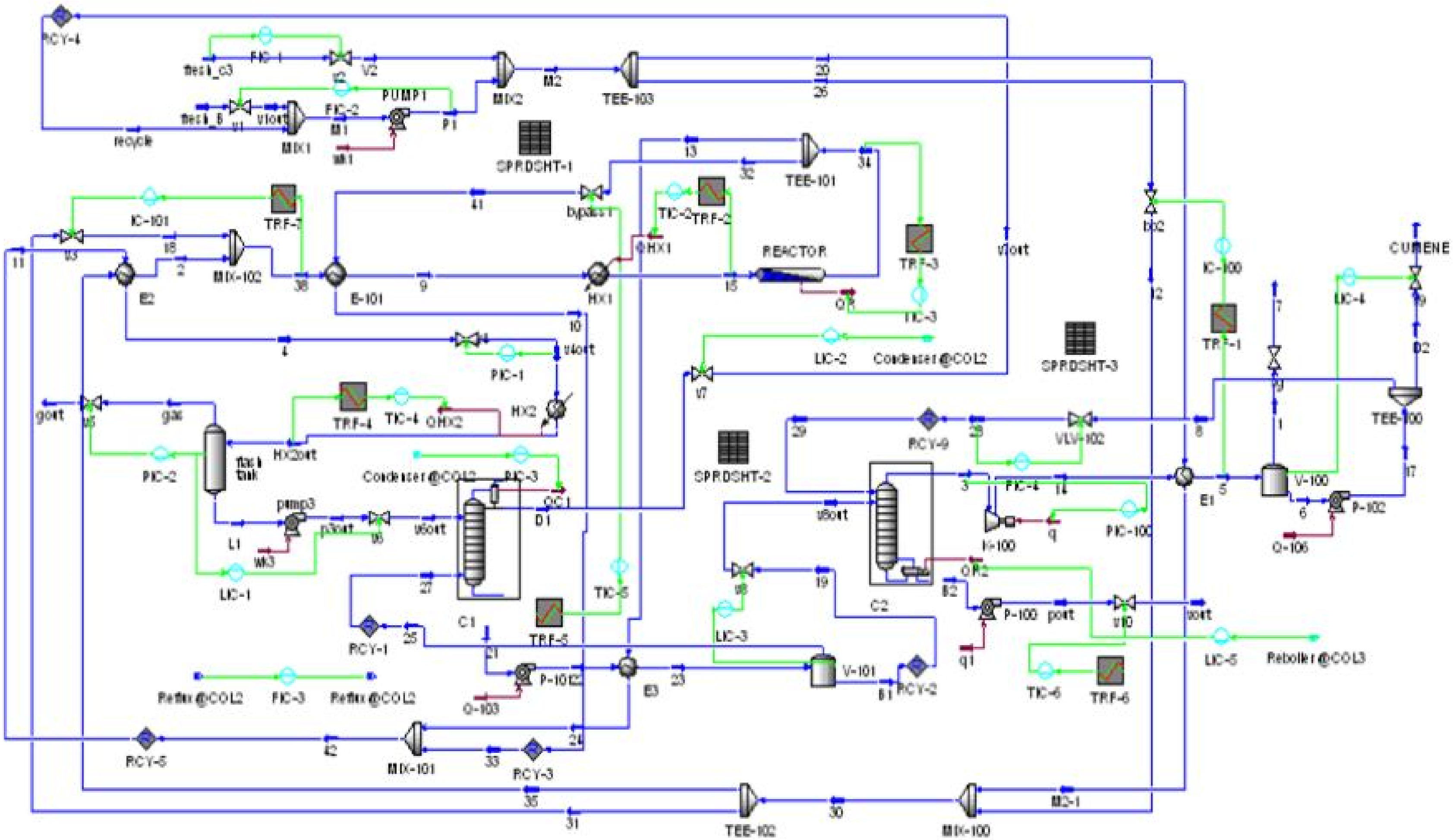


Figure 5.10 Application of control structure (CS5) for cumene process

5.3 Dynamic simulation result

In order to illustrate the dynamic behaviors of our control structures, two types of disturbance: thermal and material disturbances are used to test response of the process: change in propylene fresh feed flow rate increase from101.93 kmol/hr to112 kmol/hr at a time is equal 15 minutes to 600 minutes and decrease from 112 kmol/hr to 91.7 kmol/hr at a time is equal 600 minutes to 1200 minutes and then return to its setpoint at time equals 1200 minutes, change in outlet reactor temperature step increase from 358.5 °C to 365 °C at a time is equal 15 minutes to 600 minutes and decrease from 365 °C to 350 °C at a time is equal 600 minutes to 1200 minutes and then return to its setpoint at time equals 1200 minutes.

Temperature controllers are PIDs which are tuned using relay feedback. Five temperature measurement lags of 0.1 minute are included in the five temperature loops (temperature of the mixture between propylene and benzene, reactor inlet temperature, reactor outlet temperature, 10^{th} tray temperature of benzene (C1) column, 15^{th} tray temperature of cumene (C2) column). Flow and pressure controller are PIs and their parameters are heuristics values. Proportional-only level controllers are used and their parameters are heuristics values. All control valves at nominal operating condition are half-open.

5.3.1 Change in material disturbances of propylene fresh feeds flow rate

Dynamic response of cumene process when the propylene fresh feed flow rate is changed by step increase from101.93 kmol/hr to112 kmol/hr at a time is equal 15 minutes to 600 minutes and decrease from 112 kmol/hr to 91.7 kmol/hr at a time is equal 600 minutes to 1200 minutes and then return to its setpoint at time equals 1200 minutes The dynamic response of all control structures as show in Figure 5.11 to Figure 5.16.

In Figure 5.11 (a) setpoint of propylene fresh feed flow controller is changed as already discussed earlier. Stable regulatory is achieved with the product quality of cumene product being maintained above the desired 99.9% specification as show in Figure 5.11 (m). Moreover, gas flowrate, product flowrate, and undesirable product flowrate are influenced by changing of propylene fresh feed flowrate in the same direction, when propylene fresh feed flowrate is increased they are increased and when propylene fresh feed flowrate is decreased they are decreased too (Figure 5.11(b), Figure 5.11(k), and Figure 5.11(l), respectively). This is because that, when there are changes on the conversion of propylene fresh feed flowrate the ratio between the propylene and propane are changed. The propane is removed from the process in gas stream. Also, the rate of reaction increases product flowrate and undesirable product flowrate are increased. Reactors temperatures and column temperatures are well controlled.

For the design control structure CS1 (HEN-CS1), there are the control structures similar to the CS0. Therefore, the HEN-CS1 is same dynamic response that the CS0. Considering the quality of the product, the result show that product quality of cumene is maintained above the desired 99.9% specification. As show in Figure 5.12 (m)

For the design control structure CS2 (HEN-CS2) benzene recycle column and cumene column are fixed reflux. It has direct impact on the composition. Consider Figure 5.13 (g), which is the dynamic response of the recycle composition, notice that the composition of recycle flowrate change when changing of propylene fresh feed flowrate as well as composition of product cumene in Figure 5.13 (m). It is also maintain product quality of cumene above the desired 99.9% specification.

For the design control structure CS3 (HEN-CS3), the 16^{th} tray temperature of cumene column (C2 column) is controlled by reflux. Notice that, the tray temperature of cumene column is controlled by manipulating the reflux have dynamic response to be smooth than controlled by manipulating the reboiler duty as show in Figure 5.14(o). This design control structure. This design control structure is maintained product quality of cumene above the desired 99.9% specification as shown in Figure 5.14. (m).

For the design control structure CS4 (HEN-CS4), the 16^{th} tray temperature of cumene column (C2 column) is controlled by reflux and reboiler level in this column is controlled by reboiler duty of cumene column. This design control structure is maintained product quality of cumene above the desired 99.9% specification as shown in Figure 5.15. (m).

For the design control structure CS5 (HEN-CS5), the 16th tray temperature of cumene column (C2 column) is controlled by bottom valve of cumene column and level in this column is controlled by reboiler duty of cumene column. Notice that, the tray temperature of cumene column is returned to the setpoint more slowly than other cases as show in Figure 5.16 (o), so top product composition and bottom product composition are returned to the setpoint more slowly than other cases as well as show in Figure 5.16 (m) and Figure 5.16 (n).

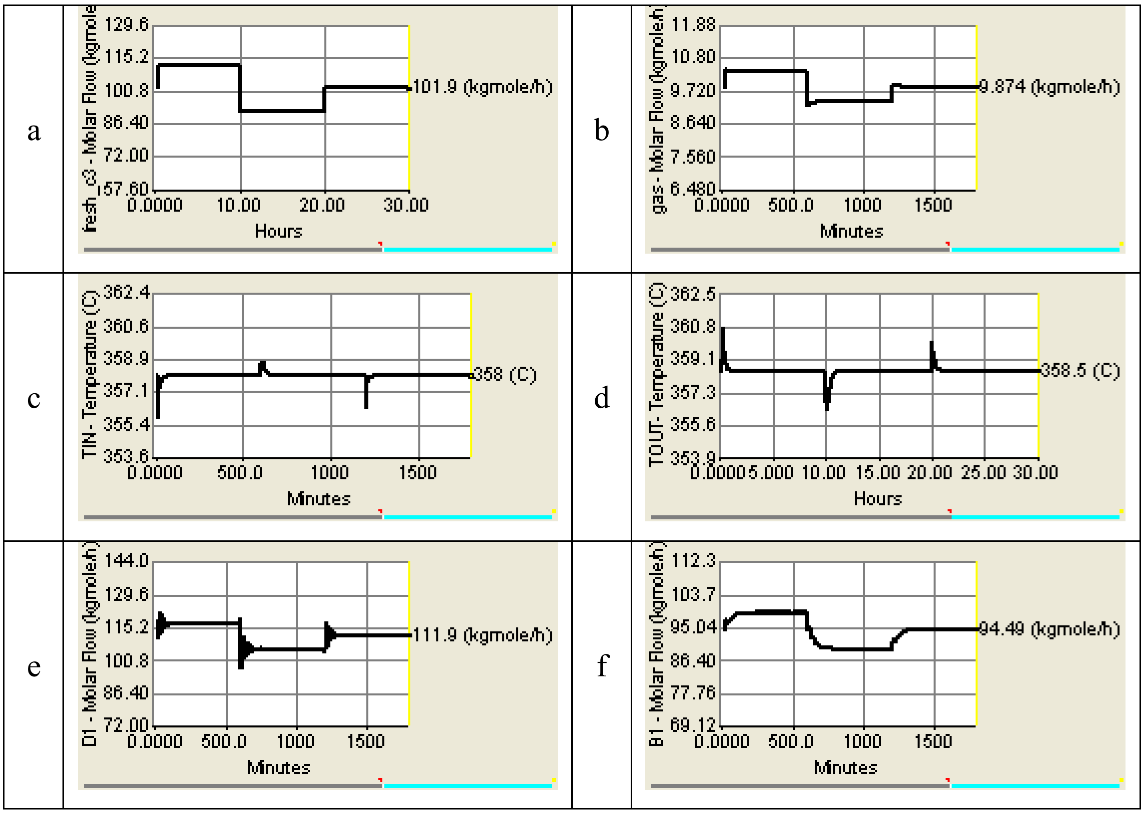


Figure5.11 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of CS0, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column

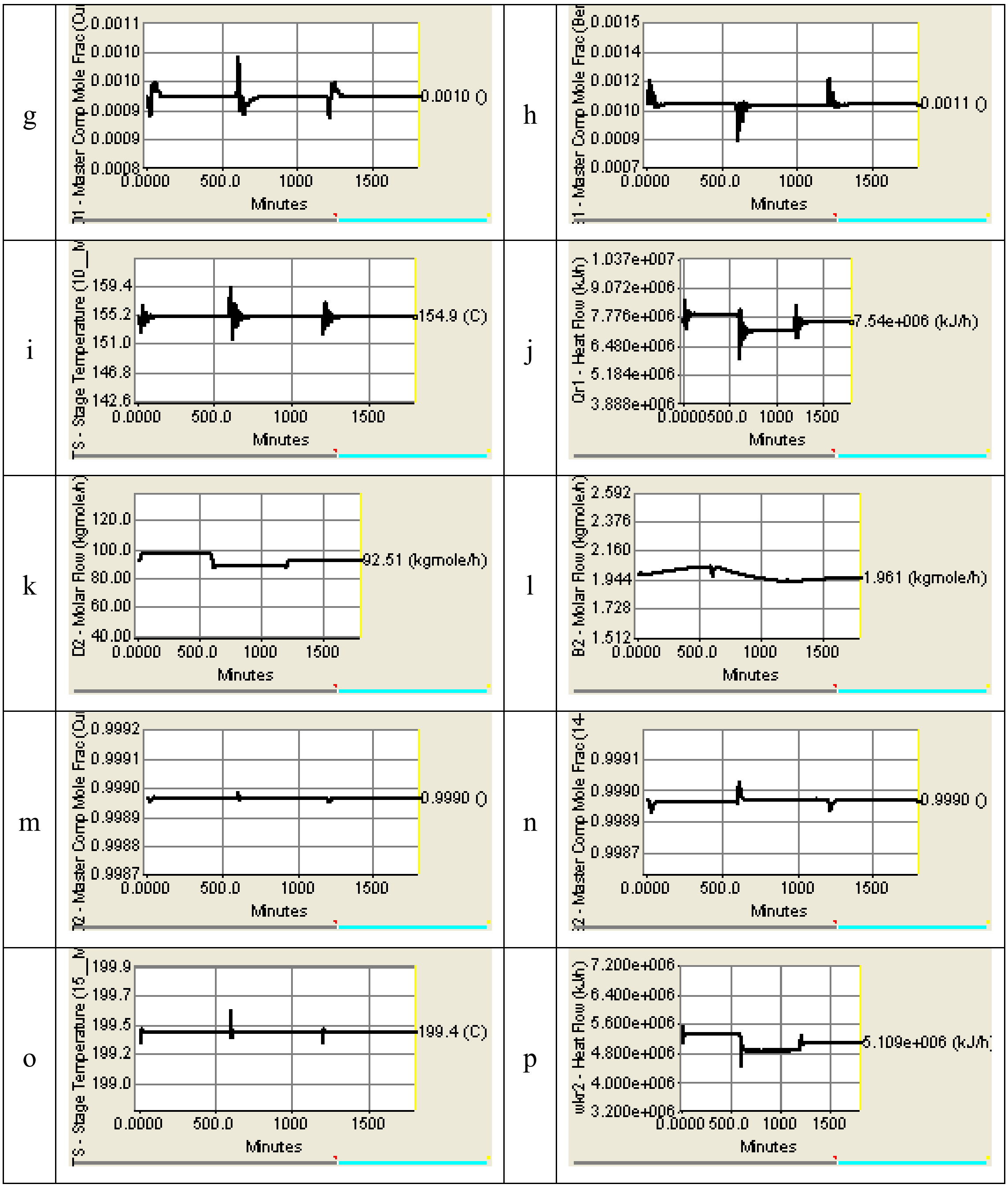


Figure5.11 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of CS0, where (g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column, (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 15th temperature of C2 column, (p) reboiler duty of C2 column

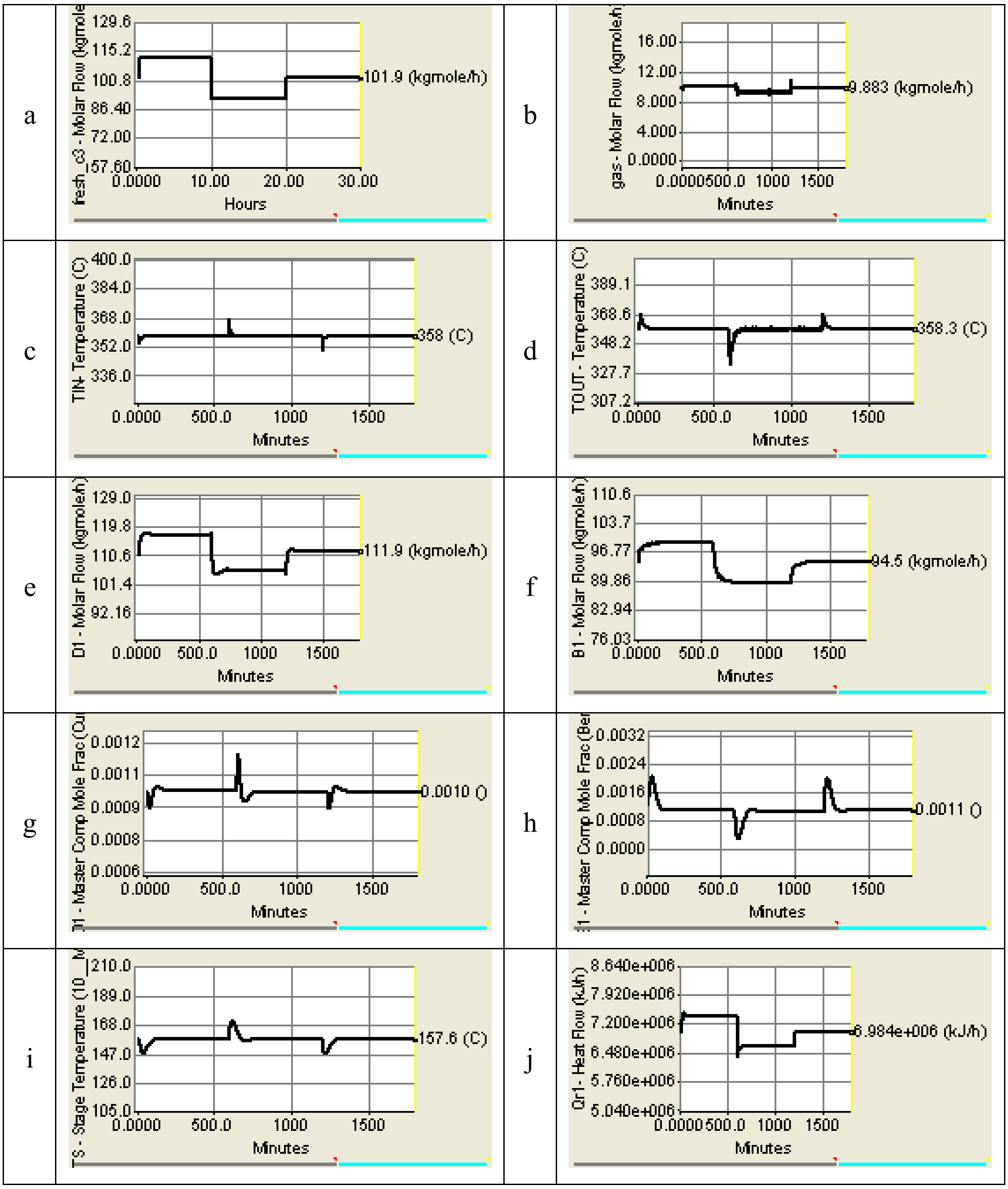


Figure5.12 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS1, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, (g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

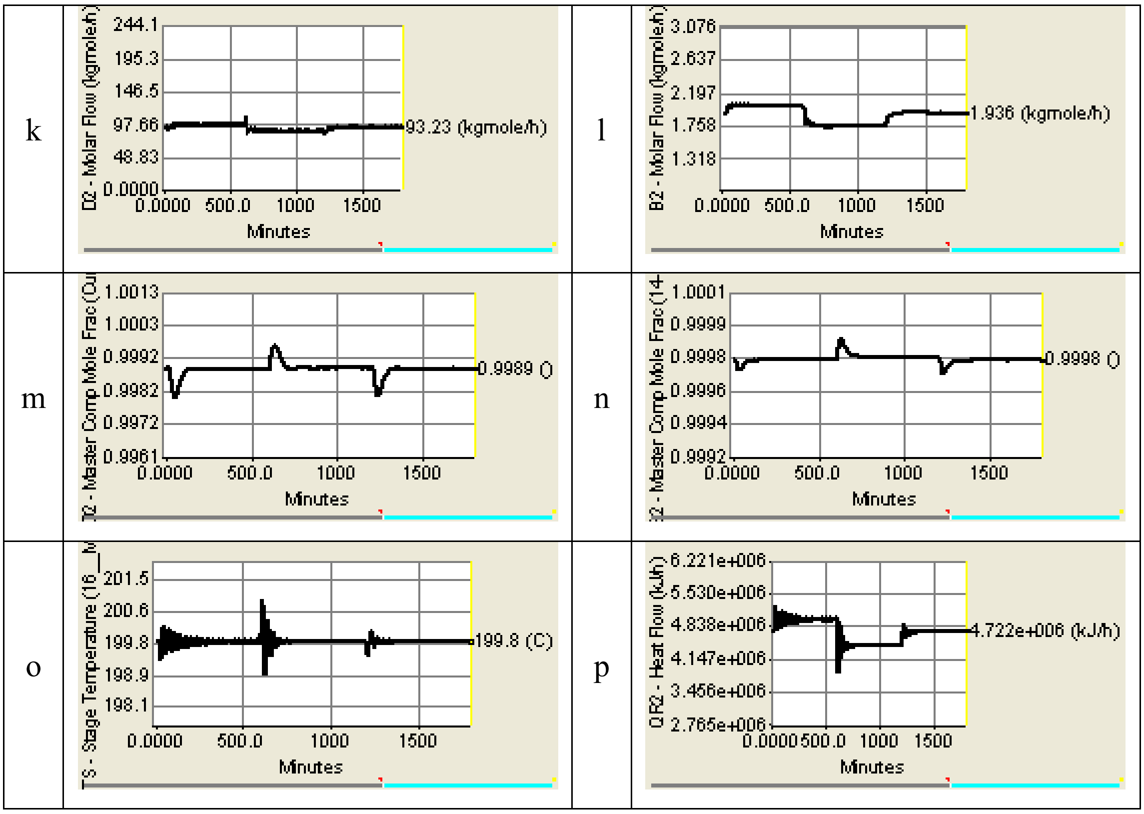


Figure5.12 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS1, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

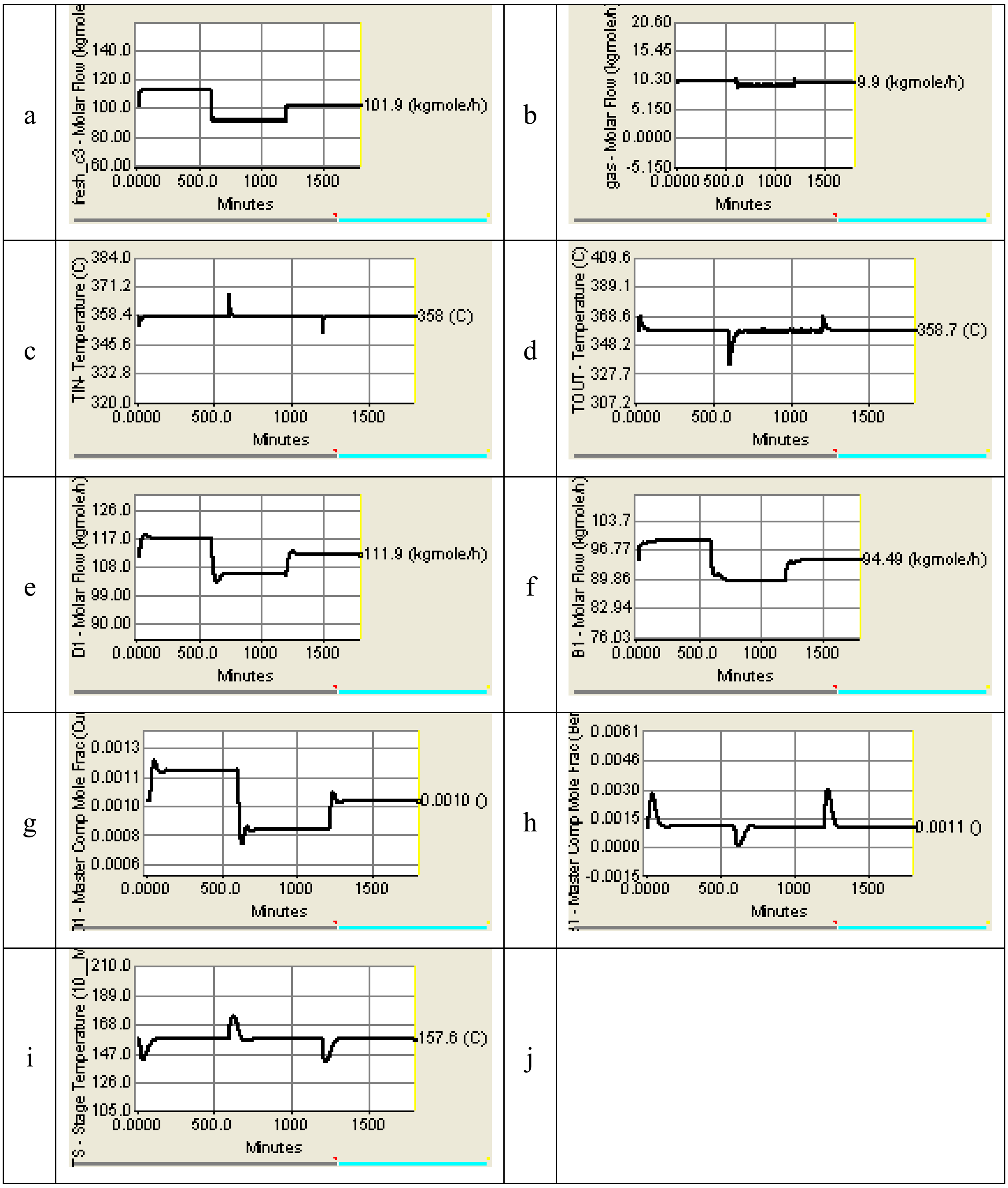


Figure5.13 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS2, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, (g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

