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PLANTWIDE CONTROL STRUCTURE DESIGN OF STYRENE PROCESS

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for the Degree of Master of Engineering Program in Chemical Engineering
Department of Chemical Engineering
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เกศสุดา มาช่วย : การออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์ของกระบวนการผลิตสไตรีน. (PLANTWIDE CONTROL STRUCTURE DESIGN OF STYRENE PROCESS) อ. ที่ปรึกษาวิทยานิพนธ์หลัก: ผศ. ดร. มนตรี วงศ์ศรี, 132 หน้า.

กระบวนการผลิตทางเคมี สิ่งสำคัญ คือการควบคุมกระบวนการเพื่อให้ระบบดำเนินงานไปตามเป้าหมายที่ต้องการ โดยให้เกิดความสูญเสียพลังงานและของเสียน้อยที่สุดโดยทั่วไป กระบวนการทางเคมีนั้นจะประกอบด้วยการหมุนเวียนมวลสารและการสะสมพลังงาน ซึ่งส่งผลทำให้ระบบการควบคุมมีความยุ่งยากมากขึ้น ดังนั้นจึงมีการนำหลักการควบคุมแบบแพลนท์ไวด์เข้ามาช่วยในการพิจารณาการควบคุมกระบวนการที่ซับซ้อน เน้นการมองภาพรวมของทั้งกระบวนการเพื่อให้ได้ระบบการควบคุมที่ดีขึ้น

ในงานวิจัยนี้จึงได้ศึกษาการออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์ตามขั้นตอนของวงศ์ศรี (2009) ประยุกต์ใช้กับกระบวนการผลิตสไตรีน โดยกระบวนการผลิตสไตรีนเป็นอุตสาหกรรมทางเคมีที่มีขนาดใหญ่ติดอันดับ 1 ใน 10 ของโลกประกอบด้วยหน่วยปฏิบัติการหลายหน่วย โดยทำการออกแบบโครงสร้างการควบคุมทั้งหมด 4 โครงสร้าง และสร้างแบบจำลองกระบวนการในสถานะคงตัวและสถานะพลวัตในโปรแกรมไฮซีส จากนั้นทำการประเมินสมรรถนะและเปรียบเทียบผลกับโครงสร้างการควบคุมอ้างอิงของลูเบน (2011) ผลที่ได้พบว่าโครงสร้างที่หนึ่งมีสมรรถนะการควบคุมดี เนื่องจากสามารถกำจัดสิ่งรบกวนที่เข้าสู่ระบบ และรักษาคุณภาพของผลิตภัณฑ์ได้ตามที่ต้องการ รวมถึงพลังงานที่ใช้ในการควบคุมน้อยกว่าโครงสร้างการควบคุมอ้างอิง โดยเปรียบเทียบจากค่าปริพันธ์ของค่าความคลาดเคลื่อนสัมบูรณ์

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The most important task in a chemical process is controlling the process to achieve the setpoint in such a way of minimizing energy loss and waste generation. In general, the chemical plant is composed of material recycle and energy integration. These affect the control system in more complexity. Plantwide control structure was adapted for developing control loops in the plant since its approach is holistic which could provide a better control system

In this work, plantwide control procedure of Wongsri (2009) is proposed and applied to the styrene process. The styrene process consists of multiple operating units. This plant is one of the top ten bulk petrochemicals in the world. Therefore, we designed four plantwide control structures (CS1 to CS4) and simulated styrene process at steady state and dynamic via HYSYS. Moreover, all designed control structures are evaluated the dynamics performance and compared with the base case control structure designed by Luyben (2011). The result shows that the designed control structure I (CS1) can handle disturbances and maintain product quality better than other control structures as compared with Integral absolute error (IAE) In addition, the energy used in all designed control structures are less than the base case control structure.

Department : Chemical Engineering Student's Signature

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NOMENCLATURES

C_p	Specific heat capacity of the stream
CS	Control Structure
CV	Controlled variable
DOF	DOF
F	Mass flow rate of the stream
HENs	Heat Exchanger Networks
IAE	Integral Absolute Error
K_u	Ultimate gain
L	Heat load
MV	Manipulated variable
P	Proportional controller
PI	Proportional -integral controller
PID	Proportional–integral–derivative controller
P_u	Ultimate period
ΔT_{\min}	Minimum temperature different

CHAPTER I

INTRODUCTIONS

This chapter introduced the importance and reasons for research, research objective, scope of research, contribution of Research, procedure and research framework.