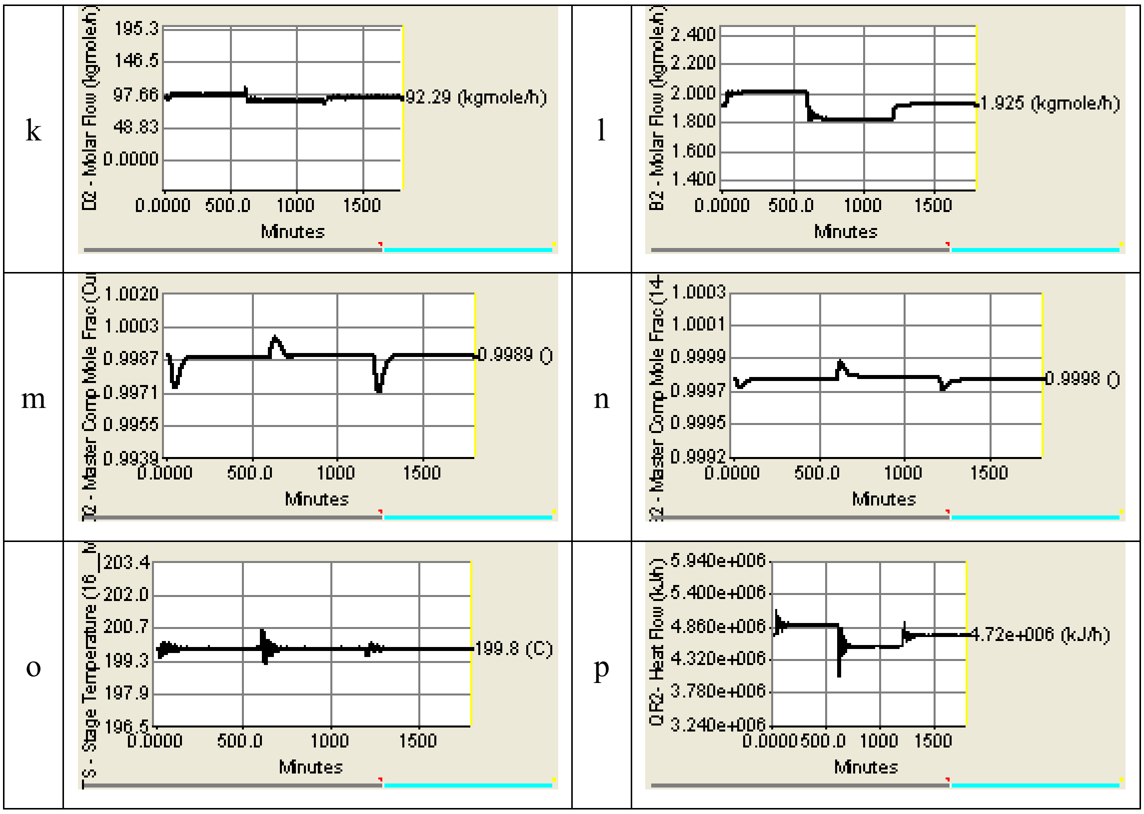


Figure5.13 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN-CS2, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

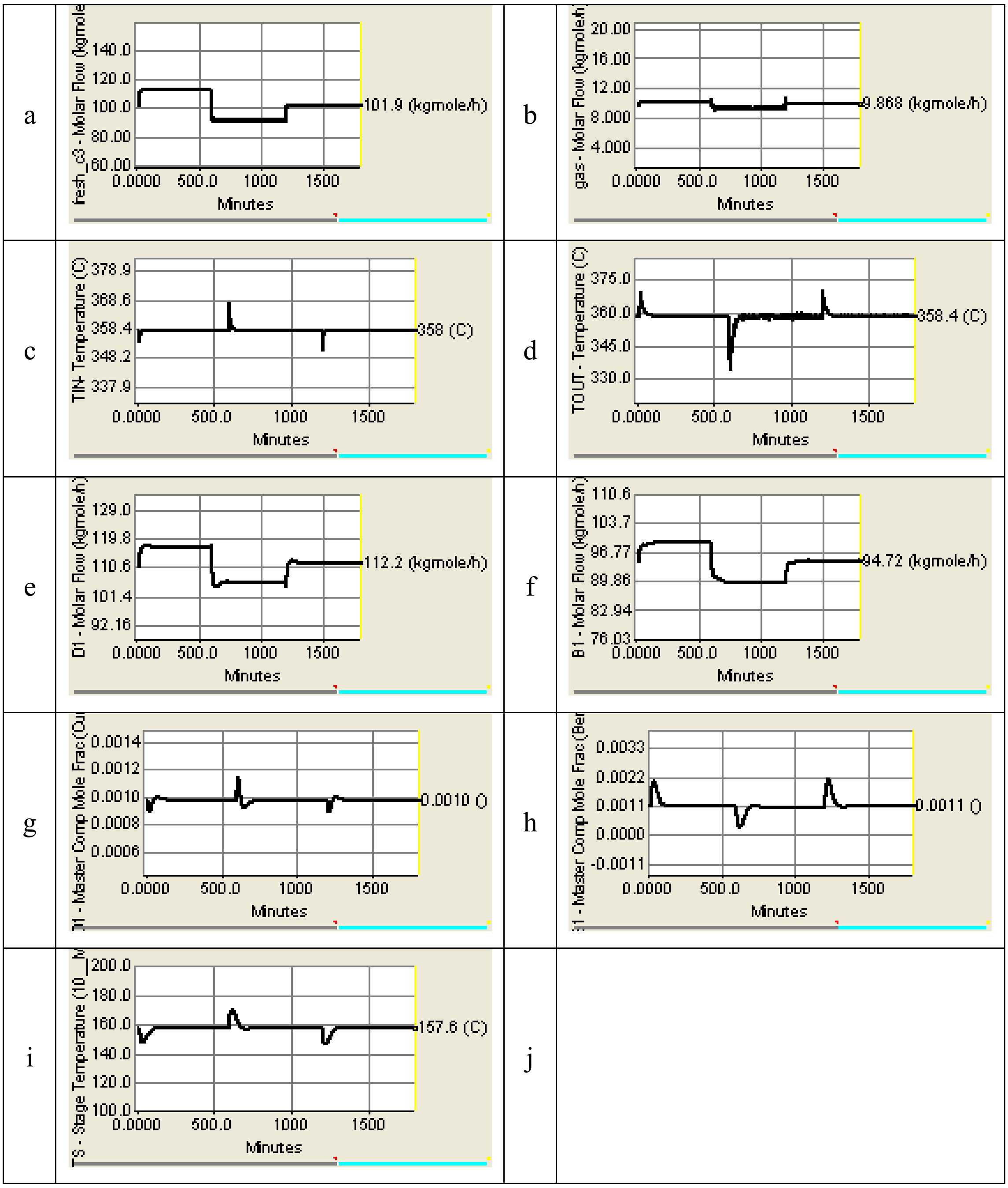


Figure5.14 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS3, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, (g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

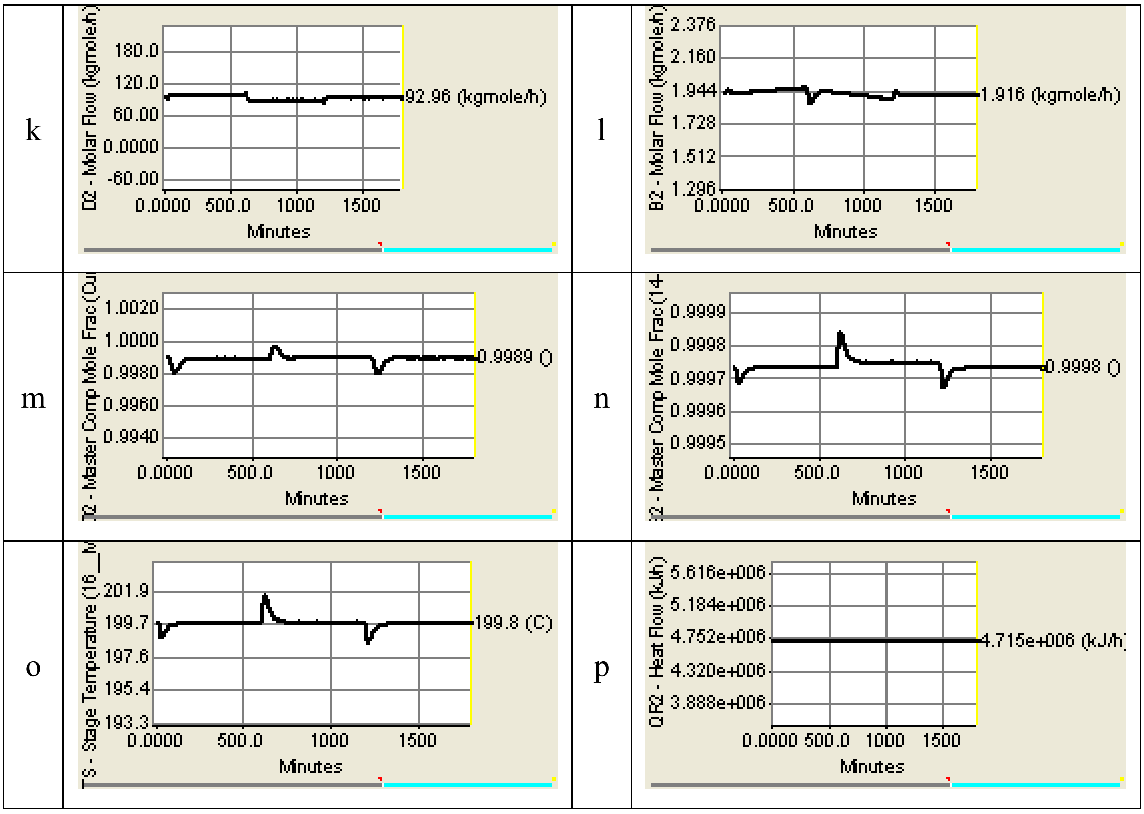


Figure5.14 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS3, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 15th temperature of C2 column, (p) reboiler duty of C2 column

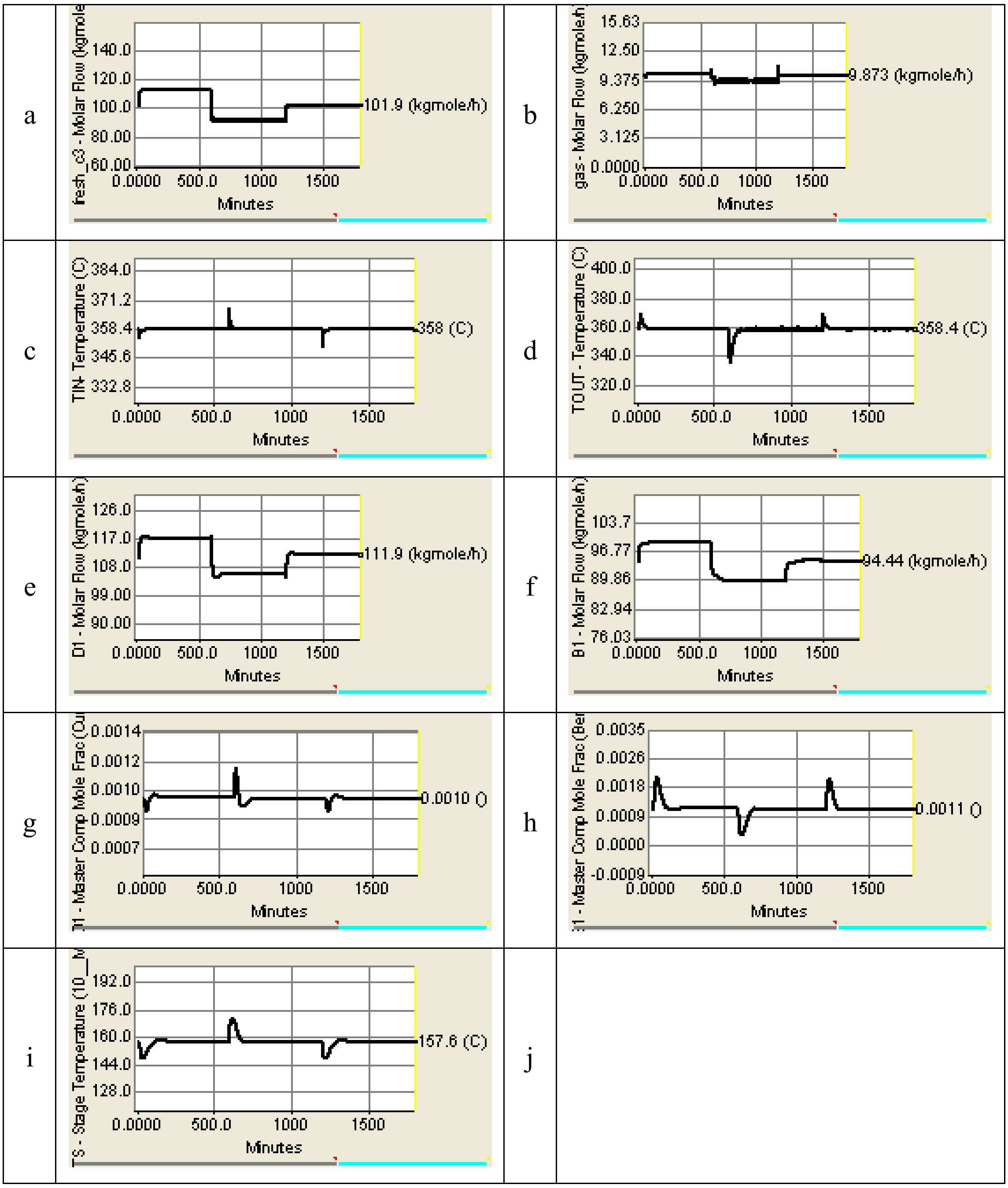


Figure5.15 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS4, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, (g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

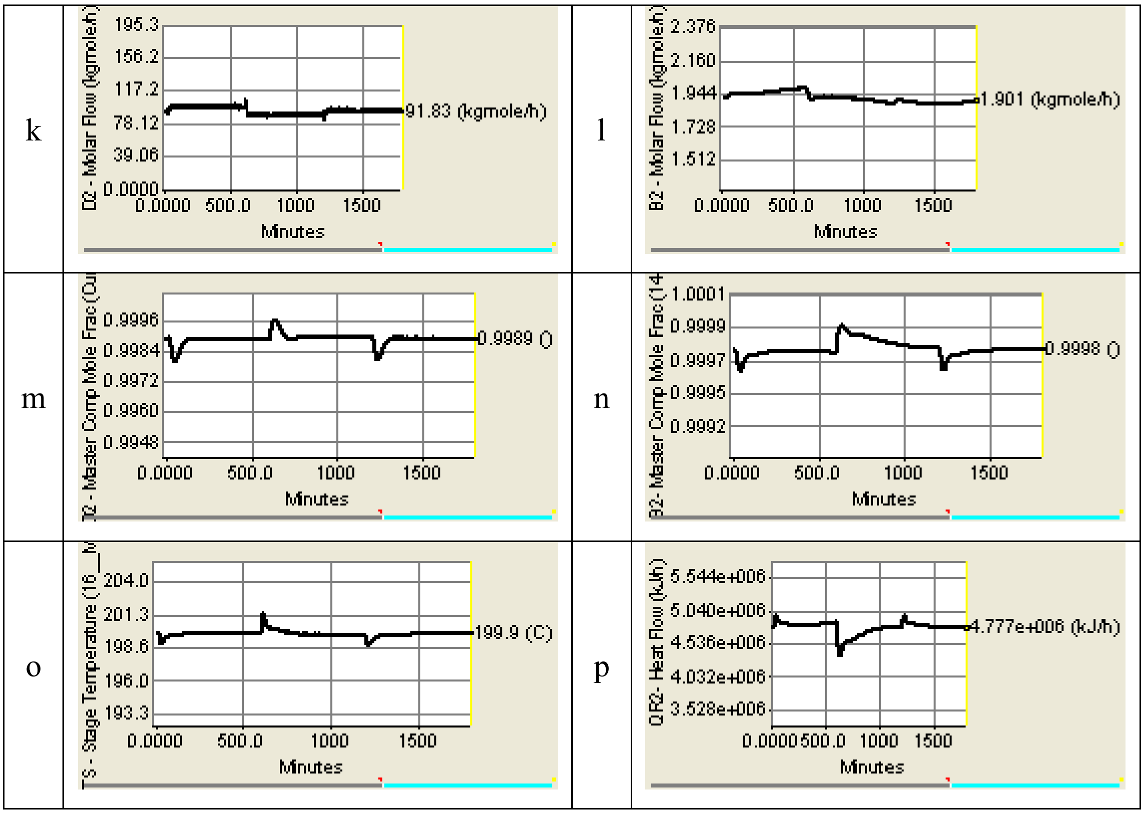


Figure5.15 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS4, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

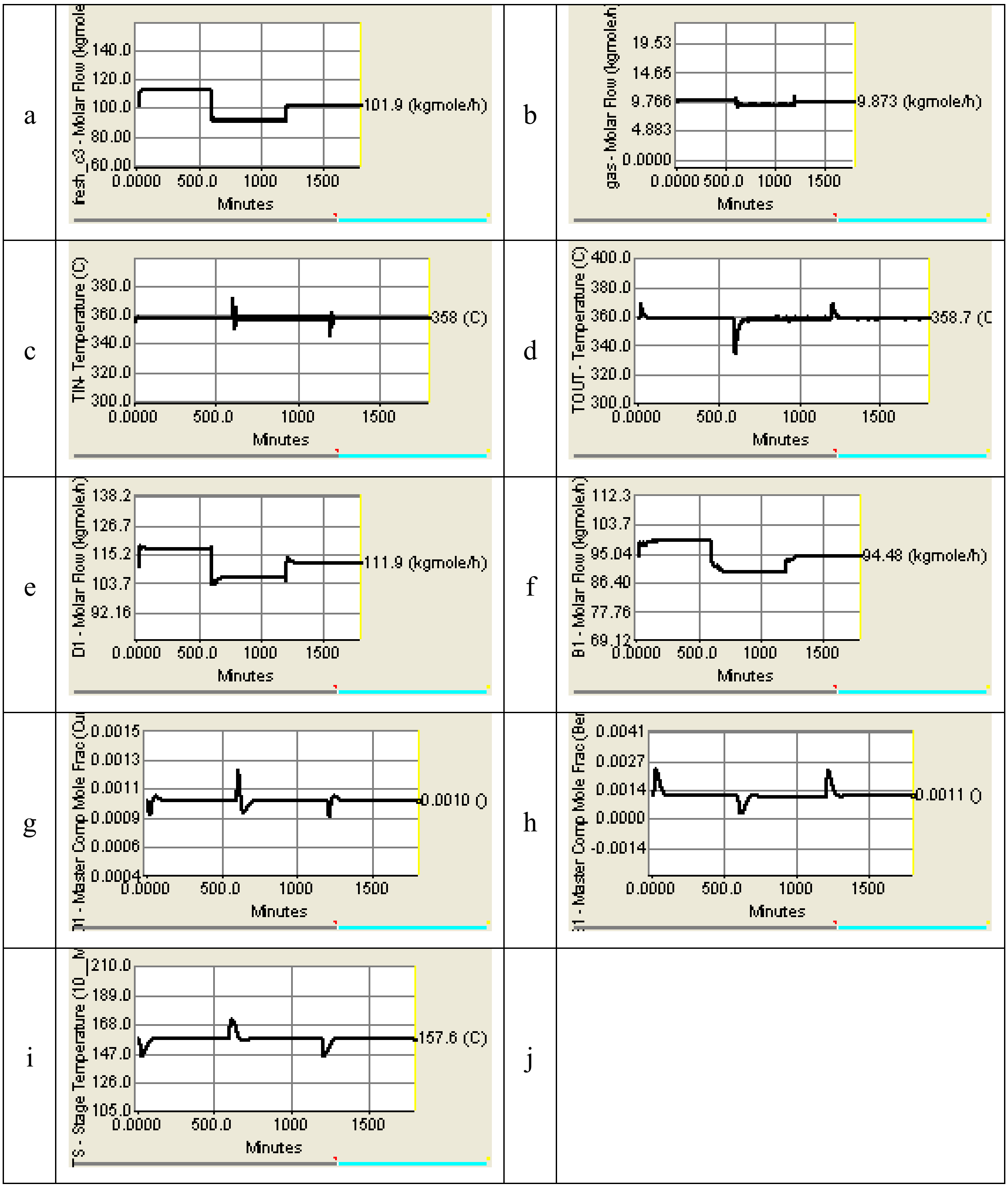


Figure5.16 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS5, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, (g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

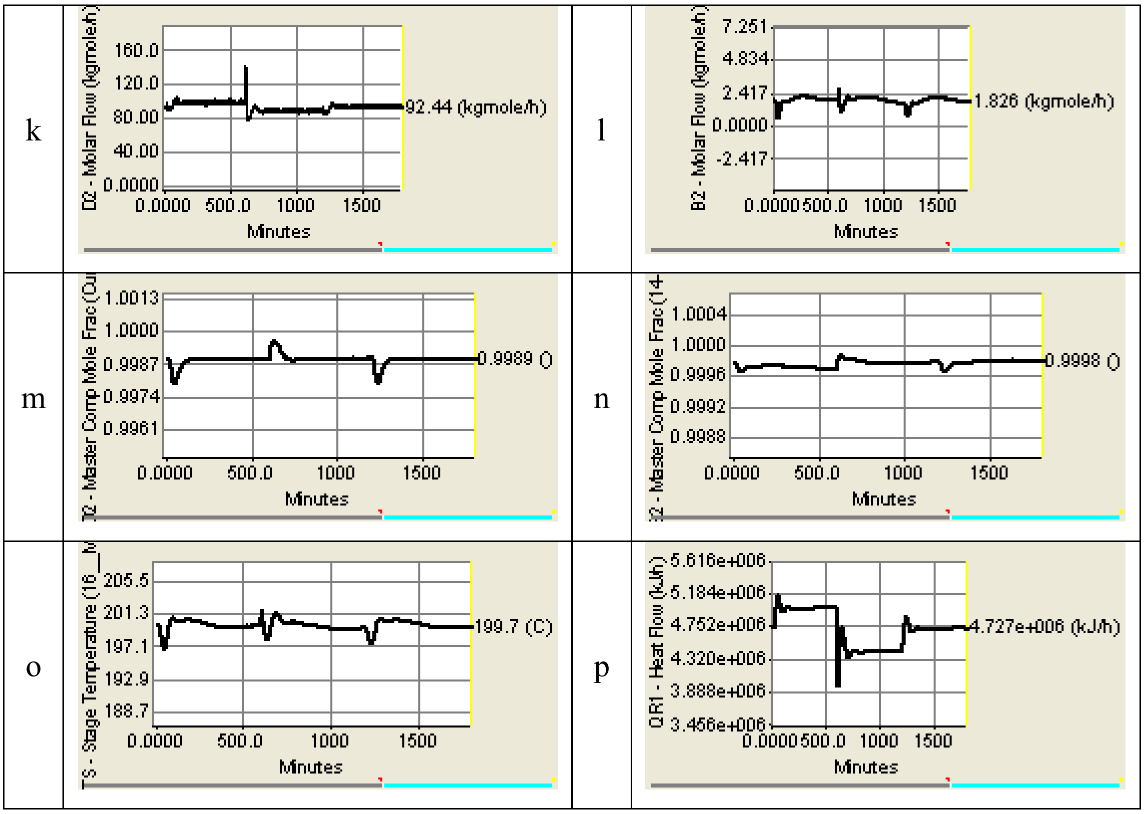


Figure5.16 Dynamic responses for cumene process by step change ±10% in propylene fresh feed flow rate of HEN CS5, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

5.3.2 Change in thermal disturbances of outlet reactor temperature

Dynamic response of cumene process when the outlet reactor temperature is changed by step increase from 358.5 °C to 365 °C at a time is equal 15 minutes to 600 minutes and decrease from 365 °C to 350 °C at a time is equal 600 minutes to 1200 minutes and then return to its setpoint at time equals 1200 minutes. The dynamic response of all control structures as show in Figure 5.17 to Figure 5.22

In Figure 5.17 (d) setpoint of reactor exit temperature controller is changed as already discussed earlier. As expected, increasing reactor temperature reduces more undesirable product. There is little change in both products. There is a significant decrease for a decrease in reactor temperature. This is occurs because the lower conversation of propylene in the reactor produces a large increase in the gas at the first time then return to setpoint as show in Figure 5.17(b). Stable regulatory is achieved with the product quality of cumene product being maintained above the desired 99.9% specification as show in Figure 5.17 (m).

For the design control structure CS1 (HEN-CS1), there are the control structures similar to the CS0. Therefore, the HEN-CS1 is same dynamic response that the CS0. Considering dynamic response of tray temperature of cumene column (Figure 5.18 (o)), the result shows that response of tray temperature of cumene column in control structure CS1 less smooth than CS0.

For the design control structure CS2 (HEN-CS2) benzene recycle column and cumene column are fixed reflux. There are same dynamic responses with control structure CS1

For the design control structure CS3 (HEN-CS3), the 16th tray temperature of cumene column (C2 column) is controlled by reflux. Considering dynamic response of tray temperature of cumene column (Figure 5.19 (o)), notice that the dynamic response of tray temperature of cumene column in control structure CS3 less smooth than all of control structures.

For the design control structure CS4 (HEN-CS4), the 16th tray temperature of cumene column (C2 column) is controlled by reflux and reboiler level in this column is controlled by reboiler duty of cumene column. There are same dynamic responses with control structure CS3 but dynamic response of tray temperature of cumene column of control structure CS4 is smoother than control structure CS3

For the design control structure CS5 (HEN-CS5), the 16th tray temperature of cumene column (C2 column) is controlled by bottom valve of cumene column and level in this column is controlled by reboiler duty of cumene column. Notice that, the tray temperature of cumene column is returned to the setpoint more slowly than other cases as show in Figure 5.22 (o), so top product and bottom product are returned to the setpoint more slowly than other control structures as well as show in Figure 5.22 (k) and Figure 5.22 (l).

However, all of control structures can handle disturbances and can maintain product quality above the desired 99.9% specification.

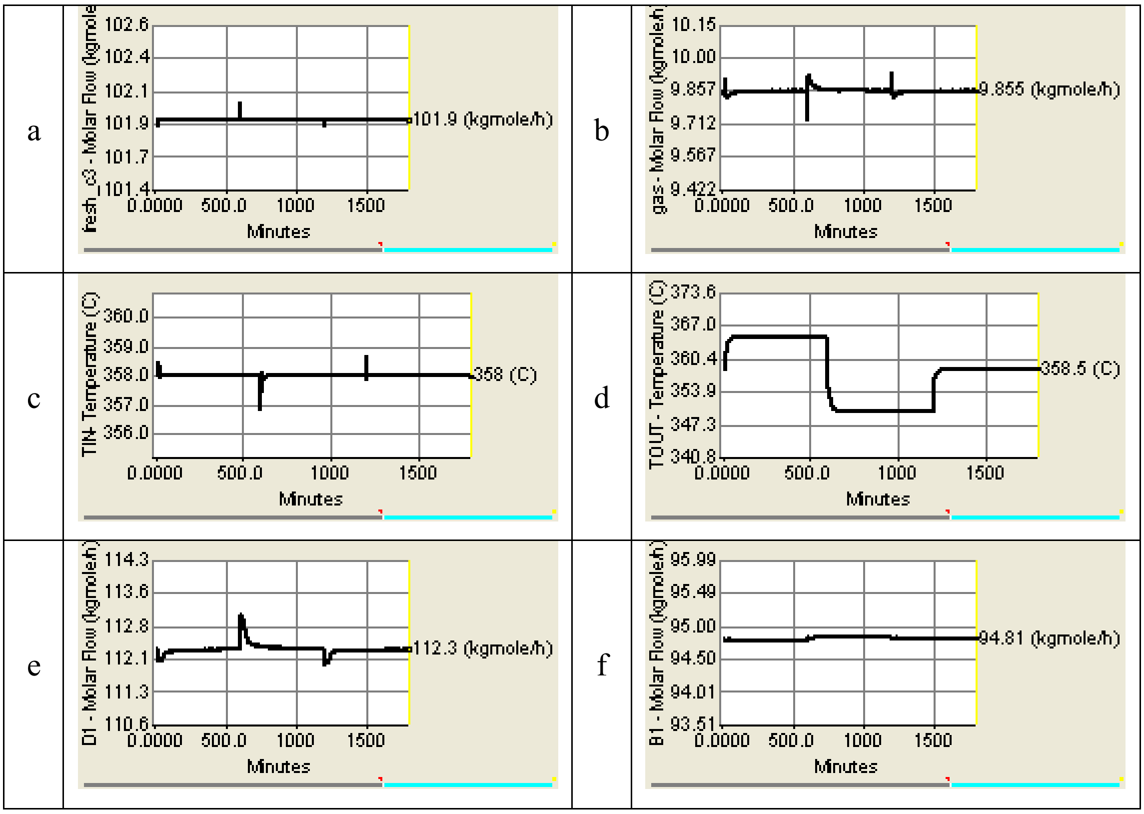


Figure5.17 Dynamic responses for cumene process when the outlet reactor temperature is changed, CS0, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column

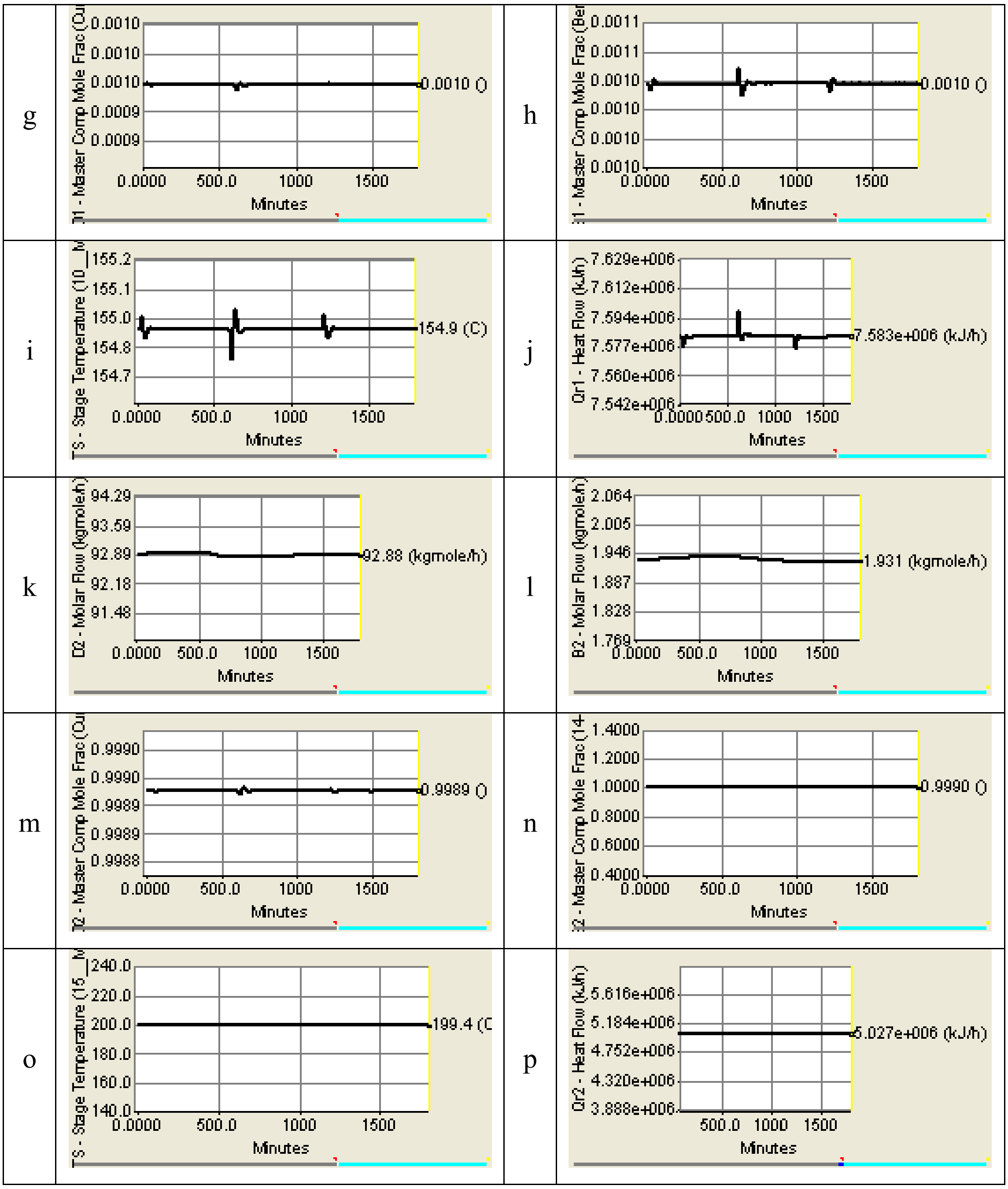


Figure5.17 Dynamic responses for cumene process when the outlet reactor temperature is changed, CS0, where (g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column, (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

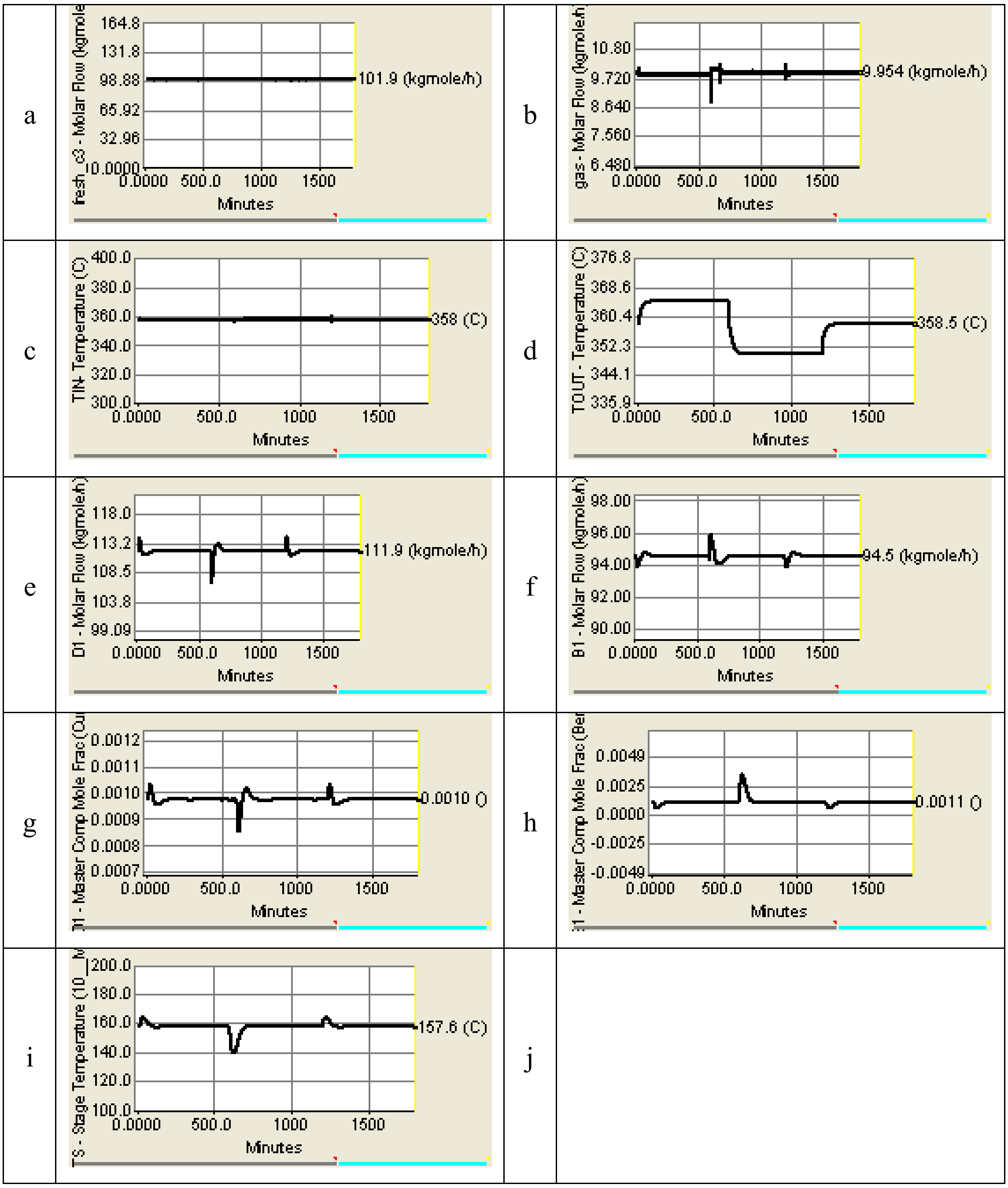


Figure5.18 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS1, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

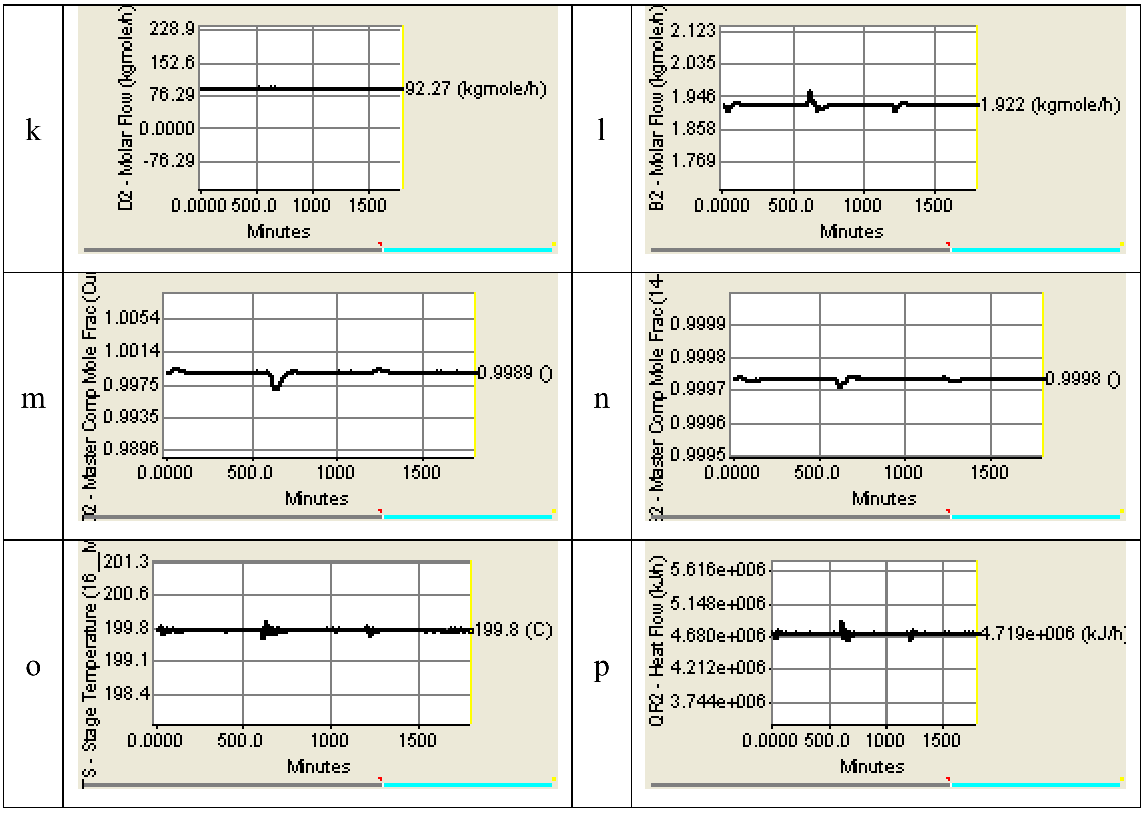


Figure5.18 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS1, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

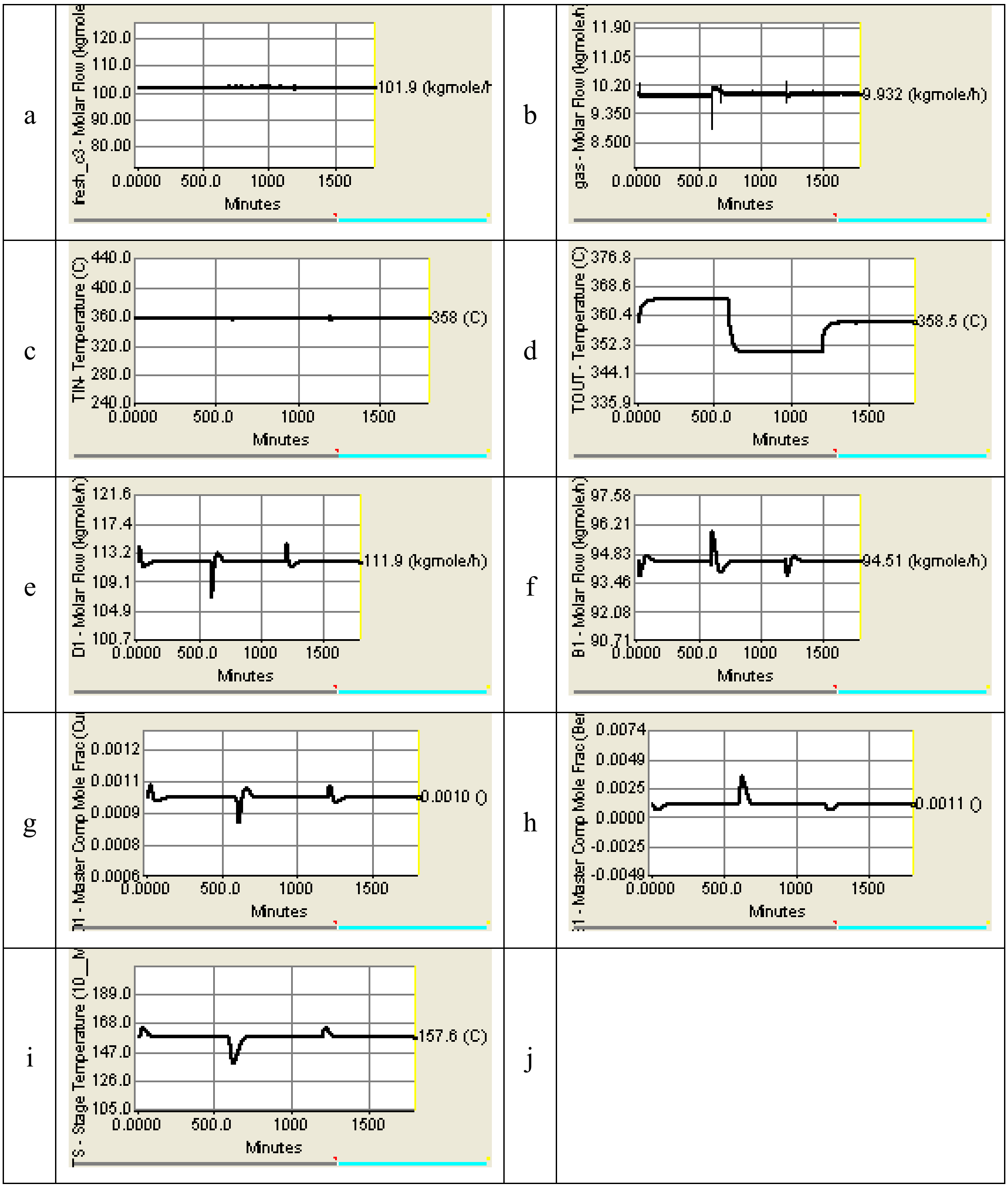


Figure5.19 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS2, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

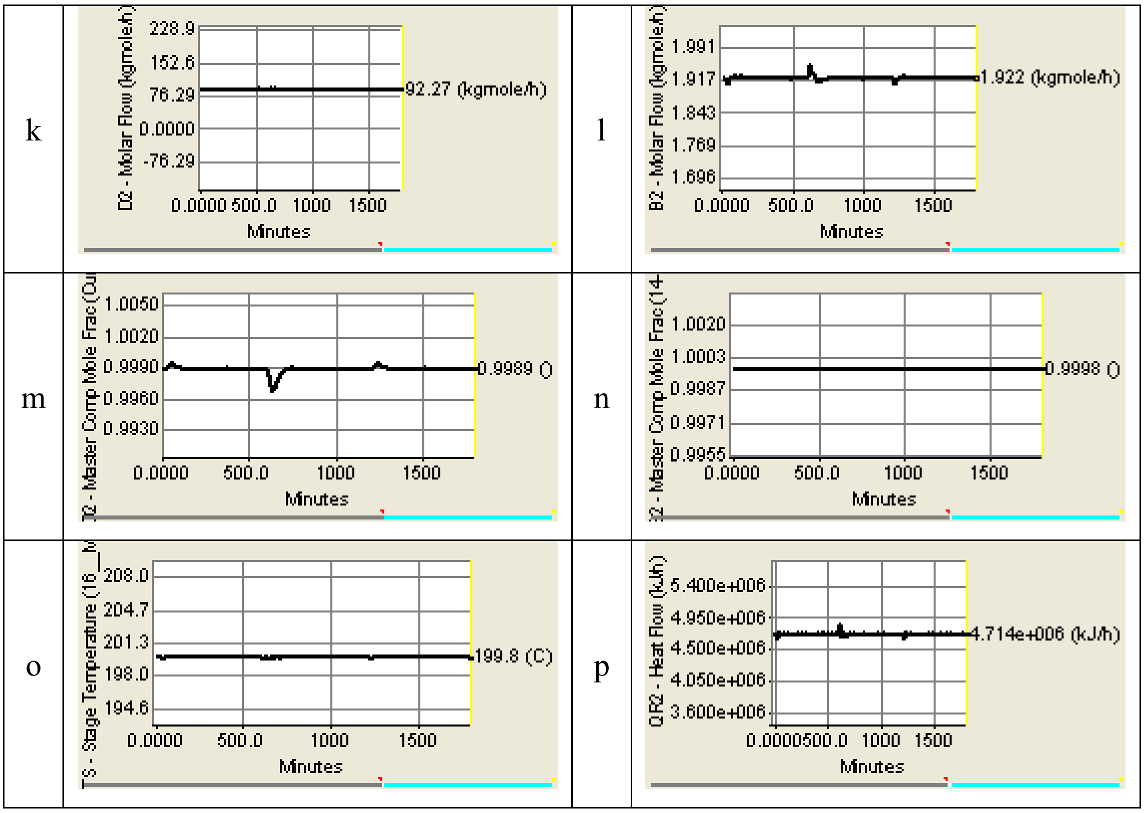


Figure5.19 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS2, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

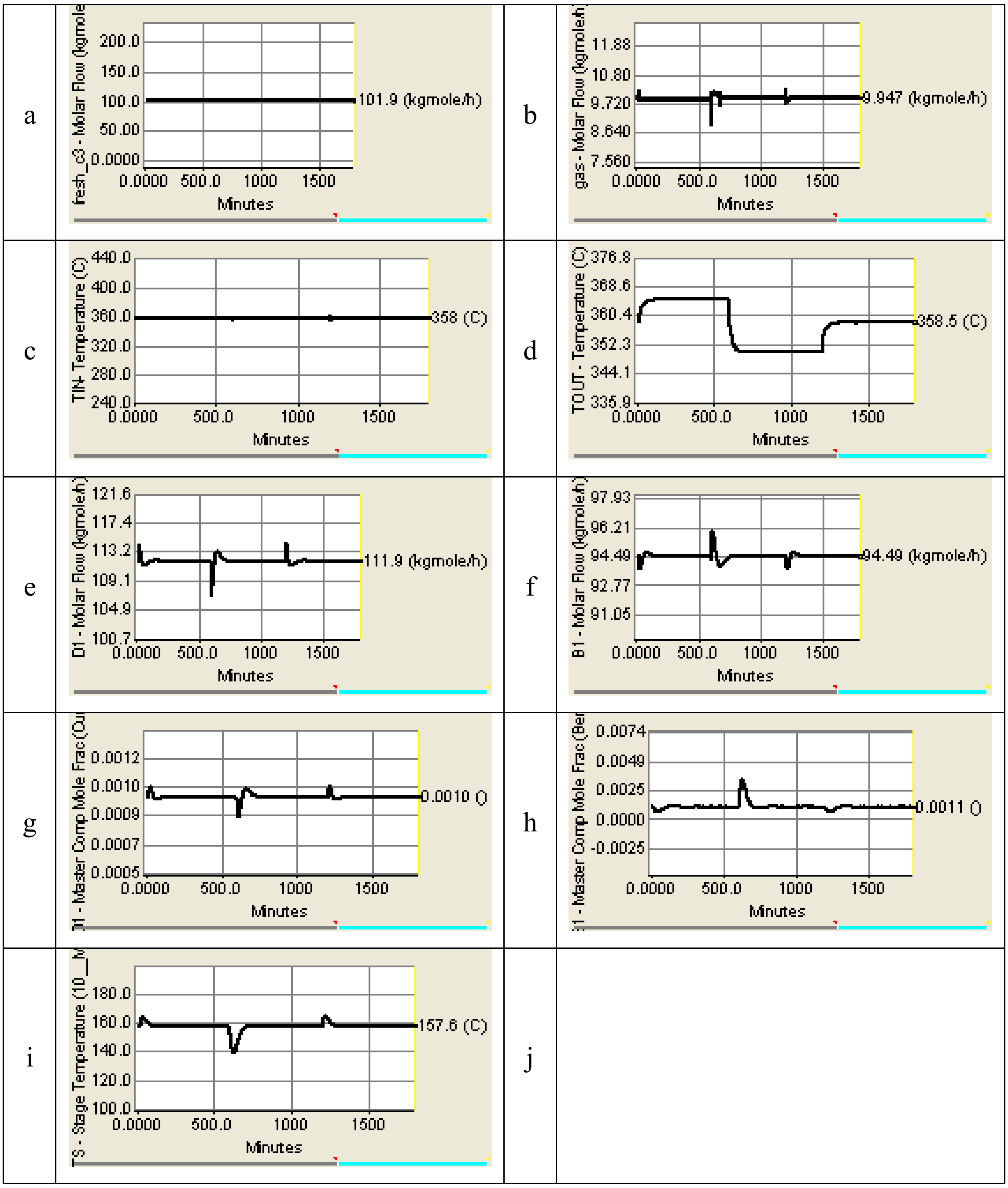


Figure5.20 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS3, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