1.1 importance and reasons for research

At present, the tendency of energy demands is rapidly increasing. With high fuel prices and the possibility of supplies in the years ahead, the development of system of using energy more effectively or energy recovery is becoming more and more important.

Moreover, industries are very competitive both in quality and cost of the production. Therefore, the 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 the required specification of the products. In the real situation, the process cannot operate smoothly; all factors do not meet the design conditions. The process always changes due to disturbance from the external factors and the internal factors, it is necessary to have the control system to control the condition and compensate any deviation occurred.

In general, most industrial processes contain a complex flow sheet with several recycle streams, energy integration, and many different unit operations. Economically, the processes can be improved by introducing recycle streams and energy integration. However, these factors lead to more interactions of the separate unit operations and create a path for disturbance propagation. These cause increases difficulties to control the process. Hence, the original process created loop control unit is unable to control the process in the desired state due to interactions of each individual unit (Umada et al., 1978).

Therefore, strategies for plantwide control (PWC) are required to operate an entire plant safely and achieve its design objectives. Essentially, the plantwide control problem is how to develop the control loops needed to operate an entire process and achieve its design objectives. The problem is extremely complex and is very much open-ended. There are a combinatorial number of possible choices and alternative strategies to control and manage the disturbance load entering the process.

Process control researchers have developed many systematic PWC methodologies and applied them to typical chemical processes. Such as the Tennessee Eastman (TE) process (Downs and Vogel, 1993), vinyl acetate monomer and toluene hydrodealkylation (HDA) plant are applied to the heuristic method ((Luyben et al., 1997). As well, the self-optimizing procedure (Skogestad, 2004) is applied to the HDA process and ammonia synthesis process. Hence, there is still a necessity to further study other complex processes in order to understand and evaluate the PWC problems.

This research will focus on new plantwide control design procedure of Wongsri (2009) applied to styrene process. Styrene is one of the top ten bulk petrochemicals in the world. More than 85% of styrene is produced by the dehydrogenation of ethyl benzene (EB). This process plant consists of many unit operations. This work is to design heat exchanger network and control structures of styrene process in order to achieve better understanding of PWC problems and evaluate PWC procedure. The performance of new designed control structures are evaluated via simulation using HYSYS.

1.2 Research Objectives

The objectives of this research are

To design and evaluate plantwide control structures of styrene process using new design procedure of Wongsri (2009).

1.3 Scope of research

The scopes of this research can be listed as follows.

1. Simulate of styrene process is HYSYS program simulator (Version 7.0).
2. Description and data of styrene process is obtained from Vasudevan et al. (2009) and Luyben (2011).
3. Plantwide control structures for styrene process are designed using new design procedure of Wongsri (2009).
4. The four control structures for the styrene process are designed.

1.4 Contribution of Research

The contributions of this research are follows;

1. Process flowsheet diagram of styrene process has been simulated.
2. The new heat exchanger network designed for styrene process.
2. The control structures are designed by using new design procedure.
3. Better understand the control structure problems and evaluate plantwide methodologies.

1.5 Research Procedures

Procedure plans of this research are;

1. Study disturbance load propagation method and plantwide process control theory.
2. Study the styrene process, and concerned information.
3. Simulate the steady state of the styrene process by using HYSYS (Version 7.0).
4. Design of disturbance load propagation for heat exchanger networks of styrene process.
5. Design new plantwide control structures followed new design procedure.
6. Simulate the dynamic of styrene process with control structures design.
7. Evaluate the dynamic performance of the designed control structures.

8. Analyze of the design and simulate result .
9. Conclude the thesis.

1.6 Research Framework

The thesis matter is classified six chapters as follow;

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

Chapter II: presents literature review related to plantwide control structures design procedures, review of previous work on the styrene process design, and design heat exchanger network.

Chapter III: covers some background information of plantwide and theory concerning with plantwide control fundamentals, new plantwide control design procedure, control issues for distillation column and energy management .

Chapter IV: describes process description and the design heat exchanger network for styrene process.

Chapter V: describes the design of plantwide control structures and dynamic simulation results and compares with control structures of Luyben (2011).