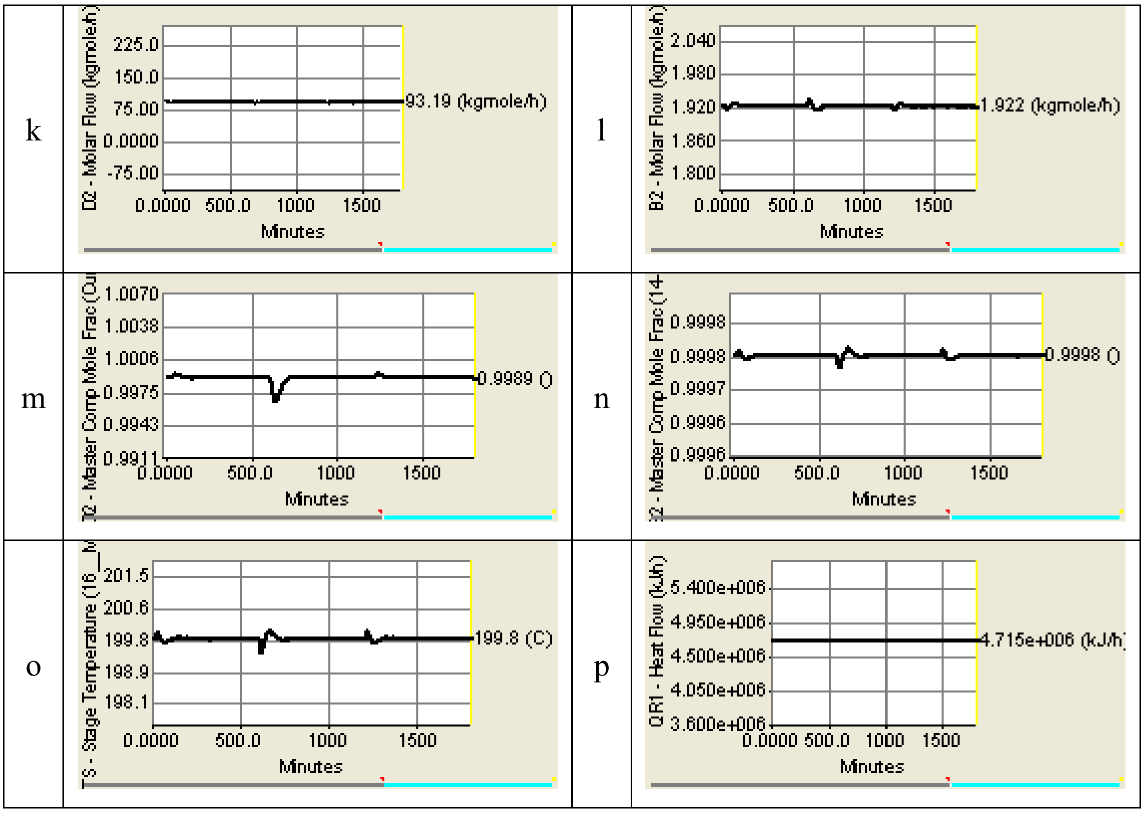


Figure5.20 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS3, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

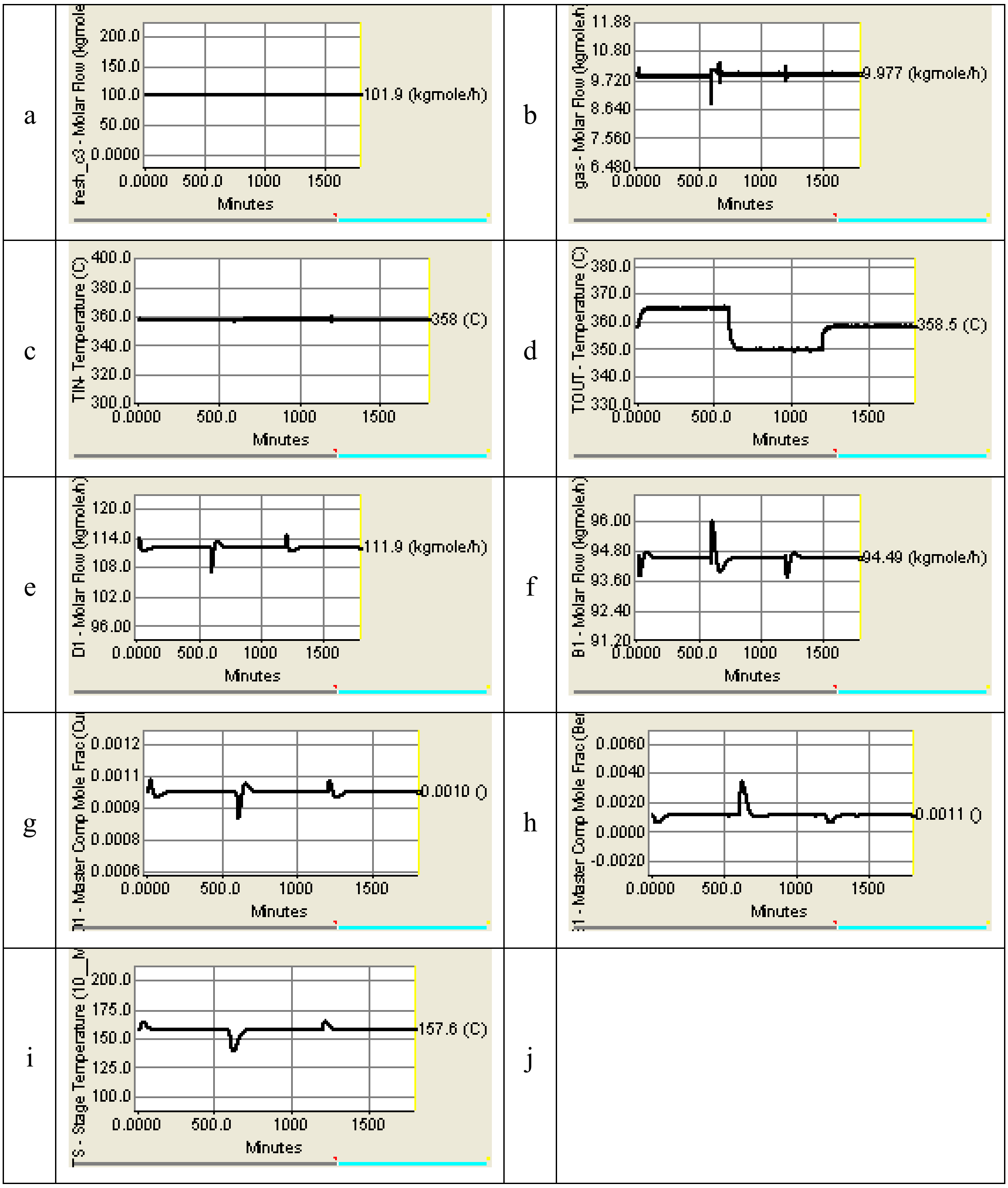


Figure5.21 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS4, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, g) recycle composition, (h) bottom composition of C1 column, (i) tray 10^{th} temperature of C1 column, (j) reboiler duty of C1 column

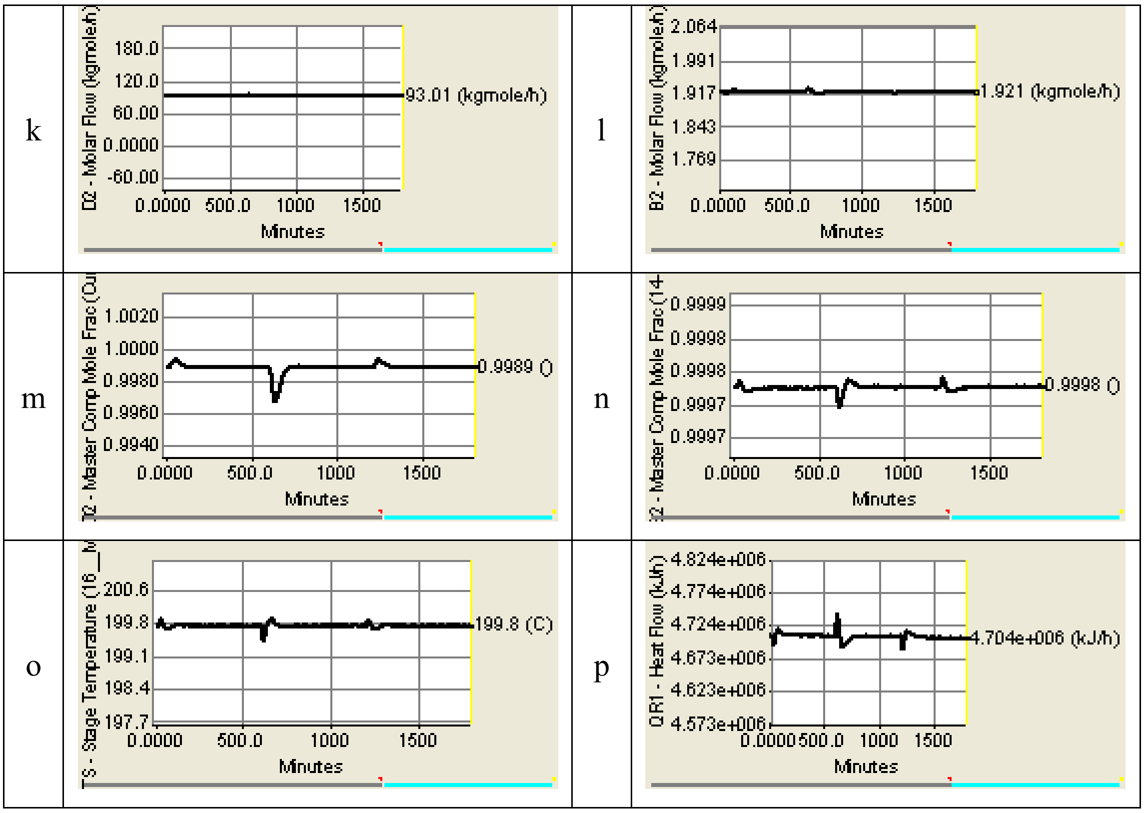


Figure5.21 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS4, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

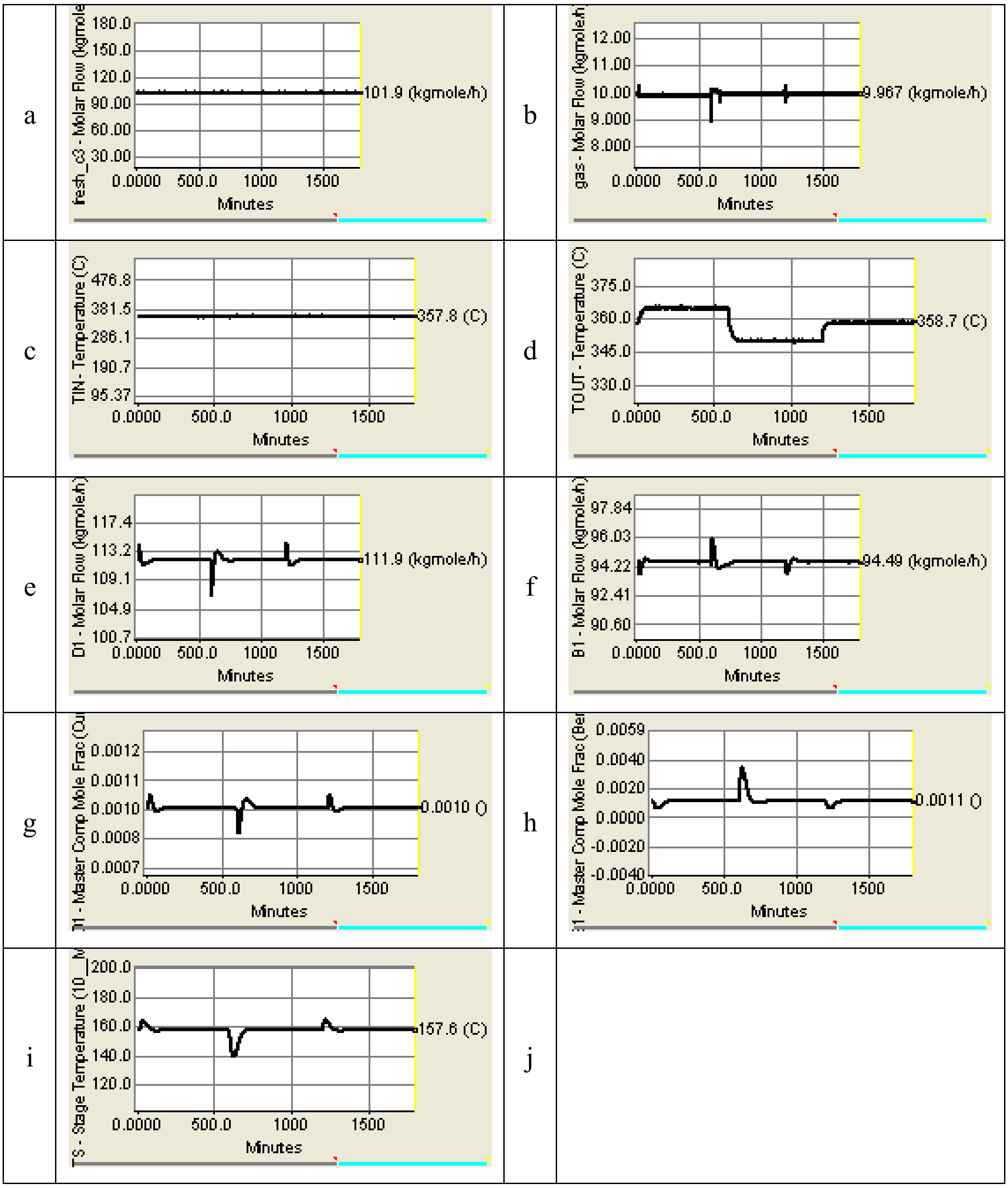


Figure5.22 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS5, where (a) propylene fresh feed flowrate, (b) gas flow rate, (c) reactor inlet temperature, (d) reactor outlet temperature, (e) recycle flow rate, (f) bottom flow rate of C1 column, g) recycle composition, (h) bottom composition of C1 column, (i) tray 10th temperature of C1 column, (j) reboiler duty of C1 column

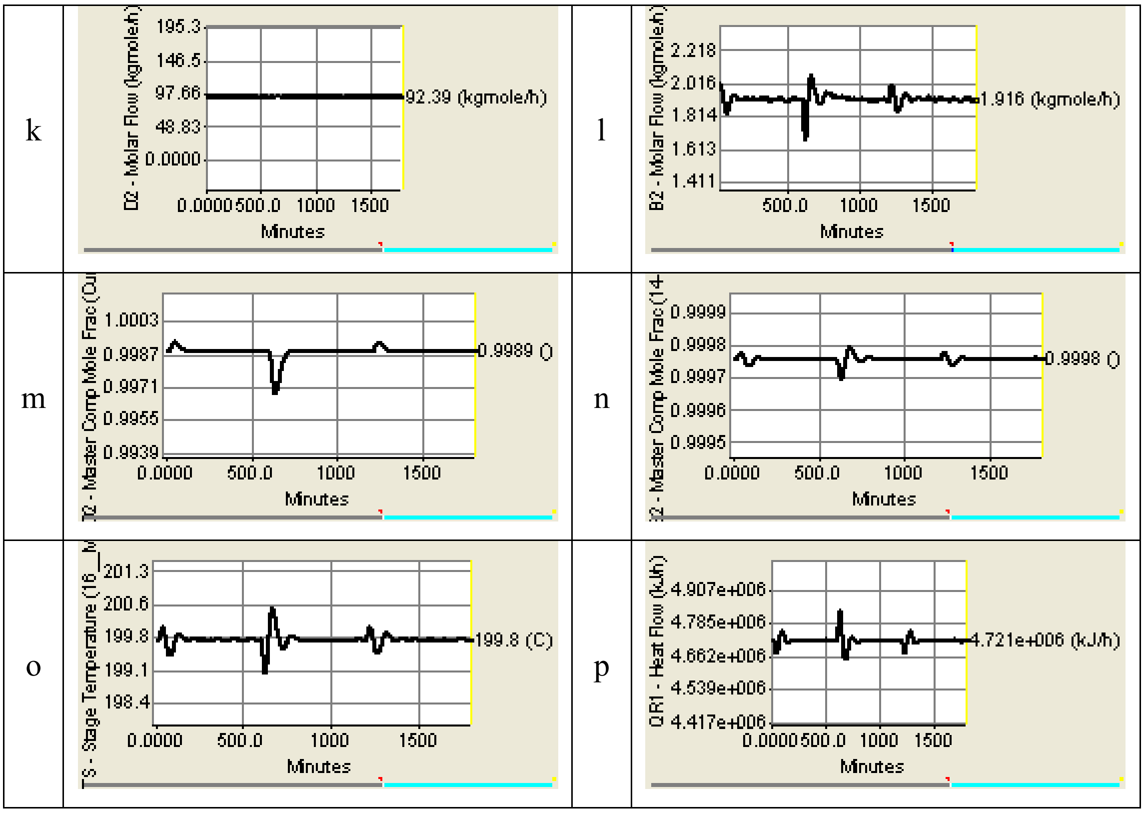


Figure5.22 Dynamic responses for cumene process when the outlet reactor temperature is changed, HEN-CS5, where (k) product flow rate, (l) by-product flow rate, (m) product composition, (n) by-product composition, (o) tray 16th temperature of C2 column, (p) reboiler duty of C2 column

5.4 Evaluation of the Dynamic Performance

The dynamic performance index is focused on time related characteristics of the controller's response to setpoint changes or deterministic disturbances. There exist several candidate performance measures such as settling time and integral absolute error (IAE). Integral absolute error is well known and widely used. For the formulation of a dynamic performance as written below:

$$\text{IAE refer to} \int \left| \varepsilon (t) \right| dt \tag{1}$$

Note that ε (t) refer to $y_{sp}(t) - y(t)$ is the deviation (error) of the response from the desired setpoint.

In this research, IAE method is used to evaluate the dynamic performance of all control structures. There are several types of control variables in this process (temperature, pressure, molar flow rate and level) so to compare it we must divide by span (the largest expected change in disturbance) of each variable. The IAE results consider in handle disturbances and maintain product quality. For energy use, the summation value of all energy use is used to evaluate the dynamic performance of all control structures.

For changing in material disturbances of propylene fresh feed flowrate for all control structure, the IAE results for handle disturbances and maintain product quality and the summation value of all energy use are shown in Table 5.7, Table 5.9, and Table 5.11, respectively.

For changing in thermal disturbances of outlet reactor temperature for all control structure, the IAE results for handle disturbances and maintain product quality and the summation value of all energy use are shown in Table 5.8, Table 5.10, and Table 5.12, respectively.

Table5.7 The IAE Result for handle disturbances to the change by step ±10% of propylene fresh feeds flow rate for temperature loops.

Controller	BC	HEN				
		CS1	CS2	CS3	CS4	CS5
TCR	0.00692	0.00786	0.00751	0.00742	0.00731	0.00774
TCC1	0.03857	0.03902	0.02695	0.03899	0.03914	0.03639
TCC2	0.00008	0.00018	0.00008	0.00031	0.00021	0.00032
TCT	0.00679	0.00493	0.00493	0.00417	0.00397	0.00477
SUM	0.05236	0.05199	0.03947	0.05089	0.05063	0.04922

Table5.8 The IAE Result for handle disturbances to the change by step increase from 358.5°C to 365°C and from 365 °C to 350 °C of outlet reactor temperature for temperature loops.

Controller	BC	HEN				
		CS1	CS2	CS3	CS4	CS5
TCR	0.00167	0.00127	0.00425	0.00469	0.00421	0.00422
TCC1	0.00162	0.02833	0.02394	0.03162	0.03205	0.0333
TCC2	0.00000	0.00003	0.00002	0.00014	0.00005	0.00021
TCT	0.00079	0.00141	0.0015	0.00152	0.00151	0.00151
SUM	0.00408	0.03104	0.02971	0.03797	0.03782	0.03924

Note TC refer to Temperature control, PC refer to Pressure control, R refer to Reactor, T refer to flash tank, C1 refer to Benzene recycle column, C2 refer to Cumene column, CC2t refer to Top composition of cumene column

Table5.9 The IAE Result for handle disturbances to the change by step ±10% of propylene fresh feeds flow rate for pressure loops.

Controller	BC	HEN				
		CS1	CS2	CS3	CS4	CS5
PCC1	0.01891	0.01693	0.01389	0.01693	0.01695	0.01618
PCC2	0.00938	0.00526	0.00696	0.01024	0.00366	0.01395
PCT	0.00846	0.00067	0.00067	0.00073	0.0007	0.00073
SUM	0.03675	0.02286	0.02152	0.0279	0.02131	0.03086

Table5.10 The IAE Result for handle disturbances to the change by step increase from 358.5°C to 365°C and from 365 °C to 350 °C of outlet reactor temperature for pressure loops.

Controller	BC	HEN				
		CS1	CS2	CS3	CS4	CS5
PCC1	0.00001	0.00024	0.0045	0.00584	0.0059	0.00849
PCC2	0	0.00419	0.00211	0.00405	0.00218	0.00211
PCT	0.00034	0.00018	0.00034	0.00035	0.00035	0.00035
SUM	0.00035	0.00461	0.00695	0.01024	0.00843	0.01095

Table5.11 The IAE Result of composition change by step increase ±10% of propylene fresh feeds flow rate

Composition	BC	HEN				
		CS1	CS2	CS3	CS4	CS5
xD2	0.00152	0.00153	0.00182	0.00115	0.00164	0.00157
xB2	0.0019	0.00188	0.00194	0.00022	0.00099	0.00204
SUM	0.00342	0.00341	0.00376	0.00137	0.00263	0.00361

Table5.12 The IAE Result of composition change by step increase from 358.5°C to 365°C and from 365 °C to 350 °C of outlet reactor temperature

Composition	BC	HEN				
		CS1	CS2	CS3	CS4	CS5
xD2	0.00117	0.0012	0.00122	0.00114	0.00112	0.00116
xB2	0.00003	0.00002	0.00002	0.00002	0.00002	0.00007
SUM	0.0012	0.00123	0.00124	0.00116	0.00114	0.0012

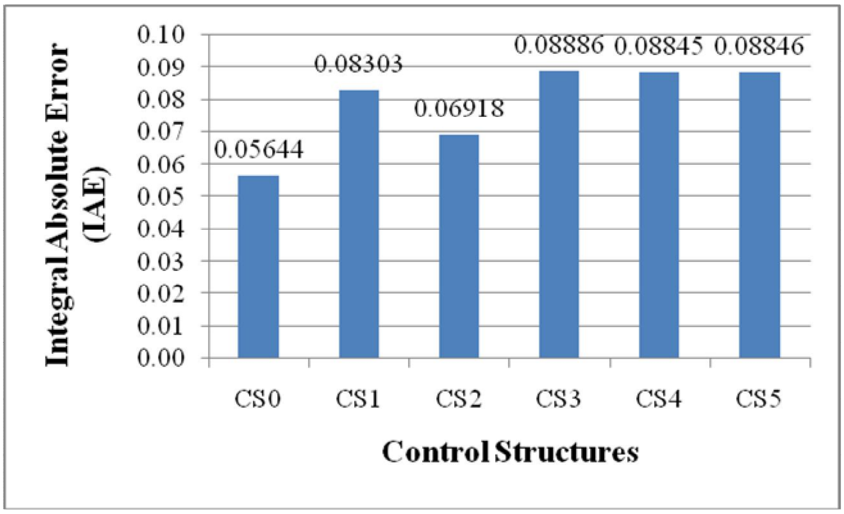


Figure5.23 The IAE summations for handle disturbances to the change in all disturbances testing for temperature loops.