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

This is follow by:

References

Appendix A: Styrene Process Stream and Equipment.

Appendix B: Parameter Tuning of Control Structures

Appendix C: Fixture Point Theorem Data.

CHAPTER II

LITERATURE REVIEW

Our purpose of this chapter is to present a review of the previous works on the plantwide control design and heat exchanger network (HEN).

2.1 Plantwide Control Structure Design

Downs and Vogel (1993) described a model of an industrial chemical process for the purpose of developing, studying and evaluating process control technology. It consisted of a reactor, separator, and recycle management associated two simultaneous gas-liquid exothermic reactions. This process was suited for variety of studies including both plantwide and multivariable control problems.

Luyben and Tyreus (1997) proposed a heuristic procedure which composes nine steps for plantwide control structure design problem and that it considered.: energy management; product quality, production rate, environmental and safety constrain. Application of the procedure was illustrated with chemical process.

Larsson and Skogestad (2000) presented the plantwide control structure design procedure based on mathematics theory, constrained, optimization and system analysis. In particular, the selection of self- optimizing control variables those give the least loss in profit. The procedure divided in two main parts: (1) Top-down analysis, to specify degree of freedom and primary controlled variables (2) Bottom-up analysis to determine secondary controlled variables and structure of control system.

Skogestad (2004) proposed an expanded version of Larsson and Skogestad (2000). A systematic approach to plantwide control begins by defining the operational and economic objectives, and the degrees of freedom available to fulfill them. Other issues include inventory and production rate control, decentralized versus multivariable control, loss in performance by bottom-up design, and a definition of a

the “complexity number” which can proceed to find the “optimal” controller for the secondary (regulatory) control layer.

Wang and McAvoy (2001) discussed an optimization-based approach to synthesizing plantwide control architectures. The plantwide controller was synthesized in three stages involving fast and slow safety variables to be controlled, followed by product variables. In each stage, a mixed integer linear program was solved to generate candidate architectures. The objective function involved a tradeoff between manipulated variable moves and transient response area. Controlling component balances and adding unit operation controls completed the plantwide control system design.

Larsson et al., (2003) presented control structure selection for a simple plant with a liquid-phase reactor, a distillation column, and recycle of unreacted reactants. The starting point is a clear definition of the operational objectives, constraints, and degrees of freedom. Active constraints should be controller to optimize the economic performance. This implies for this case study that the reactor level should be kept at its maximum, that being economically attractive. Maximizing the reactor holdup also minimizes the “snowball effect”. The main focus is no the selection of a suitable controlled variable for the remaining unconstrained degree of freedom, that use the concept of self-optimizing control to search for a constant setpoint strategy with an acceptable economic loss. Both for the case with a given feed rate where the energy costs should be minimized and for the case where the production rate should be maximized, they find that a good controlled variable is the reflux ratio L/F . This applies to single-loop control as well as multivariable model predictive control.

Konda et al., (2005) proposed the integration framework is successfully applied to the HDA process. A viable control system can be designed by the framework which combined the advantages of both and simulation. It was shown that plantwide control system cannot be accomplished just by heuristics without the aid of rigorous nonlinear simulation tools.

Suntisrikomol (2008) suggested the “Fixture Point Theorem” for HDA process to select appropriated set of controlled variables from a large number of

candidate output as plant level variables. The fixture point control theorem states that the most disturbed points must be satisfactory controlled by giving them consideration before other controlled variables and mitigating the propagation to other units. The maximum (scaled) gain is used for selecting and pairing controlled variables with manipulated variables. The five control structures were designed and evaluated performance of designed control structures by integral absolute error (IAE) value. The designed structures are fast response and the most effective on compared with the base case.

Detjareansri (2009) used plantwide control procedure of Wongsri (2009) to develop the control structures for alkylation process. Then design eight plantwide control structures for alkylation process using new design procedure of Wongsri (2009) and evaluate the dynamic performance of the designed control structures by two types of disturbances: material and thermal disturbances. The designed control structure has a good performance because it can handle disturbances entering the process and can maintain product quality as compared by IAE and total energy use low.

Vasudevan et al. (2009) showed the comparison of the development of a plantwide control for the styrene monomer plant using the integrated framework (Konda et al., 2005). In order to measure its effectiveness, the result was compared to the heuristics procedure of Luyben and co-workers (Luyben et al., 1998) and the self-optimizing control procedure (Skogestad, 2004). An analysis of the results indicated that while all the procedures give stable control structures, the self-optimizing control procedures has performance better control structures the other procedures.

Recently Luyben (2011) interested in economic optimum design of styrene process from paper by Vasudevan et al (2009). The purpose to develop a reasonable conceptual design considering capital costs, energy costs and raw material costs. The main design optimization variables in this process are the steam-to-EB ratio, reactor inlet temperature, EB recycle flow rate, and reactor size. It is found that, higher process steam flows improve yield and selectivity but increase furnace capital and fuel costs and increase the cost of providing the process steam.

2.2 Heat Exchanger Network Design

The objectives of heat exchanger network are reaching the minimum number of matches and also the maximum energy recovery. Several methods have been performed, Temperature Interval, Evolutionary Design Methods (Linnhoff and Flower 1978b), Pinch Method (Linnhoff and Hindmarsh 1983) which utilizes design heuristics and insights derived from the previous work (Linnhoff and Flower 1978a). This method has been widely employed because it is simple and can guarantee maximum energy recovery. The problem must be firstly identified whether it is (1) a heating problem or, (2) a cooling problem or, (3) both heating and cooling problem at which the network is separated by pinch. However, it is important to note that the heat must not be allowed to transfer across the pinch. The suggested matching heuristics are start matching from the pinch, do not transfer heat across the pinch, observe the heat capacity flow rate constraints, etc.

The objectives of heat exchanger network control are reaching the target temperature and keeping the minimum utility. There are recently a few research works concerned heat exchanger network control. Marselle et al. (1982) proposed that all heat exchanger in network should be equipped with bypass and all utilities should be considerably settled with control loop.

Calandranis and Stephanopoulos (1988) proposed an approach to design the control loops for a HEN and to order the control actions of the loops in order to accommodate setpoint change and reject load disturbances. From the process design point of view, Mathisen et al. (1992) provided a heuristic method for bypass placement. The resultant HEN is supposedly satisfactory in rejecting disturbances over a moderate range of operating conditions.

Wongsri (1990) developed the heuristics and procedures for resilient heat exchanger network synthesis. The heuristics are used to develop basic and derived match patterns and disturbance propagation method. This method will transfer disturbance from one stream to another stream, which remain heat. Moreover, this algorithm can find a resiliency network structure directly from the resiliency

requirement and feature minimum number of units (MNU) and maximum energy recovery (MER)

Ploypaisangsang (2003) presented the resilient heat exchanger network design procedure provided by Wongsri (1990) is use to design resilient network for the Hydrodealkylation process (HDA Process).The match pattern heuristic, shift approach and the heat load propagation technique are essential approach. Six alternatives for the HDA process are redesign to be the resiliency networks for maintaining target temperature and reaching maximum energy recovery (MER).

Gayapan (2008) studied nine alternatives (eight new alternative designs and base case) of heat exchanger networks and three control structure designs of the natural gas expander plant are proposed. The resilient heat exchanger networks are designed using the disturbance load propagation method (Wongsri, 1990) and the control structures of HENs using the heat pathway heuristics (Hermawan, 2004). The plantwide control structures are designed by using Luyben heuristic design method (1999) for the CS1 control structure and Fixture point theorem (Wongsri, 2008) for the CS2, CS3 control structure. The result shows the CS3 control structure can handle disturbances better than other control structures.

Sukkongwaree (2009) also pointed out that the procedure of Wongsri (2009) not only provides the appropriated set of controlled variables but also consider the energy management via heat exchanger networks. The result of the designed control structures showed an excellent performance of control structures through highly complexities with heat-integration recycle plants.

CHAPTER III

THEORY

This chapter is aimed to summarize heuristic approach from the previous researches and this approach heat exchanger network design. 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 Pinch Technology

Pinch technology has been developed for more than two decades and now provides a systematic methodology for analysis chemical processes and surrounding utility systems. The concept was first developed by two independent research groups 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, i.e. reaction and separation system, has been designed. 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 flow rate for each stream as shown in Table 3.1.