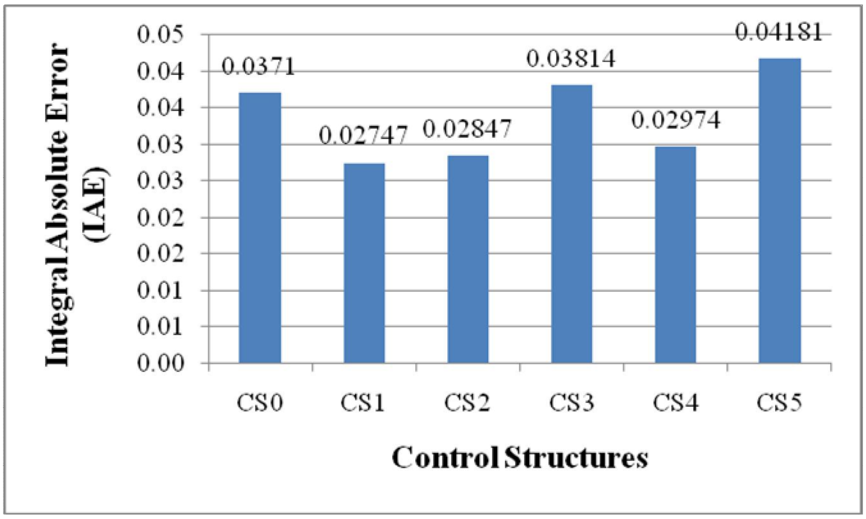


Figure5.23 The IAE summations for handle disturbances to the change in all disturbances testing for pressure loops.

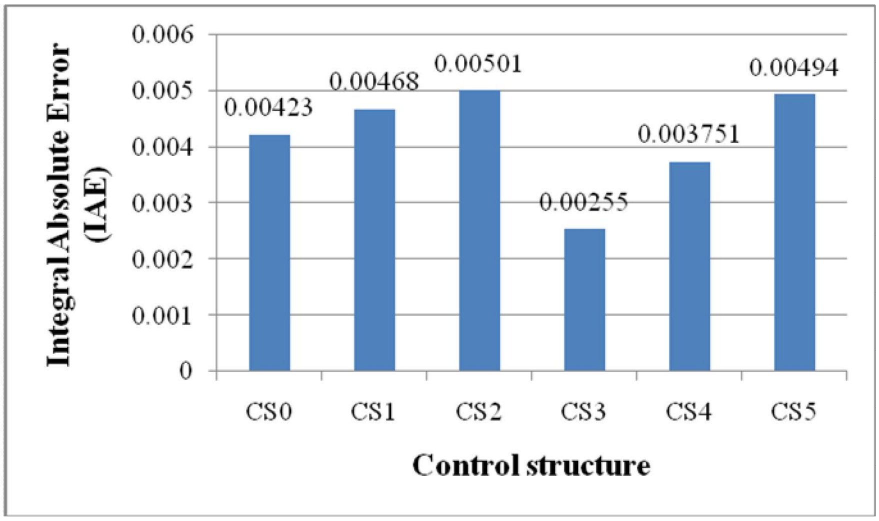


Figure5.23 The IAE Result of composition to the change in all disturbances testing

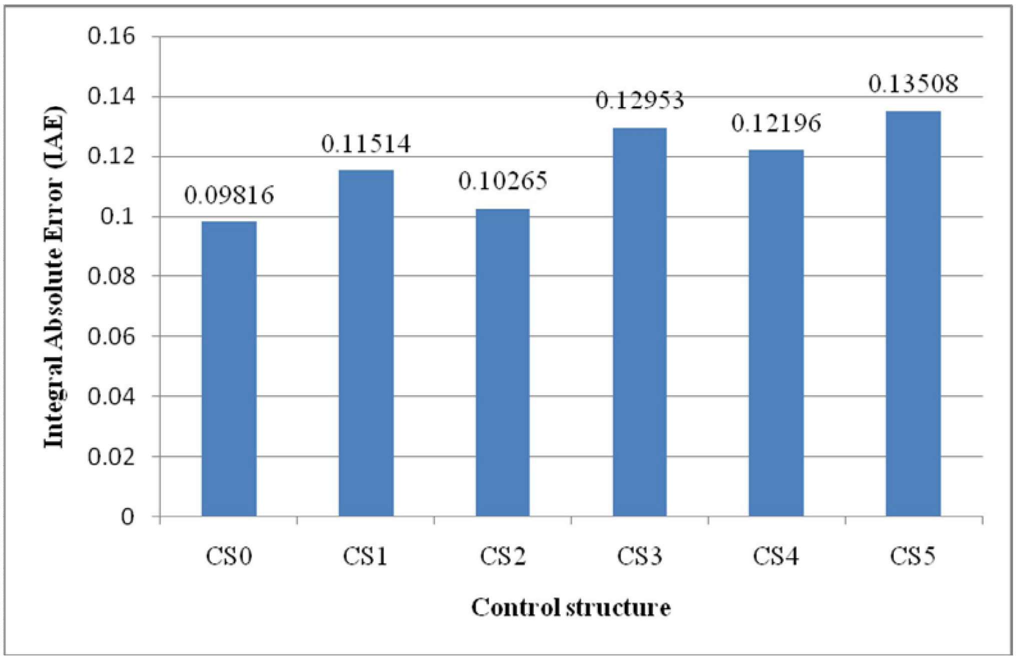


Figure5.23 The IAE summations for handle disturbances to the change in all disturbances testing

For all disturbances testing, Figure 5.23 shows the results of Integral Absolute Error (IAE) of all control structure for handle disturbances. CS0 control structure can handle disturbances the best of all control structures. Considering new design control structures, notice that the integral absolute error of control structure CS2 is close to the CS0 when compared with other control structures.

Table5.11 The summation value of all energy use to change by step increase ±10% of propylene fresh feeds flow rate

		Summation value (kW)				
Energy	CS0	CS1	CS2	CS3	CS4	CS5
Qc1	1.70E+06	1.67E+06	1.36E+06	1.67E+06	1.67E+06	9.56E+05
Qc2	1.08E+06					
Qhx1	6.77E+05	4.30E+06	4.47E+06	4.69E+06	4.58E+06	4.20E+06
Qhx2	2.31E+06	3.33E+05	3.00E+05	2.36E+05	2.14E+05	2.63E+05
Qr1	1.61E+06					
Qr2	5.38E+05	1.14E+06	8.18E+05	0.00E+00	3.93E+05	1.36E+06
Qrxn	1.99E+06	3.53E+06	3.52E+06	3.54E+06	3.53E+06	3.52E+06
Qvap	4.19E+06					
SUM	1.41E+07	1.10E+07	1.05E+07	1.01E+07	1.04E+07	1.03E+07

Note: Qc1 refer to condenser duty of benzene recycle column, Qc2 refer to condenser duty of cumene column, Qhx1 refer to HX1 heater duty, Qhx2 refer to HX2 cooler duty, Qr1 refer to reboiler duty of benzene recycle column, Qr2 refer to reboiler duty of cumene column, Qrxn refer to plug flow reactor duty, Qvap refer to vaporizer duty.

Table5.12 The summation value of all energy use to change by step increase from 358.5°C to 365°C and from 365 °C to 350 °C of outlet reactor temperature

Energy	CS0	Summation value (kW)				
	CS0	CS1	CS2	CS3	CS4	CS5
Qc1	5.01E+05	5.12E+04	3.25E+04	4.34E+04	4.38E+04	1.82E+04
Qc2	1.44E+04					
Qhx1	8.27E+05	7.18E+05	9.91E+05	1.18E+06	9.92E+05	9.13E+05
Qhx2	2.24E+05	1.15E+05	6.32E+04	7.25E+04	8.04E+04	9.27E+04
Qr1	3.73E+03					
Qr2	8.43E+03	1.36E+05	8.69E+04	0.00E+00	4.45E+04	1.09E+05
Qrxn	1.38E+06	1.35E+06	1.38E+06	1.39E+06	1.40E+06	1.39E+06
Qvap	5.16E+05					
SUM	3.47E+06	2.37E+06	2.55E+06	2.69E+06	2.56E+06	2.52E+06

Note: Qc1 refer to condenser duty of benzene recycle column, Qc2 refer to condenser duty of cumene column, Qhx1 refer to HX1 heater duty, Qhx2 refer to HX2 cooler duty, Qr1 refer to reboiler duty of benzene recycle column, Qr2 refer to reboiler duty of cumene column, Qrxn refer to plug flow reactor duty, Qvap refer to vaporizer duty.

For all disturbances testing, Figure 5.24 shows the summation value for all energy use of all control structure. Observed that all of the new design control structures are minimizes energy used for handing disturbance entering the process than CS0.

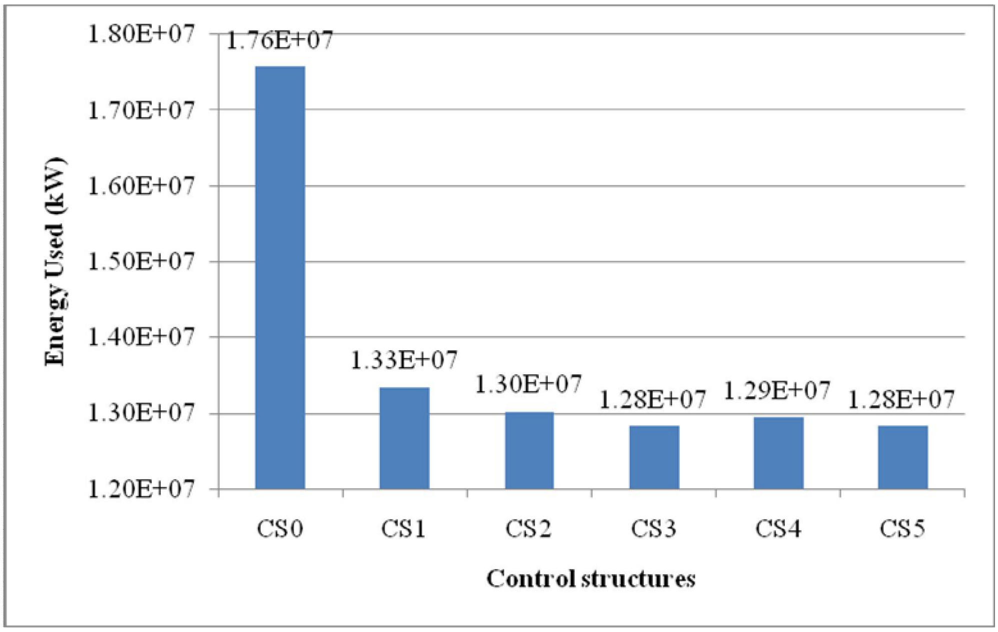


Figure5.24 The IAE summations value of energy used to change in all disturbance testing

5.5 Evaluation of cost

In this section is discusses the cost estimate for cumene process. The capital cost of equipment includes the cost of reactor, heat exchanger, pre-heater, cooler, reboiler, condenser, vessel and distillation column. The operating cost of energy includes the cost energy of reactor, pre-heater, cooler, reboiler and condenser. The costs associated with supplying a given utility are then obtained by calculating the operating costs to generate the utility. These are the costs that have been presented in Table 5.13. Total capital investment for base case structure and heat exchanger network structure are \$12,667,693.65 and \$12,624,611.06, respectively. The net total energy cost for base case structure is \$5,184,399.42 and for heat exchanger network structure is \$2,969,366.45. The total annual cost for base case is \$9,406,963.97 and for heat exchanger network structure is \$7,177,570.14

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Table 5.13 Utility costs (Analysis, Synthesis, and Design of Chemical Processes,2007

Utility	Description	Cost ($/GJ)
Steam from boilers	Process steam: latent heat only High pressure (41 barg, 254 oC)	9.83
Cooling tower water	Process cooling water: 30°C to 40°C or 45 °C	0.354
Electrical substation	Electric Distribution a. 110 V b. 220 V c. 440 V	16.8

Moreover, when disturbances entire to the process is changing in utility used, so utility costs are changed. Utility costs for each control structure to the change in all disturbances testing is shown in Figure 5.25

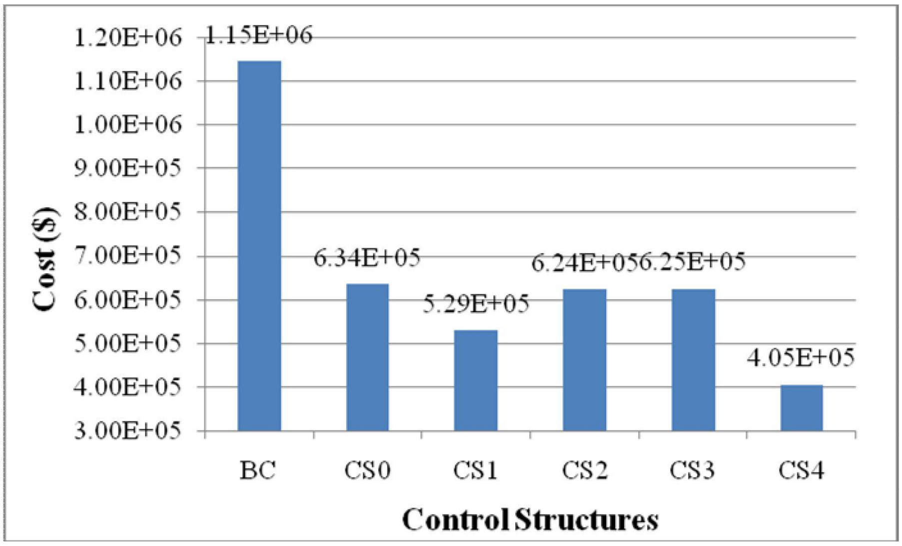

Figure 25. Utility costs for each control structure to the change in all disturbances testing

CHAPTER VI

CONCLUTIONS AND RECOMMENDATIONS

6.1 Conclusion

This research has discussed the control structure design for cumene process, using new design procedure of Wongsri (2009). This procedure based on heuristics and mathematical analysis. The preference of control variables is established. The purposed plantwide control structure design procedure for selection the best set of control structure is intuitive, simple and straightforward

The best control structure should handle disturbances entering the process, maintain product quality and minimize energy use. The major disturbances are directed or managed explicitly to achieve the minimal interaction between loops by using the material disturbances and the thermal disturbances

For the material disturbances, designed control structure CS2 can handle disturbances the best when compare with other control structures. The designed control structure CS3 can maintain product quality the best when compare with other control structures. Moreover, designed control structure CS3 is the most minimize energy use.

For the thermal disturbances testing, control structure of CS0 can handle disturbances the best when compare with other control structures. The designed control structure CS4 can maintain product quality the best when compare with other control structures and designed control structure CS1 is the most minimize energy use.

For all disturbances testing, control structure of CS0 can handle disturbances the best of all control structures and the design control structure CS2 can handle disturbances is close to the CS0. The designed control structure CS3 can maintain product quality the best when compare with other control structures. Moreover, all of control structures are the most minimize energy use when compare with CS0. When considering IAE and the most minimize energy use of all control structures it can be concluded that the new design control structure CS1 is the best.

New design procedure of Wongsri (2009) can be used to find the appropriate set of controlled variables to achieve form fixture point theorem. The best control configurations depend on the direction of controlled variable with manipulated variable. Therefore this research establishes that the Wongsri's procedure, which combines heuristics, analytical method and dynamic simulation, a useful design procedure that leads to a good-performance plantwide control system.

6.2 Recommendations

This research has been focus on new plantwide control design procedure of Wongsri (2009) applied to cumene process. There is added heat integration into the process even more. So performance of the process is slightly lower than the CS0. Deviations that occur are acceptable. If you want to improve performance of the process you need to change the controller type

Finally, study and design the control structure of the other process in plantwide control via new design procedure of Wongsri (2009).

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APPENDICES

APPENDIX A

PROCESS STREAM AND EQUIPMENT DATA

Table A.1 Data of Cumene process (CS0) for simulation

Name	fresh_c3	fresh_B	recycle	v1out	mix1out
Vapor Fraction	0.000	0.000	0.035	0.000	0.000
Temperature (c)	25.000	25.000	44.482	25.000	38.274
Pressure (BAR)	30.000	3.000	1.000	1.000	1.000
Molar Flow (kgmol/hr)	101.930	98.780	110.865	98.780	209.645
Mass Flow (kg/hr)	4299.55	7715.70	8457.37	7715.70	16173.07
Comp Mole Frac (Propane)	0.050	0.000	0.043	0.000	0.023
Comp Mole Frac (Proene)	0.950	0.000	0.011	0.000	0.006
Comp Mole Frac (Benzene)	0.000	1.000	0.945	1.000	0.971
Comp Mole Frac (Cumene)	0.000	0.000	0.001	0.000	0.001
Comp Mole Frac (14-ip-BZ)	0.000	0.000	0.000	0.000	0.000

Table A.1 (Continued) Data of Cumene process (CS0) for simulation

Name	totb	v2out	tot	htot	tubeout
Vapor Fraction	0.000	0.000	0.000	0.805	1.000
Temperature (c)	40.780	25.000	36.076	210.000	330.000
Pressure (BAR)	28.000	28.000	28.000	27.000	26.000
Molar Flow (kgmol/hr)	209.645	101.930	311.575	311.575	311.575
Mass Flow (kg/hr)	16173.07	4299.55	20472.63	20472.63	20472.63
Comp Mole Frac (Propane)	0.023	0.050	0.032	0.032	0.032
Comp Mole Frac (Proene)	0.006	0.950	0.315	0.315	0.315
Comp Mole Frac (Benzene)	0.971	0.000	0.653	0.653	0.653
Comp Mole Frac (Cumene)	0.001	0.000	0.000	0.000	0.000
Comp Mole Frac (14-ip-BZ)	0.000	0.000	0.000	0.000	0.000

Table A.1 (Continued) Data of Cumene process (CS0) for simulation

Name	p2	shellout	HX1out	p1	v4out
Vapor Fraction	1.000	0.571	1.000	1.000	1.000
Temperature (c)	358.500	271.290	358.000	358.500	210.255
Pressure (BAR)	24.000	23.000	25.000	24.000	2.000
Molar Flow (kgmol/hr)	216.278	216.278	311.575	216.277	216.278
Mass Flow (kg/hr)	20473.09	20473.09	20472.63	20472.95	20473.09
Comp Mole Frac (Propane)	0.046	0.046	0.032	0.046	0.046
Comp Mole Frac (Proene)	0.013	0.013	0.315	0.013	0.013
Comp Mole Frac (Benzene)	0.503	0.503	0.653	0.503	0.503
Comp Mole Frac (Cumene)	0.435	0.435	0.000	0.435	0.435
Comp Mole Frac (14-ip-BZ)	0.003	0.003	0.000	0.003	0.003

Table A.1 (Continued) Data of Cumene process (CS0) for simulation

Name	HX2out	gas	L1	gout	p3out
Vapor Fraction	0.046	1.000	0.000	1.000	0.000
Temperature (c)	80.000	80.000	80.000	80.000	80.085
Pressure (BAR)	1.750	1.750	1.750	1.000	2.750
Molar Flow (kgmol/hr)	216.278	9.872	206.406	9.872	206.406
Mass Flow (kg/hr)	20473.09	554.65	19918.44	554.65	19918.44
Comp Mole Frac (Propane)	0.046	0.520	0.023	0.520	0.023
Comp Mole Frac (Proene)	0.013	0.147	0.006	0.147	0.006
Comp Mole Frac (Benzene)	0.503	0.308	0.513	0.308	0.513
Comp Mole Frac (Cumene)	0.435	0.025	0.455	0.025	0.455
Comp Mole Frac (14-ip-BZ)	0.003	0.000	0.003	0.000	0.003

Table A.1 (Continued) Data of Cumene process (CS0) for simulation

Name	v6out	D1	B1	v7out	p4out
Vapor Fraction	0.000	0.000	0.000	0.035	0.000
Temperature (c)	80.073	49.787	178.504	44.512	178.582
Pressure (BAR)	1.750	1.750	1.900	1.000	2.900
Molar Flow (kgmol/hr)	206.406	111.925	94.481	111.925	94.481
Mass Flow (kg/hr)	19918.44	8538.33	11380.10	8538.33	11380.10
Comp Mole Frac (Propane)	0.023	0.043	0.000	0.043	0.000
Comp Mole Frac (Proene)	0.006	0.012	0.000	0.012	0.000
Comp Mole Frac (Benzene)	0.513	0.945	0.001	0.945	0.001
Comp Mole Frac (Cumene)	0.455	0.001	0.992	0.001	0.992
Comp Mole Frac (14-ip-BZ)	0.003	0.000	0.007	0.000	0.007

Table A.1 (Continued) Data of Cumene process (CS0) for simulation

Name	v8out	D2	B2	p5out	p6out
Vapor Fraction	0.197	0.000	0.000	0.000	0.000
Temperature (c)	151.997	151.642	167.434	151.721	167.512
Pressure (BAR)	1.000	1.000	1.000	2.000	2.000
Molar Flow (kgmol/hr)	94.481	92.858	1.623	92.858	1.623
Mass Flow (kg/hr)	11380.10	11157.02	223.08	11157.02	223.08
Comp Mole Frac (Propane)	0.000	0.000	0.000	0.000	0.000
Comp Mole Frac (Proene)	0.000	0.000	0.000	0.000	0.000
Comp Mole Frac (Benzene)	0.001	0.001	0.000	0.001	0.000
Comp Mole Frac (Cumene)	0.992	0.999	0.590	0.999	0.590
Comp Mole Frac (14-ip-BZ)	0.007	0.000	0.410	0.000	0.410

Table A.1 (Continued) Data of Cumene process (CS0) for simulation

Name	product	v10out
Vapor Fraction	0.001	0.001
Temperature (c)	151.642	167.443
Pressure (BAR)	1.000	1.000
Molar Flow (kgmol/hr)	92.858	1.623
Mass Flow (kg/hr)	11157.02	223.08
Comp Mole Frac (Propane)	0.000	0.000
Comp Mole Frac (Proene)	0.000	0.000
Comp Mole Frac (Benzene)	0.001	0.000
Comp Mole Frac (Cumene)	0.999	0.590
Comp Mole Frac (14-ip-BZ)	0.000	0.410

Table A.2 Process equipments data for CS0

Parameter	Plug Flow Reactor	Vaporizer	FEHE	Flash tank
Length(m)	6	-	-	-
Height	-	-	-	2.75
Diameter (m)	0.0763	-	-	1.376
Number of tubes	1500	-	320	-
catalyst void fraction	0.5	-	-	-
catalyst density(kg/m3)	2000	-	-	-
Heat duty (kJ/hr)	9.262E+06	1.236E+07	6.33E+06	-
UA (kJ/C-h)	-	-	7E+05	-
Hot out temperature (°C)	-	-	271.3	-

Table A.3 Column specifications for CS0

Parameter	Benzene recycle columu	Cumene column
Theoretical trays	15	20
Feed tray	6	12
Diameter (m)	1.524	1.372
Condenser volume (m3)	2	2.5
Reboiler volume (m3)	22.3	16.45
Reflux ratio	0.44	0.63
Reboiler duty (kJ/h)	7.583E+06	5.027E+06
Condenser duty (kJ/h)	5.06E+06	5.603E+06

Table A.4 Process equipments data for all of design control structures

Parameter	Plug Flow Reactor	V-100	V-101	Flash tank
Length(m)	6	-	-	-
Height	-	2.263	4.105	2.271
Diameter (m)	0.0763	1.509	2.677	1.514
Number of tubes	1500	-	-	-
catalyst void fraction	0.5	-	-	-
catalyst density(kg/m3)	2000	-	-	-
Heat duty (kJ/hr)	9.12E+06	-	-	-
UA (kJ/C-h)	-	-	-	-
Hot out temperature (°C)	-	-	-	-

Table A.4 (continued) Process equipments data for all of design control structures

Parameter	E1	E2	E3	E4
Length(m)	-	-	-	-
Height	-	-	-	-
Diameter (m)	-	-	-	-
Number of tubes	160	160	160	160
catalyst void fraction	-	-	-	-
catalyst density(kg/m3)	-	-	-	-
Heat duty (kJ/hr)	5.61E+06	3.12E+06	7.39E+06	3.79E+06
UA (kJ/C-h)	2.50E+05	9.50E+04	8.33E+04	5.50E+04
Hot out temperature (°C)	145.3	185.4	226	211

Table A.5 Column specifications for all of design control structures

Parameter	Benzene recycle columu	Cumene column
Theoretical trays	15	20
Feed tray	6	12
Diameter (m)	1.524	1.372
Condenser volume (m3)	1.818	-
Reboiler volume (m3)	-	16.44
Reflux ratio	0.44	0.63
Reboiler duty (kJ/h)	-	5.04E+06
Condenser duty (kJ/h)	5.512E+06	-

APPENDIX B

TUNING OF CONTROL STRUCTURES

B.1 Tuning Controllers

Notice throughout this work uses several types of controllers such as P, PI, and PID controllers. They depend on the control loop. In theory, control performance can be improved by the use of derivative action but in practice the use of derivative has some significant drawbacks:

1. Three tuning constants must be specified.
2. Signal noise is amplified.
3. Several types of PID control algorithms are used, so important to careful that the right algorithm is used with its matching tuning method.
4. The simulation is an approximation of the real plant. If high performance controllers are required to get good dynamics from the simulation, the real plant may not work well.