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

Stream No.	Stream type	Start Temperature (Ts), °C	Target Temperature (Tt), °C	Heat capacity flow rate (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 streams that required cooling, i.e. the source temperature is higher than that of the target. While the cold streams are referred to those required heating, i.e. the target temperature is higher than the supply. Heat Capacity flow rate is defined as the multiple between specific heat capacity and mass flow rate as shown below.

$$CP = C_p * F \quad (3.1)$$

Where CP = heat capacity flow rate (kW/°C)

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

F = mass flow rate of the stream (kg/s)

The data used here is based on the assumption that the heat capacity flow rate is constant. In practice, this assumption is valid because every streams 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 $\Delta T_{min} = 20^\circ\text{C}$, the results obtained from this method are shown in Table 3.2.

Table 3.2 The problem table for data given in Table 3.1

W				T hot (°C)	T cold (°C)	ΣW (kW/C)	ΔT (°C)	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	107.5	10	2.5	10
2	0	2.5	0	120	100	-0.5	25	117.5	-12.5	12.5	-2.5
2	0	2.5	3	90	70	-3.5	30	105	-105	0	-107.5
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	-52.5	-67.5
										Qc	

The pinch separates the problem into two 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.

3.2 Heat Exchanger network

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*.

3.2.1 Heuristics

The heuristics approach is based on the use of rules of thumb to provide a plausible direction in the solution of the problem. There are a number of design procedures using heuristic in structuring an optimal network featuring minimum number of matches and maximum energy recovery, however, there are to be the best of our knowledge that use heuristics to structure a resilient network.

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 C.1 To generate a heat exchanger network featuring the minimum number of heat transfer units, let is match eliminate at least one of the two streams – a „tick-off“ rule

Heuristic C.2 Prefer the matches that will leave a residual stream at its cold end if a problem is a heating problem and at its hot end if a problem is a cooling problem. Obviously, a match of this type will feature the maximum temperature difference.¹

Heuristic C.3 Prefer matching large heat load streams together. The significance of this rule is that the control problem (a capital cost) of a match 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 E.1 Divide the problem at the pinch into subproblems and solve them separately (Linnhoff and Hindmarsh, 1983). This is followed by the next three heuristics.

Heuristic E.2 Do not transfer heat across the pinch.

Heuristic E.3 Do not cool above the pinch.

Heuristic E.4 Do not heat below the pinch.

The laws of thermodynamics:

Rule T.1 In a heating problem, if a supply temperature of a cold stream is less than a target temperature of a hot stream by the minimum approach temperature (ΔT_{\min}) or more and the heat capacity flowrate of a hot stream is less than or equal to the heat capacity of flowrate of cold stream. The match between these two streams is feasible. (Immediately above the pinch temperature, the heat capacity flow rate of a cold stream must be greater than or equal to that of a hot stream.)

Rule T.2 In a cooling problem, if a supply temperature of a hot stream is greater than a target temperature of a cold stream by minimum approach temperature, ΔT_{\min} ,

or more and the heat capacity of flowrate of a cold stream, the match between these two streams is certainly feasible. (Immediately below the pinch temperature, the heat capacity flow rate of the hot stream must be greater than or equal to that of a cold stream.)

Rule T.1 and T.2 can be used as a quick checks in match feasibility tests.

Rule T.3 For a situation different from the above rules, match feasibility must be determined by checking whether the minimum temperature difference of a match violates the minimum approach temperature, ΔT_{\min} , specified by the design.

The heuristics that concern heat load state that one must match a large heat load hot and cold streams first. However, we want to propose two heuristics:

Heuristic N.1 We propose that for a heating subproblem, a match where the heat load of a cold stream is greater than of a hot stream should be given higher priority than the other. The reason is that the net heat load in a heating subproblem is a deficit. The sum of heat loads of cold streams is greater than that of hot streams. The proposed match will likely be present in a solution.

Heuristic N.2 Conversely, we prefer a match where the heat load of a hot stream is greater than that of a cold stream in a cooling subproblem.

3.2.2 Disturbance Propagation Design method.

In order for a stream to be resilient with a specified disturbance load must be transferred to heat sink or heat sources with the network. With the use of heuristic: To generate a heat exchanger network the featuring the minimum number of heat transfer units, let each match eliminate at least one of the two streams.