B.2 Tuning Flow, Level, Pressure and Temperature Loops

Flow Controllers

The dynamics of flow measurement are fast. The time constants for moving control valves are small. Therefore, the controller can be turned with a small integral or reset time constant. A value τ_I=0.3 minutes work in most controllers. The value of controller gain should be kept modest because flow measurement signal are sometime noisy due to the turbulent flow through the orifice plate. A value of controller gain of K_c=0.5 is often used. Derivative action should not be used.

Level Controllers

Most level controllers should use proportional-only action with a gain of 1 to 2. This provides the maximum amount of flow smoothing. Proportional control means there will be steady state offset (the level will not be returned to its set point value).

However, maintaining a liquid level at a certain value is often not necessary when the liquid capacity is simply being used as surge volume. So the recommended tuning of a level controller is K_c =2.

Pressure Controllers

Most pressure controllers can be fairly easily tuned. The process time constant is estimated by dividing the gas volume of the system by the volumetric flow rate of gas flowing through the system. Setting the integral time equal to about 2 to 4 times the process time constant and using a reasonable controller gain usually gives satisfactory pressure control. Typical pressure controller tuning constants for columns and tanks are K_c = 2 and τ_I =10 minutes.

Temperature Controllers

Temperature dynamic responses are generally slow, so PID control is used. Typically, the controller gain, K_c, should be set between 2 to 10, the integral time.

B.3 Relay-Feedback Testing

The relay-feedback test is a tool that server a quick and simple method for identifying the dynamic parameters that are important for to design a feedback controller. The results of the test are the ultimate gain and the ultimate frequency. This information is usually sufficient to permit us to calculate some reasonable controller tuning constants.

The method consists of merely inserting on-off relay in the feedback loop. The only parameter that must be specified is the height of the relay, h. This height is typically 5 to 10 percent of the controller output scale. The loop starts to oscillate around the set point with the controller output switching every time the process variable (PV) signal crosses the set point. Figure A.1 shows the PV and OP signals from a typical relay-feedback test. The maximum amplitude (a) of the PV signal is used to calculate the ultimate gain, K_u from the equation.

$$Ku = \frac{4h}{a\pi} \quad (1)$$

The period of the output PV curve is the ultimate period, P_u from these two parameters controller tuning constants can be calculated for PI and PID controllers, using a variety of tuning methods proposed in the literature that require only the ultimate gain and the ultimate frequency, e.g. Ziegler-Nichols, Tyreus-Luyben.

The test has many positive features that have led to its widespread use in real plants as well in simulation studies:

1. Only one parameter has to be specified (relay height).

2. The time it takes to run the test short, particularly compared to the extended periods required for methods like PRBS.

3. The test is closed loop, so the process is not driven away from the setpoint.

4. The information obtained is very accurate in the frequency range that is important for the design of a feedback controller.

5. The impact of load changes that occur during the test can be detected by a change to asymmetric pulses in the manipulated variable.

These entire features make relay-feedback testing a useful identification tool.

Knowing the ultimate gain, Ku and the ultimate period, Pu permits us to calculate controller settings. There are several methods that require only these two parameters. The Zieglegler-Nichols tuning equations for a PI controller are:

$$K_c = K_u /2.2 \tag{2}$$

$$\tau_I = P_u/1.2 \tag{3}$$

These tuning constants are frequently too aggressive for many chemical engineering applications. The Tyreus-Luyben tuning method provides more conservative settings with increased robustness. The TL equations for a PI controller are:

$$K_c = K_u /3.2 \tag{4}$$

$$\tau_I = 2.2P_u \tag{5}$$

B.4 Inclusion of Lags

Any real physical system has many lags. Measurement and actuator lags always exist. In simulations, however, these lags are not part of the unit models. Much more aggressive tuning is often possible on the simulation than is possible in the real plant. Thus the predictions of dynamic performance can be overly optimistic. This is poor engineering. A conservative design is needed. Realistic dynamic simulations require that we explicitly include lags and/or dead times in all the important loops. Usually this means controllers that affect Product quality or process constraint. Table A.1 summarizes some recommended lags to include in several different types of control loops.

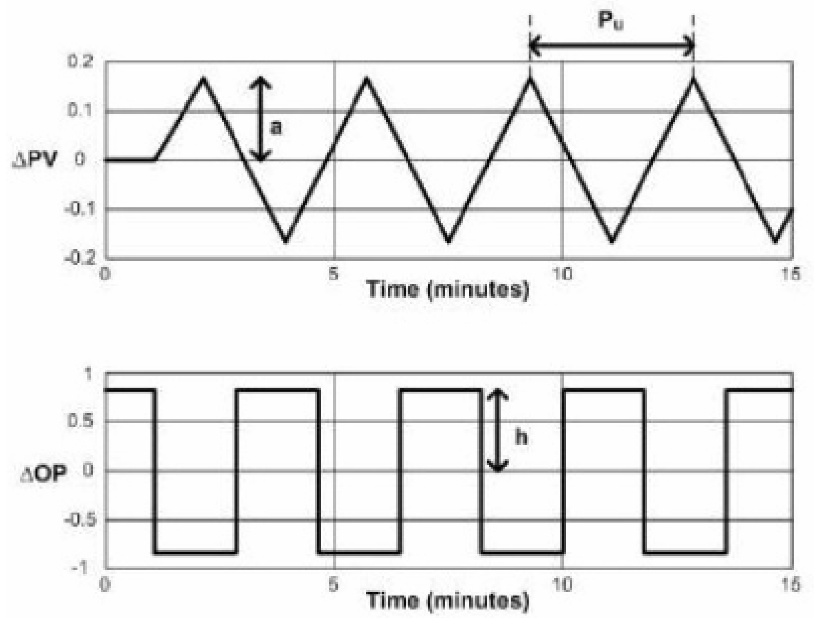

Figure B.1 Input and Output from Relay-Feedback Test

Table B.1 typical measurement lags

		Number	Time constant (minutes)	Type
Temperature	Liquid	2	0.5	First-order lags
	Gas	3	1	First-order lags
Composition	Chromatograph	1	3 to 10	Deadtime

Any real physical system has many lags. Measurement and actuator lags always exist. 1n simulations, however, these lags are not part of the unit models. Much more aggressive tuning is often possible on the simulation than is possible in the real plant. Thus the predictions of dynamic performance can be overly optimistic. This is poor engineering. A conservative design is needed.

Table B.2 Tuning parameter for the design control structure CS1

Controller	Controlled variable	Manipulated variable	Action controller	Tuning parameter			PV
				K_c	τ_I	τ_D	Rank
FIC-1	propylene fresh feed flowrate	V2	Reverse	0.199	0.011	-	0-204 kmol/h
FIC-2	total benzene flowrate	V1	Reverse	1.060	0.055	-	0-420 kmol/h
FIC-3	reflux flowrate of benzene recycle column	reflux	Reverse	0.173	0.018	-	0-92 kmol/h
FIC-4	reflux flowrate of cumene column	reflux	Reverse	0.204	0.008	-	0-119 kmol/h
LIC-1	flash tank level	V6	Direct	2.000	-	-	0-100 %
LIC-2	condenser level of benzene recycle column	V7	Direct	2.000	-	-	0-100 %
LIC-3	V-101 level	V8	Direct	2.000	-	-	0-100 %
LIC-4	V-100 level	V9	Direct	2.000	-	-	0-100 %
LIC-5	reboiler level of cumene column	V10	Direct	2.000	-	-	0-100 %
PIC-1	pressure stream	V4	Reverse	0.781	0.013	-	0-4 bar
PIC-2	flash tank pressure	V5	Direct	6.620	0.015	-	0-3.5 bar
PIC-3	condenser pressure of benzene recycle column	Qc1	Direct	5.310	0.180	-	0-3.5 bar
PIC-4	inlet V-100 pressure	Q	Direct	1.600	2.480	-	0-2 bar
TIC-1	total stream temperature	V3	Direct	2.000	10.000	1.000	111-251°C
TIC-2	inlet reactor temperature	Qhx1	Reverse	0.879	1.500	0.333	300-400°C
TIC-3	outlet reactor temperature	Qr	Direct	0.184	9.250	2.060	300-400°C
TIC-4	inlet flash tank temperature	Qhx2	Direct	1.520	1.760	0.391	30-130°C
TIC-5	tray temperature of benzene recycle column	bypass	Direct	0.685	28.200	6.270	91-191°C
TIC-6	tray temperature of cumene column	Qr2	Reverse	8.740	4.340	0.966	100-300°C

Table B.3 Tuning parameter for the design control structure CS2

Controller	Controlled variable	Manipulated variable	Action controller	Tuning parameter			PV
				Kc	τI	τD	Rank
FIC-1	propylene fresh feed flowrate	V2	Reverse	0.199	0.011	-	0-204 kmol/h
FIC-2	total benzene flowrate	V1	Reverse	1.060	0.055	-	0-420 kmol/h
FIC-3	reflux flowrate of benzene recycle column	reflux	Reverse	0.173	0.018	-	0-92 kmol/h
FIC-4	reflux flowrate of cumene column	reflux	Reverse	0.204	0.008	-	0-119 kmol/h
LIC-1	flash tank level	V6	Direct	2.000	-	-	0-100 %
LIC-2	condenser level of benzene recycle column	V7	Direct	2.000	-	-	0-100 %
LIC-3	V-101 level	V8	Direct	2.000	-	-	0-100 %
LIC-4	V-100 level	V9	Direct	2.000	-	-	0-100 %
LIC-5	reboiler level of cumene column	V10	Direct	2.000	-	-	0-100 %
PIC-1	pressure stream	V4	Reverse	0.781	0.013	-	0-4 bar
PIC-2	flash tank pressure	V5	Direct	6.620	0.015	-	0-3.5 bar
PIC-3	condenser pressure of benzene recycle column	Qc1	Direct	5.310	0.180	-	0-3.5 bar
PIC-4	inlet V-100 pressure	Q	Direct	1.600	2.480	-	0-2 bar
TIC-1	total stream temperature	V3	Direct	2.000	10.000	1.000	111-251°C
TIC-2	inlet reactor temperature	Qhx1	Reverse	0.879	1.500	0.333	300-400°C
TIC-3	outlet reactor temperature	Qr	Direct	0.184	9.250	2.060	300-400°C
TIC-4	inlet flash tank temperature	Qhx2	Direct	1.520	1.760	0.391	30-130°C
TIC-5	tray temperature of benzene recycle column	bypass	Direct	0.685	28.200	6.270	91-191°C
TIC-6	tray temperature of cumene column	Qr2	Reverse	8.740	4.340	0.966	100-300°C

Table B.4 Tuning parameter for the design control structure CS3

Controller	Controlled variable	Manipulated variable	Action controller	Tuning parameter			PV
				Kc	τI	τD	Rank
FIC-1	propylene fresh feed flowrate	V2	Reverse	0.199	0.011	-	0-204 kmol/h
FIC-2	total benzene flowrate	V1	Reverse	1.060	0.055	-	0-420 kmol/h
FIC-3	reflux flowrate of benzene recycle column	reflux	Reverse	0.173	0.018	-	0-92 kmol/h
FIC-4	reflux flowrate of cumene column	reflux	Reverse	0.204	0.008	-	0-119 kmol/h
LIC-1	flash tank level	V6	Direct	2.000	-	-	0-100 %
LIC-2	condenser level of benzene recycle column	V7	Direct	2.000	-	-	0-100 %
LIC-3	V-101 level	V8	Direct	2.000	-	-	0-100 %
LIC-4	V-100 level	V9	Direct	2.000	-	-	0-100 %
LIC-5	reboiler level of cumene column	V10	Direct	2.000	-	-	0-100 %
PIC-1	pressure stream	V4	Reverse	0.781	0.013	-	0-4 bar
PIC-2	flash tank pressure	V5	Direct	6.620	0.015	-	0-3.5 bar
PIC-3	condenser pressure of benzene recycle column	Qc1	Direct	5.310	0.180	-	0-3.5 bar
PIC-4	inlet V-100 pressure	Q	Direct	1.600	2.480	-	0-2 bar
TIC-1	total stream temperature	V3	Direct	2.000	10.000	1.000	111-251°C
TIC-2	inlet reactor temperature	Qhx1	Reverse	0.879	1.500	0.333	300-400°C
TIC-3	outlet reactor temperature	Qr	Direct	0.184	9.250	2.060	300-400°C
TIC-4	inlet flash tank temperature	Qhx2	Direct	1.520	1.760	0.391	30-130°C
TIC-5	tray temperature of benzene recycle column	bypass	Direct	0.685	28.200	6.270	91-191°C
TIC-6	tray temperature of cumene column	reflux	Direct	8.740	35.000	0.966	100-300°C

Table B.5 Tuning parameter for the design control structure CS4

Controller	Controlled variable	Manipulated variable	Action controller	Tuning parameter			PV
				Kc	τI	τD	Rank
FIC-1	propylene fresh feed flowrate	V2	Reverse	0.199	0.011	-	0-204 kmol/h
FIC-2	total benzene flowrate	V1	Reverse	1.060	0.055	-	0-420 kmol/h
FIC-3	reflux flowrate of benzene recycle column	reflux	Reverse	0.173	0.018	-	0-92 kmol/h
FIC-4	reflux flowrate of cumene column	reflux	Reverse	0.204	0.008	-	0-119 kmol/h
LIC-1	flash tank level	V6	Direct	2.000	-	-	0-100 %
LIC-2	condenser level of benzene recycle column	V7	Direct	2.000	-	-	0-100 %
LIC-3	V-101 level	V8	Direct	2.000	-	-	0-100 %
LIC-4	V-100 level	V9	Direct	2.000	-	-	0-100 %
LIC-5	reboiler level of cumene column	Qr2	Direct	2.000	-	-	0-100 %
PIC-1	pressure stream	V4	Reverse	0.781	0.013	-	0-4 bar
PIC-2	flash tank pressure	V5	Direct	6.620	0.015	-	0-3.5 bar
PIC-3	condenser pressure of benzene recycle column	Qc1	Direct	5.310	0.180	-	0-3.5 bar
PIC-4	inlet V-100 pressure	Q	Direct	1.600	2.480	-	0-2 bar
TIC-1	total stream temperature	V3	Direct	2.000	10.000	1.000	111-251°C
TIC-2	inlet reactor temperature	Qhx1	Reverse	0.879	1.500	0.333	300-400°C
TIC-3	outlet reactor temperature	Qr	Direct	0.184	9.250	2.060	300-400°C
TIC-4	inlet flash tank temperature	Qhx2	Direct	1.520	1.760	0.391	30-130°C
TIC-5	tray temperature of benzene recycle column	bypass	Direct	0.685	28.200	6.270	91-191°C
TIC-6	tray temperature of cumene column	reflux	Direct	8.740	35.000	0.966	100-300°C

Table B.6 Tuning parameter for the design control structure CS5

Controller	Controlled variable	Manipulated variable	Action controller	Tuning parameter			PV
				Kc	τI	τD	Rank
FIC-1	propylene fresh feed flowrate	V2	Reverse	0.199	0.011	-	0-204 kmol/h
FIC-2	total benzene flowrate	V1	Reverse	1.060	0.055	-	0-420 kmol/h
FIC-3	reflux flowrate of benzene recycle column	reflux	Reverse	0.173	0.018	-	0-92 kmol/h
FIC-4	reflux flowrate of cumene column	reflux	Reverse	0.204	0.008	-	0-119 kmol/h
LIC-1	flash tank level	V6	Direct	2.000	-	-	0-100 %
LIC-2	condenser level of benzene recycle column	V7	Direct	2.000	-	-	0-100 %
LIC-3	V-101 level	V8	Direct	2.000	-	-	0-100 %
LIC-4	V-100 level	V9	Direct	2.000	-	-	0-100 %
LIC-5	reboiler level of cumene column	Qr2	Direct	2.000	-	-	0-100 %
PIC-1	pressure stream	V4	Reverse	0.781	0.013	-	0-4 bar
PIC-2	flash tank pressure	V5	Direct	6.620	0.015	-	0-3.5 bar
PIC-3	condenser pressure of benzene recycle column	Qc1	Direct	5.310	0.180	-	0-3.5 bar
PIC-4	inlet V-100 pressure	Q	Direct	1.600	2.480	-	0-2 bar
TIC-1	total stream temperature	V3	Direct	2.000	10.000	1.000	111-251°C
TIC-2	inlet reactor temperature	Qhx1	Reverse	0.879	1.500	0.333	300-400°C
TIC-3	outlet reactor temperature	Qr	Direct	0.184	9.250	2.060	300-400°C
TIC-4	inlet flash tank temperature	Qhx2	Direct	1.520	1.760	0.391	30-130°C
TIC-5	tray temperature of benzene recycle column	bypass	Direct	0.685	28.200	6.270	91-191°C
TIC-6	tray temperature of cumene column	V10	Direct	13.100	52.000	11.600	100-300°C

APPENDIX C

FICTURE POINT THEOREM DATA

Table C.1 List of Manipulate Variables of the process

Manipulated variable	Description
V2	propylene fresh feed valve
V1	benzene fresh feed valve
reflux	benzene recycle column reflux valve
reflux	cumene column reflux valve
V6	flash tank bottom valve
V7	benzene recycle column distillatate valve
V8	benzene recycle column bottom valve
V9	cumene column distillatate valve
V10	cumene column bottom valve
V4	stream valve
V5	flash tank top valve
Qc1	condenser duty of benzene recycle column
Q	compressor power
V3	total stream by pass valve
Qhx1	QHX1 duty
Qr	reactor duty
Qhx2	QHX2 duty
by pass valve	mixed out stream valve
Qr2	reboiler duty of cumene column

Table C.2 IAE results of Flow rate for the process

STREAM	v1	v2	v4	v5	v6
B1 - Molar Flow	4.3509	0.3097	0.0037	0.8208	1.4026
B2 - Molar Flow	0.0450	0.0021	0.0001	0.0130	0.0292
D1 - Molar Flow	6.0279	0.1637	0.0045	0.6350	0.5992
D2 - Molar Flow	1.7252	0.1076	0.0020	0.3734	2.6015
fresh_B - Molar Flow	6.0650	0.5050	0.0055	1.5970	1.6668
fresh_c3 - Molar Flow	0.6509	1.0261	0.0009	0.2052	0.2193
gas - Molar Flow	2.4595	0.2420	0.0012	0.6198	0.0896
gout - Molar Flow	2.4595	0.2420	0.0012	0.6198	0.0896
htot - Molar Flow	0.9123	0.5110	0.0032	1.3785	1.2862
HX1out - Molar Flow	0.9125	0.5109	0.0031	1.3784	1.2861
HX2out - Molar Flow	0.5930	0.4682	0.0232	1.1840	1.0792
L1 - Molar Flow	17.8790	2.3505	0.0228	14.0425	16.3851
mix1out - Molar Flow	0.5686	0.5144	0.0028	1.1743	1.0678
p1 - Molar Flow	0.5627	0.4674	0.0143	1.1823	1.0769
p2 - Molar Flow	0.5627	0.4674	0.0143	1.1823	1.0769
p3out - Molar Flow	17.8790	2.3505	0.0228	14.0425	16.3851
p4out - Molar Flow	4.3509	0.3097	0.0037	0.8208	1.4026
p5out - Molar Flow	1.7252	0.1076	0.0020	0.3734	2.6015
p6out - Molar Flow	0.0450	0.0021	0.0001	0.0130	0.0292
product - Molar Flow	1.7252	0.1076	0.0020	0.3734	2.6015
pump1out - Molar Flow	0.5686	0.5144	0.0028	1.1743	1.0678
pump2out - Molar Flow	0.6509	1.0261	0.0009	0.2052	0.2193
recycle - Molar Flow	6.0279	0.1637	0.0045	0.6350	0.5992
shellout - Molar Flow	0.5623	0.4671	0.0169	1.1814	1.0767
tot - Molar Flow	0.9130	0.5110	0.0031	1.3796	1.2864
TotB - Molar Flow	0.5686	0.5144	0.0028	1.1743	1.0678
tubeout - Molar Flow	0.9124	0.5109	0.0032	1.3784	1.2861
v10out - Molar Flow	0.0450	0.0021	0.0001	0.0130	0.0292
v1out - Molar Flow	6.0650	0.5050	0.0055	1.5970	1.6668
v2out - Molar Flow	0.6509	1.0261	0.0009	0.2052	0.2193
v4out - Molar Flow	0.5623	0.4671	0.0169	1.1814	1.0767
v6out - Molar Flow	17.8790	2.3505	0.0228	14.0425	16.3851
v7out - Molar Flow	6.0279	0.1637	0.0045	0.6350	0.5992
v8out - Molar Flow	4.3509	0.3097	0.0037	0.8208	1.4026

Table C.2 (Continued) IAE results of Flow rate for the process

STREAM	v7	v8	v9	v10	qv
B1 - Molar Flow	1.6641	11.4751	1.5028	0.0018	15.6298
B2 - Molar Flow	0.0089	0.0470	0.0532	0.1331	0.1415
D1 - Molar Flow	2.6186	0.0455	0.0048	0.0010	13.4509
D2 - Molar Flow	1.3730	1.0383	4.4529	0.0005	5.7599
fresh_B - Molar Flow	1.6717	0.0586	0.0057	0.0007	23.6718
fresh_c3 - Molar Flow	0.1130	0.0056	0.0001	0.0000	7.2608
gas - Molar Flow	0.0440	0.0004	0.0001	0.0000	4.4851
gout - Molar Flow	0.0440	0.0004	0.0001	0.0000	4.4851
htot - Molar Flow	1.0503	0.0183	0.0016	0.0010	18.5883
HX1out - Molar Flow	1.0502	0.0182	0.0015	0.0009	18.5884
HX2out - Molar Flow	1.0473	0.0137	0.0013	0.0001	11.8645
L1 - Molar Flow	7.9976	0.1825	0.0173	0.0010	26.2106
mix1out - Molar Flow	1.0556	0.0133	0.0011	0.0007	11.3305
p1 - Molar Flow	1.0475	0.0137	0.0013	0.0000	11.8757
p2 - Molar Flow	1.0475	0.0137	0.0013	0.0000	11.8757
p3out - Molar Flow	7.9976	0.1825	0.0173	0.0010	26.2106
p4out - Molar Flow	1.6641	11.4751	1.5028	0.0018	15.6298
p5out - Molar Flow	1.3730	1.0383	4.4529	0.0005	5.7599
p6out - Molar Flow	0.0089	0.0470	0.0532	0.1331	0.1415
product - Molar Flow	1.3730	1.0383	4.4529	0.0005	5.7599
pump1out - Molar Flow	1.0556	0.0133	0.0011	0.0007	11.3305
pump2out - Molar Flow	0.1130	0.0056	0.0001	0.0000	7.2608
recycle - Molar Flow	2.6186	0.0455	0.0048	0.0010	13.4509
shellout - Molar Flow	1.0469	0.0137	0.0012	0.0000	11.8661
tot - Molar Flow	1.0510	0.0182	0.0014	0.0006	18.5920
TotB - Molar Flow	1.0556	0.0133	0.0011	0.0007	11.3305
tubeout - Molar Flow	1.0503	0.0183	0.0016	0.0009	18.5883
v10out - Molar Flow	0.0089	0.0470	0.0532	0.1331	0.1415
v1out - Molar Flow	1.6717	0.0586	0.0057	0.0007	23.6718
v2out - Molar Flow	0.1130	0.0056	0.0001	0.0000	7.2608
v4out - Molar Flow	1.0469	0.0137	0.0012	0.0000	11.8661
v6out - Molar Flow	7.9976	0.1825	0.0173	0.0010	26.2106
v7out - Molar Flow	2.6186	0.0455	0.0048	0.0010	13.4509
v8out - Molar Flow	1.6641	11.4751	1.5028	0.0018	15.6298

Table C.2 (Continued) IAE results of Flow rate for the process

STREAM	qhx1	qr	qhx2	qc1	qr1
B1 - Molar Flow	2.7327	9.8029	15.5931	1.5719	15.4915
B2 - Molar Flow	0.0175	0.0426	0.0762	0.0127	0.1946
D1 - Molar Flow	1.9408	5.4843	14.3197	2.0418	20.1338
D2 - Molar Flow	0.9028	6.9583	12.1395	1.3448	7.2760
fresh_B - Molar Flow	3.7294	13.1402	19.5597	2.2607	24.3387
fresh_c3 - Molar Flow	1.4266	5.8777	2.2873	0.1851	1.4182
gas - Molar Flow	0.8373	2.5520	12.8733	0.0960	12.5505
gout - Molar Flow	0.8373	2.5520	12.8733	0.0960	12.5505
htot - Molar Flow	3.4701	14.6023	7.5225	0.4035	5.6212
HX1out - Molar Flow	3.4702	14.6028	7.5228	0.4035	5.6208
HX2out - Molar Flow	2.1116	9.1289	13.8502	0.2303	4.2437
L1 - Molar Flow	5.1268	23.4124	36.9730	5.8655	31.1239
mix1out - Molar Flow	2.0438	8.7291	5.2403	0.2192	4.2047
p1 - Molar Flow	2.1054	9.1276	5.6526	0.2302	4.2375
p2 - Molar Flow	2.1054	9.1276	5.6526	0.2302	4.2375
p3out - Molar Flow	5.1268	23.4124	36.9730	5.8655	31.1239
p4out - Molar Flow	2.7327	9.8029	15.5931	1.5719	15.4915
p5out - Molar Flow	0.9028	6.9583	12.1395	1.3448	7.2760
p6out - Molar Flow	0.0175	0.0426	0.0762	0.0127	0.1946
product - Molar Flow	0.9028	6.9583	12.1395	1.3448	7.2760
pump1out - Molar Flow	2.0438	8.7291	5.2403	0.2192	4.2047
pump2out - Molar Flow	1.4266	5.8777	2.2873	0.1851	1.4182
recycle - Molar Flow	1.9408	5.4843	14.3197	2.0418	20.1338
shellout - Molar Flow	2.1041	9.1202	5.7056	0.2301	4.2379
tot - Molar Flow	3.4711	14.6061	7.5269	0.4037	5.6218
TotB - Molar Flow	2.0438	8.7291	5.2403	0.2192	4.2047
tubeout - Molar Flow	3.4701	14.6025	7.5225	0.4035	5.6210
v10out - Molar Flow	0.0175	0.0426	0.0762	0.0127	0.1946
v1out - Molar Flow	3.7294	13.1402	19.5597	2.2607	24.3387
v2out - Molar Flow	1.4266	5.8777	2.2873	0.1851	1.4182
v4out - Molar Flow	2.1041	9.1202	5.7056	0.2301	4.2379
v6out - Molar Flow	5.1268	23.4124	36.9730	5.8655	31.1239
v7out - Molar Flow	1.9408	5.4843	14.3197	2.0418	20.1338
v8out - Molar Flow	2.7327	9.8029	15.5931	1.5719	15.4915

Table C.2 (Continued) IAE results of Flow rate for the process

STREAM	qc2	qr2	c1-reflux	c2-reflux	SUM
B1 - Molar Flow	0.0725	6.1609	1.6525	1.3963	91.6355
B2 - Molar Flow	0.0026	0.2053	0.0168	0.0393	1.0805
D1 - Molar Flow	0.0012	0.0182	2.2366	0.0044	69.7318
D2 - Molar Flow	0.3678	8.6813	2.3056	5.2573	62.6678
fresh_B - Molar Flow	0.0008	0.0234	3.0275	0.0052	101.3334
fresh_c3 - Molar Flow	0.0000	0.0016	0.2804	0.0000	20.9588
gas - Molar Flow	0.0001	0.0002	0.0124	0.0001	36.8634
gout - Molar Flow	0.0001	0.0002	0.0124	0.0001	36.8634
htot - Molar Flow	0.0009	0.0077	1.0703	0.0015	56.4508
HX1out - Molar Flow	0.0007	0.0078	1.0703	0.0014	56.4506
HX2out - Molar Flow	0.0002	0.0053	0.8110	0.0012	46.6567
L1 - Molar Flow	0.0016	0.0734	9.8818	0.0156	197.5628
mix1out - Molar Flow	0.0006	0.0053	0.7912	0.0009	36.9643
p1 - Molar Flow	0.0000	0.0052	0.8110	0.0011	38.4124
p2 - Molar Flow	0.0000	0.0052	0.8110	0.0011	38.4124
p3out - Molar Flow	0.0016	0.0734	9.8818	0.0156	197.5628
p4out - Molar Flow	0.0725	6.1609	1.6525	1.3963	91.6355
p5out - Molar Flow	0.3678	8.6813	2.3056	5.2573	62.6678
p6out - Molar Flow	0.0026	0.2053	0.0168	0.0393	1.0805
product - Molar Flow	0.3678	8.6813	2.3056	5.2573	62.6678
pump1out - Molar Flow	0.0006	0.0053	0.7912	0.0009	36.9643
pump2out - Molar Flow	0.0000	0.0016	0.2804	0.0000	20.9588
recycle - Molar Flow	0.0012	0.0182	2.2366	0.0044	69.7318
shellout - Molar Flow	0.0000	0.0052	0.8104	0.0011	38.4471
tot - Molar Flow	0.0005	0.0078	1.0710	0.0013	56.4664
TotB - Molar Flow	0.0006	0.0053	0.7912	0.0009	36.9643
tubeout - Molar Flow	0.0008	0.0078	1.0703	0.0015	56.4503
v10out - Molar Flow	0.0026	0.2053	0.0168	0.0393	1.0805
v1out - Molar Flow	0.0008	0.0234	3.0275	0.0052	101.3334
v2out - Molar Flow	0.0000	0.0016	0.2804	0.0000	20.9588
v4out - Molar Flow	0.0000	0.0052	0.8104	0.0011	38.4471
v6out - Molar Flow	0.0016	0.0734	9.8818	0.0156	197.5628
v7out - Molar Flow	0.0012	0.0182	2.2366	0.0044	69.7318
v8out - Molar Flow	0.0725	6.1609	1.6525	1.3963	91.6355

Table C.3 IAE results of Pressure for the process

STREAM	v1	v2	v4	v5	v6
B1 - Pressure	0.3139	0.0196	0.0003	0.0770	0.0792
B2 - Pressure	0.0626	0.0028	0.0001	0.0186	0.0452
D1 - Pressure	0.3154	0.0196	0.0003	0.0774	0.0798
D2 - Pressure	0.0629	0.0028	0.0001	0.0188	0.0460
fresh_B - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
gas - Pressure	0.1422	0.0661	0.0003	0.1110	0.0114
gout - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
htot - Pressure	0.1167	0.0788	0.0002	0.0407	0.0429
HX1out - Pressure	0.1092	0.0767	0.0005	0.0349	0.0384
HX2out - Pressure	0.1422	0.0661	0.0003	0.1110	0.0114
L1 - Pressure	0.1422	0.0661	0.0003	0.1110	0.0114
mix1out - Pressure	0.3664	0.0192	0.0002	0.0604	0.0635
p1 - Pressure	0.1145	0.0777	0.0002	0.0402	0.0433
p2 - Pressure	0.1145	0.0777	0.0002	0.0402	0.0433
p3out - Pressure	0.1656	0.0496	0.0002	0.0465	0.2430
p5out - Pressure	0.0410	0.0019	0.0000	0.0119	0.0872
p6out - Pressure	0.0441	0.0020	0.0001	0.0130	0.0313
product - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
pump1out - Pressure	0.1342	0.0761	0.0002	0.0365	0.0392
pump2out - Pressure	0.1114	0.1746	0.0002	0.0351	0.0374
recycle - Pressure	0.3664	0.0192	0.0002	0.0604	0.0635
shellout - Pressure	0.1192	0.0774	0.0004	0.0482	0.0508
tot - Pressure	0.1227	0.0806	0.0002	0.0387	0.0411
TotB - Pressure	0.1227	0.0806	0.0002	0.0387	0.0411
tubeout - Pressure	0.1058	0.0786	0.0002	0.0298	0.0332
v10out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
v1out - Pressure	0.3664	0.0192	0.0002	0.0604	0.0635
v2out - Pressure	0.1227	0.0806	0.0002	0.0387	0.0411
v4out - Pressure	0.1260	0.0591	0.0004	0.0983	0.0100
v6out - Pressure	0.3150	0.0196	0.0003	0.0773	0.0796
v7out - Pressure	0.3664	0.0192	0.0002	0.0604	0.0635
v8out - Pressure	0.0627	0.0028	0.0001	0.0187	0.0457

Table C.3 (Continued) IAE results of Pressure for the process

STREAM	v7	v8	v9	v10	qv
B1 - Pressure	0.1089	0.0029	0.0003	0.0000	1.0479
B2 - Pressure	0.0126	0.0367	0.0743	0.0000	0.1827
D1 - Pressure	0.1095	0.0029	0.0003	0.0000	1.0516
D2 - Pressure	0.0126	0.0359	0.0752	0.0000	0.1831
fresh_B - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
gas - Pressure	0.0108	0.0004	0.0000	0.0000	1.4020
gout - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
htot - Pressure	0.0217	0.0009	0.0000	0.0000	1.4703
HX1out - Pressure	0.0195	0.0010	0.0002	0.0001	1.3073
HX2out - Pressure	0.0108	0.0004	0.0000	0.0000	1.4020
L1 - Pressure	0.0108	0.0004	0.0000	0.0000	1.4020
mix1out - Pressure	0.0632	0.0022	0.0002	0.0000	0.8565
p1 - Pressure	0.0214	0.0009	0.0001	0.0000	1.4007
p2 - Pressure	0.0214	0.0009	0.0001	0.0000	1.4007
p3out - Pressure	0.0398	0.0012	0.0001	0.0000	1.1174
p5out - Pressure	0.0304	0.0234	0.2265	0.0000	0.1222
p6out - Pressure	0.0088	0.0268	0.0517	0.0520	0.1317
product - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
pump1out - Pressure	0.0217	0.0009	0.0000	0.0000	1.3271
pump2out - Pressure	0.0193	0.0009	0.0001	0.0000	1.2673
recycle - Pressure	0.0632	0.0022	0.0002	0.0000	0.8565
shellout - Pressure	0.0246	0.0011	0.0000	0.0000	1.5468
tot - Pressure	0.0212	0.0009	0.0000	0.0000	1.3909
TotB - Pressure	0.0212	0.0009	0.0000	0.0000	1.3909
tubeout - Pressure	0.0178	0.0008	0.0001	0.0000	1.1978
v10out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
v1out - Pressure	0.0632	0.0022	0.0002	0.0000	0.8565
v2out - Pressure	0.0212	0.0009	0.0000	0.0000	1.3909
v4out - Pressure	0.0095	0.0004	0.0000	0.0000	1.2707
v6out - Pressure	0.1093	0.0029	0.0003	0.0000	1.0508
v7out - Pressure	0.0632	0.0022	0.0002	0.0000	0.8565
v8out - Pressure	0.0126	0.0358	0.0743	0.0000	0.1827

Table C.3 (Continued) IAE results of Pressure for the process

STREAM	qhx1	qr	qhx2	qc1	qr1
B1 - Pressure	0.1666	0.5885	0.9711	0.1084	1.1767
B2 - Pressure	0.0238	0.0594	0.1069	0.0178	0.2633
D1 - Pressure	0.1673	0.5918	0.9767	0.1089	1.1786
D2 - Pressure	0.0238	0.0592	0.1067	0.0178	0.2646
fresh_B - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
gas - Pressure	0.2497	0.7136	1.2899	0.0282	0.7814
gout - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
htot - Pressure	0.2852	1.1548	0.4348	0.0346	0.2630
HX1out - Pressure	0.3080	1.2550	0.4129	0.0318	0.2327
HX2out - Pressure	0.2497	0.7136	1.2899	0.0282	0.7814
L1 - Pressure	0.2497	0.7136	1.2899	0.0282	0.7814
mix1out - Pressure	0.1406	0.5123	0.7724	0.0862	0.8797
p1 - Pressure	0.3141	1.3058	0.4498	0.0338	0.2564
p2 - Pressure	0.3141	1.3058	0.4498	0.0338	0.2564
p3out - Pressure	0.2185	0.6550	1.1474	0.0549	0.9172
p5out - Pressure	0.0159	0.1430	0.2538	0.0298	0.1754
p6out - Pressure	0.0167	0.0414	0.0740	0.0124	0.1906
product - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
pump1out - Pressure	0.2563	1.0328	0.4245	0.0355	0.2819
pump2out - Pressure	0.2446	0.9854	0.3866	0.0316	0.2435
recycle - Pressure	0.1406	0.5123	0.7724	0.0862	0.8797
shellout - Pressure	0.3272	1.3690	0.5056	0.0371	0.2895
tot - Pressure	0.2692	1.0873	0.4260	0.0346	0.2679
TotB - Pressure	0.2692	1.0873	0.4260	0.0346	0.2679
tubeout - Pressure	0.2883	1.1713	0.3746	0.0292	0.2108
v10out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
v1out - Pressure	0.1406	0.5123	0.7724	0.0862	0.8797
v2out - Pressure	0.2692	1.0873	0.4260	0.0346	0.2679
v4out - Pressure	0.2233	0.6285	1.1227	0.0251	0.7042
v6out - Pressure	0.1671	0.5911	0.9757	0.1088	1.1773
v7out - Pressure	0.1406	0.5123	0.7724	0.0862	0.8797
v8out - Pressure	0.0238	0.0595	0.1072	0.0178	0.2637

Table C.3 (Continued) IAE results of Pressure for the process

STREAM	qc2	qr2	c1-reflux	c2-reflux	SUM
B1 - Pressure	0.0000	0.0012	0.1476	0.0002	4.8101
B2 - Pressure	0.0036	0.3030	0.0242	0.0681	1.3057
D1 - Pressure	0.0000	0.0012	0.1488	0.0002	4.8302
D2 - Pressure	0.0036	0.3051	0.0243	0.0706	1.3131
fresh_B - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
gas - Pressure	0.0000	0.0002	0.0162	0.0000	4.8236
gout - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
htot - Pressure	0.0000	0.0005	0.0533	0.0000	3.9985
HX1out - Pressure	0.0001	0.0003	0.0479	0.0002	3.8763
HX2out - Pressure	0.0000	0.0002	0.0162	0.0000	4.8236
L1 - Pressure	0.0000	0.0002	0.0162	0.0000	4.8236
mix1out - Pressure	0.0000	0.0009	0.1157	0.0002	3.9398
p1 - Pressure	0.0000	0.0004	0.0525	0.0000	4.1120
p2 - Pressure	0.0000	0.0004	0.0525	0.0000	4.1120
p3out - Pressure	0.0000	0.0005	0.0592	0.0001	4.7163
p5out - Pressure	0.0079	0.2021	0.0584	0.1134	1.5443
p6out - Pressure	0.0025	0.2173	0.0169	0.0468	0.9801
product - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
pump1out - Pressure	0.0000	0.0005	0.0525	0.0000	3.7200
pump2out - Pressure	0.0000	0.0003	0.0478	0.0001	3.5862
recycle - Pressure	0.0000	0.0009	0.1157	0.0002	3.9398
shellout - Pressure	0.0000	0.0005	0.0596	0.0000	4.4571
tot - Pressure	0.0000	0.0005	0.0525	0.0000	3.8344
TotB - Pressure	0.0000	0.0005	0.0525	0.0000	3.8344
tubeout - Pressure	0.0000	0.0003	0.0429	0.0000	3.5814
v10out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000
v1out - Pressure	0.0000	0.0009	0.1157	0.0002	3.9398
v2out - Pressure	0.0000	0.0005	0.0525	0.0000	3.8344
v4out - Pressure	0.0000	0.0002	0.0144	0.0000	4.2929
v6out - Pressure	0.0000	0.0012	0.1483	0.0002	4.8247
v7out - Pressure	0.0000	0.0009	0.1157	0.0002	3.9398
v8out - Pressure	0.0036	0.3034	0.0243	0.0691	1.3077

Table C.4 IAE results of Temperature for the process

STREAM	v1	v2	v4	v5
B1 - Temperature	7.7128	0.3904	0.0068	2.5577
B2 - Temperature	1.9424	0.0864	0.0027	0.6857
D1 - Temperature	11.7022	0.5251	0.0135	8.2487
D2 - Temperature	2.5667	0.0463	0.0035	1.3460
fresh_B - Temperature	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Temperature	0.0000	0.0000	0.0000	0.0000
gas - Temperature	3.1147	2.1931	0.0090	2.2522
gout - Temperature	10.2964	2.1925	0.0090	2.2510
htot - Temperature	1.4687	0.0783	0.0036	1.2002
HX1out - Temperature	1.0668	1.6264	0.0039	1.6556
HX2out - Temperature	3.1964	2.2462	0.0091	2.4110
L1 - Temperature	3.1148	2.1931	0.0090	2.2522
mix1out - Temperature	5.9522	0.3191	0.0079	4.6639
p1 - Temperature	0.7921	3.4807	0.0025	1.1298
p2 - Temperature	0.7915	3.4793	0.0025	1.1290
p3out - Temperature	3.1194	2.1910	0.0090	2.2560
p4out - Temperature	7.7072	0.3901	0.0067	2.5562
p5out - Temperature	2.5650	0.0464	0.0034	1.3454
p6out - Temperature	1.9406	0.0863	0.0026	0.6853
product - Temperature	0.3385	0.0982	0.0011	0.6602
pump1out - Temperature	5.9164	0.3230	0.0079	4.6487
pump2out - Temperature	0.0111	0.0174	0.0000	0.0035
recycle - Temperature	12.7994	0.5655	0.0138	8.3360
shellout - Temperature	0.9665	1.0933	0.0041	1.2684
tot - Temperature	4.1697	0.1917	0.0055	3.2480
TotB - Temperature	5.9164	0.3230	0.0079	4.6487
tubeout - Temperature	1.0045	1.7735	0.0033	1.5477
v10out - Temperature	0.4197	0.0219	0.0013	0.0324
v1out - Temperature	0.0000	0.0000	0.0000	0.0000
v4out - Temperature	3.1421	2.3387	0.0086	3.7048
v2out - Temperature	0.0111	0.0174	0.0000	0.0035
v6out - Temperature	3.1193	2.1910	0.0090	2.2557
v7out - Temperature	12.7994	0.5655	0.0138	8.3360
v8out - Temperature	2.5037	0.0718	0.0033	1.1155

Table C.4 (Continued) IAE results of Temperature for the process

STREAM	v6	v7	v8	v9
B1 - Temperature	7.9514	2.9215	0.0891	0.0081
B2 - Temperature	2.3810	0.4419	4.3440	2.5066
D1 - Temperature	7.6667	4.6645	0.1346	0.0125
D2 - Temperature	7.8325	0.6821	1.2806	2.9270
fresh_B - Temperature	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Temperature	0.0000	0.0000	0.0000	0.0000
gas - Temperature	2.6828	1.6098	0.0408	0.0037
gout - Temperature	2.6824	1.6091	0.0407	0.0036
htot - Temperature	1.1391	0.5891	0.0193	0.0024
HX1out - Temperature	1.5410	1.3194	0.0228	0.0028
HX2out - Temperature	2.7374	1.6755	0.0426	0.0039
L1 - Temperature	2.6829	1.6098	0.0408	0.0037
mix1out - Temperature	4.3079	2.4062	0.0797	0.0074
p1 - Temperature	1.0457	1.2422	0.0111	0.0011
p2 - Temperature	1.0453	1.2415	0.0111	0.0011
p3out - Temperature	2.5455	1.6121	0.0408	0.0037
p4out - Temperature	7.9495	2.9184	0.0986	0.0093
p5out - Temperature	7.7704	0.6597	1.2797	2.8588
p6out - Temperature	2.3794	0.4415	4.3438	2.5045
product - Temperature	7.5670	0.5161	0.0416	2.6167
pump1out - Temperature	4.2935	2.4015	0.0794	0.0074
pump2out - Temperature	0.0038	0.0019	0.0001	0.0000
recycle - Temperature	7.8206	3.8221	0.1367	0.0127
shellout - Temperature	1.1884	0.8784	0.0183	0.0022
tot - Temperature	3.0049	1.6737	0.0556	0.0052
TotB - Temperature	4.2935	2.4015	0.0794	0.0074
tubeout - Temperature	1.4406	1.2114	0.0210	0.0025
v10out - Temperature	0.5731	0.0461	5.3192	0.3948
v1out - Temperature	0.0000	0.0000	0.0000	0.0000
v4out - Temperature	3.4172	2.6722	0.0572	0.0055
v2out - Temperature	0.0038	0.0019	0.0001	0.0000
v6out - Temperature	2.7486	1.7879	0.0408	0.0037
v7out - Temperature	7.8206	3.8221	0.1367	0.0127
v8out - Temperature	5.3561	0.6431	1.2964	2.7628

Table C.4 (Continued) IAE results of Temperature for the process

STREAM	v10	qv	qhx1	qr
B1 - Temperature	0.0006	21.8014	3.2128	14.4217
B2 - Temperature	0.0008	2.8417	0.6472	2.0450
D1 - Temperature	0.0006	43.5456	7.0299	30.4343
D2 - Temperature	0.0004	7.1853	0.4998	1.7234
fresh_B - Temperature	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Temperature	0.0000	0.0000	0.0000	0.0000
gas - Temperature	0.0002	50.6931	9.3481	34.3206
gout - Temperature	0.0002	50.6787	9.3455	41.8946
htot - Temperature	0.0005	20.0984	1.9764	8.4055
HX1out - Temperature	0.0014	24.2010	15.4975	25.5415
HX2out - Temperature	0.0002	51.5273	9.5733	35.6075
L1 - Temperature	0.0002	50.6931	9.3482	34.3216
mix1out - Temperature	0.0003	27.7412	4.2141	16.9895
p1 - Temperature	0.0000	8.7874	11.9723	45.4188
p2 - Temperature	0.0000	8.7824	11.9674	45.4004
p3out - Temperature	0.0002	50.9326	9.3427	34.3058
p4out - Temperature	0.0008	21.7836	3.2099	14.4026
p5out - Temperature	0.0003	7.1812	0.4996	1.6516
p6out - Temperature	0.0047	2.8375	0.6465	2.0427
product - Temperature	0.0007	2.4102	0.8192	1.5927
pump1out - Temperature	0.0003	27.7056	4.2122	16.9866
pump2out - Temperature	0.0000	0.1261	0.0243	0.0981
recycle - Temperature	0.0006	43.8839	7.2097	31.7439
shellout - Temperature	0.0000	19.4117	3.6481	14.5430
tot - Temperature	0.0002	19.5337	2.9735	11.9619
TotB - Temperature	0.0003	27.7056	4.2122	16.9866
tubeout - Temperature	0.0015	22.5563	6.2275	24.7771
v10out - Temperature	0.0010	3.4957	0.2515	0.2820
v1out - Temperature	0.0000	0.0000	0.0000	0.0000
v4out - Temperature	0.0007	71.4943	13.1691	50.3233
v2out - Temperature	0.0000	0.1261	0.0243	0.0981
v6out - Temperature	0.0002	49.5144	9.2112	34.5489
v7out - Temperature	0.0006	43.8839	7.2097	31.7439
v8out - Temperature	0.0006	6.8702	0.5665	2.0289

Table C.4 (Continued) IAE results of Temperature for the process

STREAM	qhx2	qc1	qr1	qc2
B1 - Temperature	26.6367	2.8392	27.3261	0.0006
B2 - Temperature	3.8415	0.5969	6.3631	0.1118
D1 - Temperature	51.9031	3.5785	34.3034	0.0008
D2 - Temperature	4.5536	0.8672	10.4489	0.1459
fresh_B - Temperature	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Temperature	0.0000	0.0000	0.0000	0.0000
gas - Temperature	64.5432	1.0560	13.5854	0.0003
gout - Temperature	71.9236	1.0558	18.1482	0.0002
htot - Temperature	7.4520	0.4791	6.3755	0.0006
HX1out - Temperature	8.4638	0.4983	6.1242	0.0015
HX2out - Temperature	66.6529	1.0773	13.8079	0.0003
L1 - Temperature	64.5930	1.0560	13.5802	0.0003
mix1out - Temperature	28.2848	2.1933	23.0381	0.0005
p1 - Temperature	3.8389	0.1323	3.0190	0.0000
p2 - Temperature	3.8361	0.1322	3.0170	0.0000
p3out - Temperature	64.5640	1.0580	13.5882	0.0003
p4out - Temperature	26.6047	2.8364	27.3093	0.0007
p5out - Temperature	4.3939	0.8606	10.4419	0.1398
p6out - Temperature	3.8370	0.5964	6.3569	0.1116
product - Temperature	4.2965	0.7513	1.0339	0.0089
pump1out - Temperature	28.1655	2.1821	22.9076	0.0005
pump2out - Temperature	0.0385	0.0032	0.0242	0.0000
recycle - Temperature	53.9161	3.6500	34.4242	0.0008
shellout - Temperature	6.8284	0.4005	5.8113	0.0000
tot - Temperature	19.7437	1.5334	16.1120	0.0003
TotB - Temperature	28.1655	2.1821	22.9076	0.0005
tubeout - Temperature	7.8452	0.4633	6.2318	0.0015
v10out - Temperature	0.3970	0.0903	2.9303	0.0271
v1out - Temperature	0.0000	0.0000	0.0000	0.0000
v4out - Temperature	20.6734	1.4525	15.2552	0.0008
v2out - Temperature	0.0385	0.0032	0.0242	0.0000
v6out - Temperature	64.8120	1.1190	13.5891	0.0003
v7out - Temperature	53.9161	3.6500	34.4242	0.0008
v8out - Temperature	5.0206	0.8030	10.1504	0.1328