We can see that in a the match of two heat load variable stream, the variation in heat load of the smaller stream S1 will cause a variation to the residual of the larger stream S2 by the same degree: in effect the disturbance load of S1 is shifted to the residual of S2. If the residual stream S2 is matched to S3, which has lager heat load,

the same situation will happen. The combined disturbance load of S1 and S2 will cause the variation in the heat load to the residual S3. Hence, it is easy to see that the disturbance load in residual S3 is the combination of its own disturbance load and those obtained from S1 and S2. Or, if S2 is matched to a smaller heat load stream S4, the new disturbance load of residual S2 will be the sum disturbance load of S1 and S4. From this observation, in order to be resilient, a smaller process stream with specified disturbance load must be matched to a larger stream that can tolerate its disturbance. In other words, the propagated disturbance will not overshoot the target temperature of the larger process stream.

However, the amount of disturbance load can be shifted from one stream to another depends upon the type of match patterns and residual heat load. Hence, in design we must choose a pattern that yields the maximum resiliency. We can state that the resiliency requirement for a match pattern selection is that the entire disturbance load from smaller heat load stream must be tolerated by residual stream. Otherwise, the target temperature of the smaller stream will fluctuate by the unshifted disturbance. Of course, utility exchanger will finally handle the propagated disturbance. In short, the minimum heat load value of a larger stream must be less than a maximum heat load value of a smaller stream

By choosing the minimum heat load condition for the design, the new input temperature of a residual stream to its design condition according to the propagated disturbance. The propagated disturbance will proportionally cause more temperature variation in the residual stream and the range of temperature variation of the residual stream will be larger than its original range.

The propagated disturbance of a stream is disturbance caused by a variation in heat load of up-path stream to which such a stream is matched. Only a residual stream will have a propagated disturbance. The new disturbance load of a residual stream will be the sum of its own disturbance (if any) and the propagated disturbance. See Figure 3.1 and 3.2

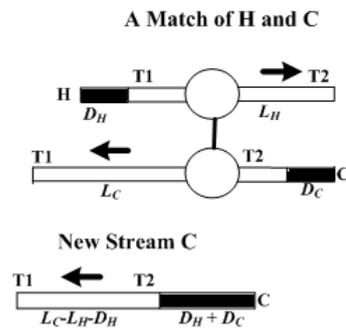


Figure 3.1 A Concept of Propagated Disturbance

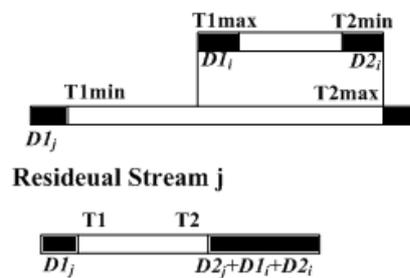


Figure 3.2 A General Concept of Propagated Disturbance

Hence, a stream with no original variation in heat load will be subjected to variation in heat load if it is matched to a stream with disturbance. Another design consideration is that the disturbance load travel path should be as short as possible, i.e. the least number of streams involved. Otherwise, the accumulated disturbance will be at high level. From the control point of view, it is difficult to achieve good control if the order of the process and the transportation lag are high. From the design viewpoint, one may not find heat sinks or sources that can handle the large amount of propagated disturbance.

3.3 Match Pattern.

A heuristic approach to design or synthesize a resilient HEN has been presented by Wongsri (1990). A resilient network is defined as a network that provides a down path for variable process streams so that their specified input heat load disturbances can be shifted to the heaters or coolers in their network without violation in the specified target temperatures and MER. HEN synthesis is usually considered as a combinatorial matching problem. Match patterns are the descriptions of the match configuration of two, and possibly more, process streams and their properties that are thermally connected with the heat exchangers.

3.3.1. Classe of Match Patterns.

There are four match patterns for a pair of hot and cold streams according to the match position and the length (heat load) of stream. The four match patterns are considered to the basic match pattern classes and simply called A, B, C, and D as shown in Figures 3.3 to 3.6. Any eligible match must belong to one of the four match pattern classes.

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 (Figure 3.3).

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 (Figure 3.4).

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 (Figure 3.5).

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 (Figure 3.6).