Table C.4 (Continued) IAE results of Temperature for the process

STREAM	qr2	c1-reflux	c2-reflux	SUM
B1 - Temperature	0.0342	5.6387	0.0071	123.5569
B2 - Temperature	12.1550	0.9573	4.8841	46.8351
D1 - Temperature	0.0534	7.5376	0.0112	211.3662
D2 - Temperature	11.1267	2.7270	2.7846	58.7475
fresh_B - Temperature	0.0000	0.0000	0.0000	0.0000
fresh_c3 - Temperature	0.0000	0.0000	0.0000	0.0000
gas - Temperature	0.0157	2.3061	0.0033	187.7779
gout - Temperature	0.0157	2.3051	0.0033	214.4558
htot - Temperature	0.0069	1.0720	0.0022	50.3697
HX1out - Temperature	0.0088	1.2888	0.0026	88.8679
HX2out - Temperature	0.0166	2.3877	0.0035	192.9765
L1 - Temperature	0.0157	2.3061	0.0033	187.8239
mix1out - Temperature	0.0316	4.3881	0.0067	124.6326
p1 - Temperature	0.0036	0.6621	0.0010	81.5406
p2 - Temperature	0.0036	0.6616	0.0010	81.5030
p3out - Temperature	0.0157	2.3088	0.0033	187.8971
p4out - Temperature	0.0399	5.6346	0.0082	123.4668
p5out - Temperature	11.1185	2.6826	2.7142	58.2131
p6out - Temperature	12.1473	0.9565	4.8817	46.8026
product - Temperature	0.1968	2.4948	2.4723	27.9167
pump1out - Temperature	0.0315	4.3695	0.0066	124.2458
pump2out - Temperature	0.0000	0.0048	0.0000	0.3570
recycle - Temperature	0.0541	7.6730	0.0114	216.0745
shellout - Temperature	0.0062	1.0214	0.0019	57.0920
tot - Temperature	0.0221	3.0626	0.0047	87.3023
TotB - Temperature	0.0315	4.3695	0.0066	124.2458
tubeout - Temperature	0.0086	1.2098	0.0023	76.3294
v10out - Temperature	0.6640	0.0325	2.0930	17.0730
v1out - Temperature	0.0000	0.0000	0.0000	0.0000
v4out - Temperature	0.0225	3.1690	0.0050	190.9120
v2out - Temperature	0.0000	0.0048	0.0000	0.3570
v6out - Temperature	0.0157	2.5827	0.0033	187.5526
v7out - Temperature	0.0541	7.6730	0.0114	216.0745
v8out - Temperature	10.7378	2.1530	2.5649	54.7813

Table C.5 IAE results of Level for the process Unit

UNITS	v1	v2	v4
Condenser - Liquid Percent Level	3.2138	0.0618	0.0008
Reboiler - Vessel Liq Percent Level	12.1766	2.2702	0.0109
Condenser - Liquid Percent Level	0.2826	0.0491	0.0003
Reboiler - Vessel Liq Percent Level	2.9710	0.2455	0.0034
flash tank - Liquid Percent Level	79.7253	14.5003	0.0365

Table C.5 (Continued) IAE results of Level for the process Unit

UNITS	v5	v6	v7
Condenser - Liquid Percent Level	1.0259	0.7085	1.3348
Reboiler - Vessel Liq Percent Level	9.2361	19.3644	3.7936
Condenser - Liquid Percent Level	0.1054	0.7196	0.3443
Reboiler - Vessel Liq Percent Level	0.4130	2.9000	0.4915
flash tank - Liquid Percent Level	44.7662	106.1567	23.2982

Table C.5 (Continued) IAE results of Level for the process Unit

UNITS	v8	v9	v10
Condenser - Liquid Percent Level	0.0039	0.0006	0.0001
Reboiler - Vessel Liq Percent Level	12.0526	1.1186	0.0004
Condenser - Liquid Percent Level	0.2350	1.6441	0.0002
Reboiler - Vessel Liq Percent Level	14.7923	4.5490	0.2476
flash tank - Liquid Percent Level	0.5815	0.0422	0.0023

Table C.5 (Continued) IAE results of Level for the process Unit

UNITS	qv	qhx1	qr
Condenser - Liquid Percent Level	2.2917	0.4233	2.5212
Reboiler - Vessel Liq Percent Level	14.9192	3.3826	9.9127
Condenser - Liquid Percent Level	1.7106	0.4116	2.1906
Reboiler - Vessel Liq Percent Level	11.8620	1.9965	3.5110
flash tank - Liquid Percent Level	129.1800	44.0668	115.5010

Table C.5 (Continued) IAE results of Level for the process Unit

UNITS	qhx2	qc1	qr1
Condenser - Liquid Percent Level	2.0486	6.9542	5.4012
Reboiler - Vessel Liq Percent Level	15.7641	5.4017	46.5516
Condenser - Liquid Percent Level	3.4422	0.3466	1.2538
Reboiler - Vessel Liq Percent Level	5.4186	0.8659	11.5114
flash tank - Liquid Percent Level	131.5488	29.1062	107.7882

Table C.5 (Continued) IAE results of Level for the process Unit

UNITS	qc2	qr2	c1-reflux
Condenser - Liquid Percent Level	0.0001	0.0022	2.0584
Reboiler - Vessel Liq Percent Level	0.1020	4.8027	8.4925
Condenser - Liquid Percent Level	7.8243	1.8433	0.2436
Reboiler - Vessel Liq Percent Level	0.2204	14.6592	0.2688
flash tank - Liquid Percent Level	0.0044	0.1867	30.5070

Table C.5 (Continued) IAE results of Level for the process Unit

UNITS	c2-reflux	SUM
Condenser - Liquid Percent Level	0.0005	28.0517
Reboiler - Vessel Liq Percent Level	1.0458	170.3980
Condenser - Liquid Percent Level	1.6470	24.2941
Reboiler - Vessel Liq Percent Level	8.1600	85.0871
flash tank - Liquid Percent Level	0.0374	857.0358

Table C.6 IAE results of Pressure for the process Unit

UNITS	v1	v2	v4	v5
Condenser - Vessel Pressure	0.31540	0.01962	0.00025	0.07741
Reboiler - Vessel Pressure	0.31394	0.01955	0.00025	0.07699
Condenser - Vessel Pressure	0.06291	0.00284	0.00008	0.01875
Reboiler - Vessel Pressure	0.06259	0.00284	0.00008	0.01858
flash tank - Vessel Pressure	0.14222	0.06610	0.00035	0.11103

Table C.6 (Continued) IAE results of Pressure for the process Unit

UNITS	v6	v7	v8	v9
Condenser - Vessel Pressure	0.07976	0.10946	0.00287	0.00027
Reboiler - Vessel Pressure	0.07922	0.10885	0.00286	0.00027
Condenser - Vessel Pressure	0.04599	0.01258	0.03587	0.07520
Reboiler - Vessel Pressure	0.04518	0.01258	0.03673	0.07432
flash tank - Vessel Pressure	0.01135	0.01082	0.00040	0.00004

Table C.6 (Continued) IAE results of Pressure for the process Unit

UNITS	v10	qv	qhx1	qr
Condenser - Vessel Pressure	0.00001	1.36678	0.16728	0.59176
Reboiler - Vessel Pressure	0.00001	1.37563	0.16655	0.58849
Condenser - Vessel Pressure	0.00002	0.36363	0.02379	0.05923
Reboiler - Vessel Pressure	0.00002	0.37958	0.02379	0.05944
flash tank - Vessel Pressure	0.00001	1.69648	0.24968	0.71360

Table C.6 (Continued) IAE results of Pressure for the process Unit

UNITS	qhx2	qc1	qr1	qc2
Condenser - Vessel Pressure	0.97673	0.10890	1.17860	0.00002
Reboiler - Vessel Pressure	0.97110	0.10841	1.17668	0.00002
Condenser - Vessel Pressure	0.10672	0.01779	0.26457	0.00364
Reboiler - Vessel Pressure	0.10694	0.01776	0.26327	0.00360
flash tank - Vessel Pressure	1.28991	0.02823	0.78136	0.00002

Table C.6 (Continued) IAE results of Pressure for the process Unit

UNITS	qr2	c1-reflux	c2-reflux	SUM
Condenser - Vessel Pressure	0.00116	0.14877	0.00024	5.14530
Reboiler - Vessel Pressure	0.00115	0.14764	0.00024	5.13788
Condenser - Vessel Pressure	0.30511	0.02434	0.07056	1.49363
Reboiler - Vessel Pressure	0.30300	0.02417	0.06811	1.50258
flash tank - Vessel Pressure	0.00017	0.01623	0.00004	5.11806

Table C.7 IAE results of Temperature for the process Unit

UNITS	v1	v2	v4	v5
Condenser - Vessel Temperature	11.7053	0.5253	0.0135	8.2531
Reboiler - Vessel Temperature	7.7148	0.3905	0.0068	2.5586
Condenser - Vessel Temperature	2.5677	0.0464	0.0035	1.3465
Reboiler - Vessel Temperature	1.9427	0.0864	0.0027	0.6859
flash tank - Vessel Temperature	3.1148	2.1931	0.0090	2.2522

Table C.7 (Continued) IAE results of Temperature for the process Unit

UNITS	v6	v7	v8	v9
Condenser - Vessel Temperature	7.6676	4.6660	0.1347	0.0126
Reboiler - Vessel Temperature	7.9525	2.9217	0.0891	0.0082
Condenser - Vessel Temperature	7.7764	0.6613	1.2810	2.8562
Reboiler - Vessel Temperature	2.3814	0.4420	4.3450	2.5068
flash tank - Vessel Temperature	2.6829	1.6098	0.0408	0.0037

Table C.7 (Continued) IAE results of Temperature for the process Unit

UNITS	v10	qv	qhx1	qr
Condenser - Vessel Temperature	0.0006	43.5606	7.0326	30.4449
Reboiler - Vessel Temperature	0.0009	21.8074	3.2138	14.4200
Condenser - Vessel Temperature	0.0004	7.1883	0.5001	1.6570
Reboiler - Vessel Temperature	0.0008	2.8422	0.6473	2.0450
flash tank - Vessel Temperature	0.0002	50.6931	9.3482	34.3216

Table C.7 (Continued) IAE results of Temperature for the process Unit

UNITS	qhx2	qc1	qr1	qc2
Condenser - Vessel Temperature	51.9192	3.5769	34.3074	0.0008
Reboiler - Vessel Temperature	26.6359	2.8389	27.3325	0.0009
Condenser - Vessel Temperature	4.4066	0.8617	10.4523	0.1403
Reboiler - Vessel Temperature	3.8412	0.5970	6.3640	0.1117
flash tank - Vessel Temperature	64.5930	1.0560	13.5802	0.0003

Table C.7 (Continued) IAE results of Temperature for the process Unit

UNITS	qr2	c1-reflux	c2-reflux	SUM
Condenser - Vessel Temperature	0.0534	7.5397	0.0112	211.4253
Reboiler - Vessel Temperature	0.0343	5.6399	0.0073	123.5739
Condenser - Vessel Temperature	11.1293	2.6866	2.7198	58.2812
Reboiler - Vessel Temperature	12.1565	0.9575	4.8850	46.8412
flash tank - Vessel Temperature	0.0157	2.3061	0.0033	187.8239

Table C.8 IAE results of Tray Temperature for Benzene recycle column

TRAY	v1	v2	v4	v5
Stage Temperature (1__Main TS)	6.6533	0.4314	0.0060	2.2214
Stage Temperature (2__Main TS)	6.6207	0.4797	0.0062	2.4332
Stage Temperature (3__Main TS)	6.6517	0.5884	0.0069	3.2177
Stage Temperature (4__Main TS)	6.7225	0.7503	0.0077	4.4154
Stage Temperature (5__Main TS)	6.8656	0.8459	0.0079	5.1870
Stage Temperature (6__Main TS)	7.0807	0.8000	0.0069	5.0160
Stage Temperature (7__Main TS)	9.0371	0.6373	0.0084	12.2707
Stage Temperature (8__Main TS)	14.9707	0.1350	0.0165	29.6276
Stage Temperature (9__Main TS)	24.3823	1.4191	0.0390	48.3734
Stage Temperature (10__Main TS)	26.2157	2.7105	0.0554	45.4682
Stage Temperature (11__Main TS)	17.8459	1.9649	0.0395	25.1804
Stage Temperature (12__Main TS)	11.2159	0.5917	0.0179	10.4038
Stage Temperature (13__Main TS)	8.5635	0.1426	0.0088	4.4529

Table C.8 (Continued) IAE results of Tray Temperature for Benzene recycle column

TRAY	v6	v7	v8	v9
Stage Temperature (1__Main TS)	2.1888	2.4655	0.0666	0.0064
Stage Temperature (2__Main TS)	2.2621	2.4511	0.0689	0.0066
Stage Temperature (3__Main TS)	2.6562	2.4798	0.0765	0.0071
Stage Temperature (4__Main TS)	3.1689	2.5354	0.0861	0.0078
Stage Temperature (5__Main TS)	3.3114	2.6092	0.0890	0.0080
Stage Temperature (6__Main TS)	2.8682	2.6798	0.0815	0.0075
Stage Temperature (7__Main TS)	4.6483	2.9801	0.1248	0.0108
Stage Temperature (8__Main TS)	12.1461	4.3612	0.2810	0.0226
Stage Temperature (9__Main TS)	33.5974	7.8980	0.6336	0.0493
Stage Temperature (10__Main TS)	67.0745	11.4690	0.8861	0.0681
Stage Temperature (11__Main TS)	77.6639	9.6020	0.6450	0.0497
Stage Temperature (12__Main TS)	51.4466	5.7162	0.3075	0.0244
Stage Temperature (13__Main TS)	22.2523	3.6453	0.1443	0.0122

Table C.8 (Continued) IAE results of Tray Temperature for Benzene recycle column

TRAY	v10	qv	qhx1	qr
Stage Temperature (1__Main TS)	0.0009	22.6937	3.7448	14.4345
Stage Temperature (2__Main TS)	0.0011	24.3513	3.8372	14.3391
Stage Temperature (3__Main TS)	0.0010	27.8292	4.1185	14.7043
Stage Temperature (4__Main TS)	0.0007	31.0632	4.4230	14.9687
Stage Temperature (5__Main TS)	0.0007	31.2374	4.4160	14.6261
Stage Temperature (6__Main TS)	0.0009	28.9314	4.0735	13.8049
Stage Temperature (7__Main TS)	0.0011	34.2255	2.9792	12.0465
Stage Temperature (8__Main TS)	0.0016	42.4699	1.7997	26.7536
Stage Temperature (9__Main TS)	0.0029	53.2064	9.1852	46.7924
Stage Temperature (10__Main TS)	0.0039	59.5346	18.2173	43.5173
Stage Temperature (11__Main TS)	0.0029	49.3872	14.4749	21.4953
Stage Temperature (12__Main TS)	0.0016	31.8115	5.0388	10.3394
Stage Temperature (13__Main TS)	0.0011	22.1117	1.5361	12.5446

Table C.8 (Continued) IAE results of Tray Temperature for Benzene recycle column

TRAY	qhx2	qc1	qr1	qc2
Stage Temperature (1__Main TS)	24.9227	2.3732	37.3742	0.0010
Stage Temperature (2__Main TS)	24.9121	2.3719	52.3395	0.0011
Stage Temperature (3__Main TS)	26.1452	2.3878	62.2491	0.0010
Stage Temperature (4__Main TS)	27.7815	2.4077	64.7647	0.0008
Stage Temperature (5__Main TS)	28.1835	2.4165	60.6124	0.0007
Stage Temperature (6__Main TS)	26.2495	2.4070	54.0606	0.0009
Stage Temperature (7__Main TS)	24.3267	2.4647	94.3564	0.0012
Stage Temperature (8__Main TS)	41.7880	4.3680	129.4070	0.0021
Stage Temperature (9__Main TS)	63.5566	10.9922	135.5221	0.0048
Stage Temperature (10__Main TS)	65.8051	16.4673	105.0612	0.0067
Stage Temperature (11__Main TS)	48.9833	12.3444	63.7392	0.0049
Stage Temperature (12__Main TS)	35.6291	5.6331	39.4037	0.0023
Stage Temperature (13__Main TS)	29.4614	3.3984	30.1753	0.0011

Table C.8 (Continued) IAE results of Tray Temperature for Benzene recycle column

TRAY	qr2	c1-reflux	c2-reflux	SUM
Stage Temperature (1__Main TS)	0.0267	3.6921	0.0058	123.3089
Stage Temperature (2__Main TS)	0.0274	4.0762	0.0060	140.5915
Stage Temperature (3__Main TS)	0.0299	5.1461	0.0064	158.3029
Stage Temperature (4__Main TS)	0.0332	6.4121	0.0069	169.5567
Stage Temperature (5__Main TS)	0.0341	6.3355	0.0071	166.7940
Stage Temperature (6__Main TS)	0.0318	4.3519	0.0067	152.4595
Stage Temperature (7__Main TS)	0.0461	5.6052	0.0096	205.7797
Stage Temperature (8__Main TS)	0.0985	10.4545	0.0200	318.7235
Stage Temperature (9__Main TS)	0.2161	23.5787	0.0435	459.4930
Stage Temperature (10__Main TS)	0.2977	40.8080	0.0601	503.7267
Stage Temperature (11__Main TS)	0.2164	40.6305	0.0439	384.3141
Stage Temperature (12__Main TS)	0.1056	24.0653	0.0216	231.7758
Stage Temperature (13__Main TS)	0.0525	11.0502	0.0108	149.5651

Table C.9 IAE results of Tray Temperature for Cumene column

STREAM	v1	v2	v4	v5
Stage Temperature (1__Main TS)	2.4431	0.0921	0.0031	0.8364
Stage Temperature (2__Main TS)	2.4184	0.0971	0.0030	0.7979
Stage Temperature (3__Main TS)	2.4070	0.0968	0.0031	0.7927
Stage Temperature (4__Main TS)	2.3962	0.0964	0.0029	0.7897
Stage Temperature (5__Main TS)	2.3846	0.0957	0.0030	0.7871
Stage Temperature (6__Main TS)	2.3708	0.0950	0.0029	0.7843
Stage Temperature (7__Main TS)	2.3524	0.0939	0.0028	0.7815
Stage Temperature (8__Main TS)	2.3241	0.0921	0.0028	0.7783
Stage Temperature (9__Main TS)	2.2754	0.0889	0.0028	0.7737
Stage Temperature (10__Main TS)	2.1854	0.0828	0.0027	0.7655
Stage Temperature (11__Main TS)	2.0168	0.0712	0.0024	0.7503
Stage Temperature (12__Main TS)	1.7072	0.0502	0.0021	0.7194
Stage Temperature (13__Main TS)	0.5970	0.1352	0.0010	0.3725
Stage Temperature (14__Main TS)	7.3950	0.6559	0.0085	0.4661
Stage Temperature (15__Main TS)	14.3186	1.0174	0.0135	1.0488
Stage Temperature (16__Main TS)	10.2592	0.6186	0.0076	0.4692
Stage Temperature (17__Main TS)	2.9552	0.1720	0.0010	0.2680
Stage Temperature (18__Main TS)	0.6920	0.0228	0.0018	0.5819

Table C.9 (Continued) IAE results of Tray Temperature for Cumene column

STREAM	v6	v7	v8	v9
Stage Temperature (1__Main TS)	2.8701	0.5233	1.3652	2.9310
Stage Temperature (2__Main TS)	2.5248	0.5110	1.3616	2.9157
Stage Temperature (3__Main TS)	2.4971	0.5088	1.3540	2.8985
Stage Temperature (4__Main TS)	2.4916	0.5071	1.3446	2.8788
Stage Temperature (5__Main TS)	2.4896	0.5055	1.3314	2.8543
Stage Temperature (6__Main TS)	2.4914	0.5036	1.3107	2.8199
Stage Temperature (7__Main TS)	2.5002	0.5009	1.2750	2.7649
Stage Temperature (8__Main TS)	2.5227	0.4968	1.2092	2.6693
Stage Temperature (9__Main TS)	2.5713	0.4891	1.0844	2.4939
Stage Temperature (10__Main TS)	2.6691	0.4737	0.8461	2.1692
Stage Temperature (11__Main TS)	2.8555	0.4426	0.3981	1.5820
Stage Temperature (12__Main TS)	3.1922	0.3826	0.4325	0.5908
Stage Temperature (13__Main TS)	5.0337	0.1271	6.6292	6.1705
Stage Temperature (14__Main TS)	12.4647	1.2620	27.5932	19.7243
Stage Temperature (15__Main TS)	21.0283	1.9833	61.9980	22.8449
Stage Temperature (16__Main TS)	17.3633	1.0598	70.9977	11.0796
Stage Temperature (17__Main TS)	8.6514	0.1385	44.8945	2.0996
Stage Temperature (18__Main TS)	4.0043	0.3093	17.5795	1.4054

Table C.9 (Continued) IAE results of Tray Temperature for Cumene column

STREAM	v10	qv	qhx1	qr
Stage Temperature (1__Main TS)	0.0007	6.4859	0.7668	2.1765
Stage Temperature (2__Main TS)	0.0007	6.5365	0.8084	2.2116
Stage Temperature (3__Main TS)	0.0007	6.5152	0.8071	2.2056
Stage Temperature (4__Main TS)	0.0007	6.4864	0.8026	2.1961
Stage Temperature (5__Main TS)	0.0006	6.4551	0.7973	2.1849
Stage Temperature (6__Main TS)	0.0007	6.4189	0.7903	2.1697
Stage Temperature (7__Main TS)	0.0007	6.3726	0.7800	2.1453
Stage Temperature (8__Main TS)	0.0007	6.3053	0.7626	2.1003
Stage Temperature (9__Main TS)	0.0007	6.1965	0.7309	2.0104
Stage Temperature (10__Main TS)	0.0008	6.0129	0.6704	1.8271
Stage Temperature (11__Main TS)	0.0008	5.7058	0.5560	1.4600
Stage Temperature (12__Main TS)	0.0009	5.2220	0.3461	0.7747
Stage Temperature (13__Main TS)	0.0015	2.8532	1.2128	3.7785
Stage Temperature (14__Main TS)	0.0032	16.1613	5.9272	13.3556
Stage Temperature (16__Main TS)	0.0029	47.7452	6.9489	7.7713
Stage Temperature (17__Main TS)	0.0016	27.0648	2.2775	1.3080
Stage Temperature (18__Main TS)	0.0011	6.7805	0.0990	1.2408

Table C.9 (Continued) IAE results of Tray Temperature for Cumene column

STREAM	qhx2	qc1	qr1	qc2
Stage Temperature (1__Main TS)	4.2717	0.7148	9.9118	0.1404
Stage Temperature (2__Main TS)	4.2495	0.7048	9.8163	0.1396
Stage Temperature (3__Main TS)	4.2346	0.7017	9.7728	0.1389
Stage Temperature (4__Main TS)	4.2190	0.6988	9.7324	0.1380
Stage Temperature (5__Main TS)	4.2002	0.6955	9.6908	0.1370
Stage Temperature (6__Main TS)	4.1733	0.6912	9.6454	0.1355
Stage Temperature (7__Main TS)	4.1272	0.6848	9.5913	0.1334
Stage Temperature (8__Main TS)	4.0382	0.6741	9.5186	0.1298
Stage Temperature (9__Main TS)	3.8555	0.6546	9.4070	0.1235
Stage Temperature (10__Main TS)	3.4784	0.6179	9.2175	0.1124
Stage Temperature (11__Main TS)	2.7296	0.5490	8.8798	0.0928
Stage Temperature (12__Main TS)	1.3746	0.4256	8.2759	0.0599
Stage Temperature (13__Main TS)	7.1488	0.5156	3.6623	0.1842
Stage Temperature (14__Main TS)	22.0331	2.7548	12.8204	0.8110
Stage Temperature (15__Main TS)	23.4619	4.0604	37.6824	1.1964
Stage Temperature (16__Main TS)	10.2299	2.2368	41.0612	0.6965
Stage Temperature (17__Main TS)	1.3656	0.4513	20.2970	0.1788
Stage Temperature (18__Main TS)	0.0011	6.7805	0.0990	1.2408

Table C.9 (Continued) IAE results of Tray Temperature for Cumene column

STREAM	qhx2	qc1	qr1	qc2
Stage Temperature (1__Main TS)	4.2717	0.7148	9.9118	0.1404
Stage Temperature (2__Main TS)	4.2495	0.7048	9.8163	0.1396
Stage Temperature (3__Main TS)	4.2346	0.7017	9.7728	0.1389
Stage Temperature (4__Main TS)	4.2190	0.6988	9.7324	0.1380
Stage Temperature (5__Main TS)	4.2002	0.6955	9.6908	0.1370
Stage Temperature (6__Main TS)	4.1733	0.6912	9.6454	0.1355
Stage Temperature (7__Main TS)	4.1272	0.6848	9.5913	0.1334
Stage Temperature (8__Main TS)	4.0382	0.6741	9.5186	0.1298
Stage Temperature (9__Main TS)	3.8555	0.6546	9.4070	0.1235
Stage Temperature (10__Main TS)	3.4784	0.6179	9.2175	0.1124
Stage Temperature (11__Main TS)	2.7296	0.5490	8.8798	0.0928
Stage Temperature (12__Main TS)	1.3746	0.4256	8.2759	0.0599
Stage Temperature (13__Main TS)	7.1488	0.5156	3.6623	0.1842
Stage Temperature (14__Main TS)	22.0331	2.7548	12.8204	0.8110
Stage Temperature (15__Main TS)	23.4619	4.0604	37.6824	1.1964
Stage Temperature (16__Main TS)	10.2299	2.2368	41.0612	0.6965
Stage Temperature (17__Main TS)	1.3656	0.4513	20.2970	0.1788
Stage Temperature (18__Main TS)	13.9865	0.9927	10.4491	64.0644

VITA

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