When the residual heat load in a match pattern is matched to a utility stream, it is a 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 requirements then the chances that the match pattern will be matched to a utility stream is high. Therefore, we give a match pattern which residual less than the minimum heating or cooling requirement a high priority in match pattern

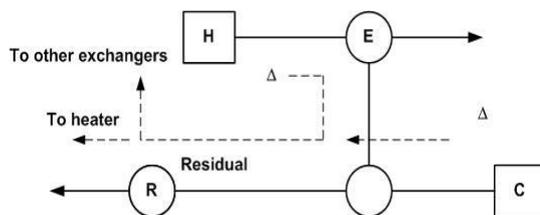


Figure 3.3 Class A Match Pattern.

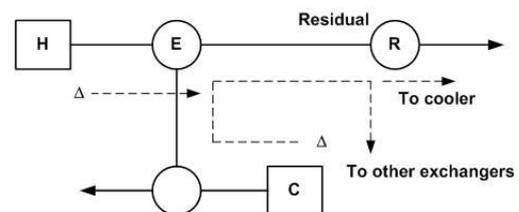


Figure 3.4 Class B Match Pattern.

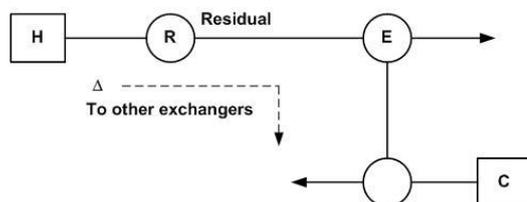


Figure 3.5 Class C Match Pattern.

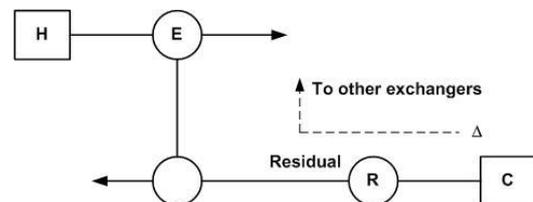


Figure 3.6 Class D Match Pattern

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 Heuristic N.3 and N.4 to further subclassify matches of class A and B and give the following subclass match 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.)

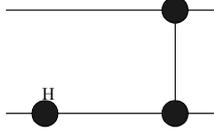
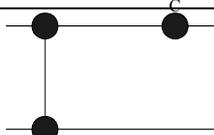
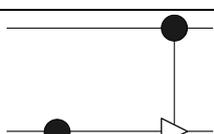
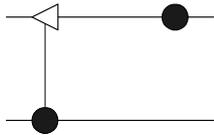
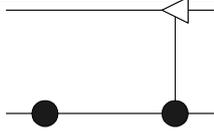
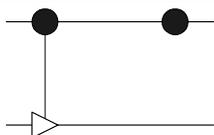
Subclass BK. A match of this subclass is a member of Class B, a cooling problem where the residual is less than or equal to the minimum cooling requirement. (A letter K in the subclass name denotes that the residual is matched to a cooling utility).

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 3.3

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 math in which the heat capacity flowrate of the cold stream is greater than or equal to that of the hot stream. Similarly in 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 math 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 heat problem are AH, A[H], B[C], A[C], B[H],C[H], D[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.

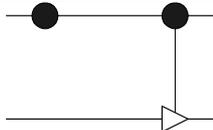
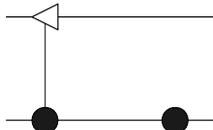
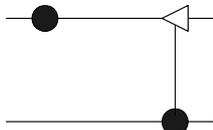
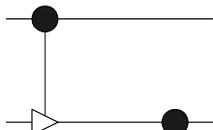
Match Operators	Conditions	Actions
 <p>Pattern AH</p>	$T_H^s \geq T_C^{t**}$ $L_H \leq L_C$ $T_H^s \geq T_C^s + L_H W_C^{-1}$ $L_C - L_H \leq Q_{\min}^{heating}$	<p>Match H and C</p> <p>Status of H \Leftarrow Matched***</p> $T_C^s \Leftarrow T_C^s + L_H W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 <p>Pattern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

* T^t =target temp, T^s =supply temp, W =heat capacity flowrate, L , Q =heat load.

** Cold stream temperatures are shifted up by ΔT_{\min} .

*** 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
 <p>Pattern C[H]</p>	$T_H^t \geq T_C^s$ $L_H > L_C$ $W_H \leq W_C$	Match H and C Status of C \Leftarrow Matched $T_H^t \Leftarrow T_H^t - L_C W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 <p>Pattern D[C]</p>	$T_H^s \geq T_C^t$ $L_H < L_C$ $W_H \geq W_C$	Match H and C Status of H \Leftarrow Matched $T_C^t \Leftarrow T_C^t + L_H W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 <p>Pattern C[C]</p>	$T_H^t \geq T_C^s$ $L_H > L_C$ $W_C < W_H$ $T_C^t \leq T_H^t + L_C W_H^{-1}$	Match H and C Status of C \Leftarrow Matched $T_H^t \Leftarrow T_H^t - L_C W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 <p>Pattern D[H]</p>	$T_H^s \geq T_C^t$ $L_H \leq L_C$ $W_H < W_C$ $T_H^t \geq T_C^t - L_H W_C^{-1}$	Match H and C Status of H \Leftarrow Matched $T_C^t \Leftarrow T_C^t + L_H W_C^{-1}$ $L_C \Leftarrow L_C - L_H$

* T^t =target temp, T^s =supply temp, W =heat capacity flowrate, L , Q =heat load.

** Cold stream temperatures are shifted up by ΔT_{\min} .

*** 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.

greater than that of the hot stream and the residual of the cold stream is matched to the heating utility.

3.4 Plantwide Control

A typical chemical plant flowsheet has a mixture of multiple units connected both in series and parallel that consists of reaction sections, separation sections and heat exchanger network. So plantwide process control involves the system and strategies required to control entire plant consisting of many interconnected unit operations

3.5 Integrated Process

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

- 1) The effect of material recycles.
- 2) The effect of energy integration.
- 3) The need to account for chemical component inventories.

If they did not have to worry about these issues, then they 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.

3.5.1 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. Therefore the reactor effluent by necessity contains both reactants and products. Separation and recycle of reactants are essential if the process is to be economically viable.

- 2) Improve economics.

In most systems it is simply cheaper to build a reactor with incomplete conversion and recycle reactants than it is to reach the necessary conversion level in one reactor or several in series. 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 $A \rightarrow B \rightarrow C$, 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 (an excess of one reactant or a product) so that the reactor temperature increase will not be too large. High temperature can potentially create several unpleasant events: it can

lead to thermal runaways, it can deactivate catalysts, it can cause undesirable side reactions, it can cause mechanical failure of equipment, etc. So the heat of reaction is absorbed by the sensible heat required to raise the temperature of the excess material in the stream flowing through the reactor.

5) Prevent side reactions.

A large excess of one of the reactants 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. These include average molecular weight, molecular weight distribution, degree of branching, particle size etc. 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.

3.6.2 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.

3.6.3 Chemical component inventories.

In chemical process can characterize a plant's chemical species into three types: reactants, products, and inert. The real problem usually arises when we consider reactants (because of recycle) and account for their inventories within the entire process. Every molecule of reactants fed into the plant must either be consumed or leave as impurity or purge. Because of their value so they prevent reactants from leaving. This means we must ensure that every mole of reactant fed to the process is consumed by the reactions.

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

3.6 Basic Concepts of Plantwide Control

Buckley basics

Buckley (1964) was the first to suggest the idea of separating the plantwide control problem into two parts: material balance control and product quality control. He suggested looking first at the flow of material through the system. A logical arrangement of level and pressure control loop is established, using the flowrates of the liquid and gas process streams. No controller tuning or inventory sizing is done at this step. The idea is to establish the inventory control system by setting up this “hydraulic” control structure as the first step. Then he proposed to establishing the product-quality control loops by choosing appropriate manipulated variables. The time constants of the closed-loop product-quality loops are estimated. They try to make these as small as possible so that good, tight control is achieved, but stability constraints impose limitations on the achievable performance

Douglas doctrines

Douglas (1988) 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. He 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 they need tight control of stream compositions exiting the process to avoid losses of reactants and products. The second rests on the principle that yield is worth more than energy.

Downs drill

Dows (1992) pointed out the importance of balancing the chemical component around the chemical plant and checking to see that the control structure an effective handles these components. The concepts of overall component balances go back to our first course in chemical engineering, where they learned how to apply mass and energy balances to system, microscopic or macroscopic. They did these balances for individual unit operations, for section of a plant, and for entire processes. He must ensure that all components (reactants, products, and inerts) have a way to leave or be consumed in the operations.

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 reactions types such as $A+B \rightarrow \text{product}$. 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 shift in the two products (M and D) that are generated. An excess of A will result in the production of more M and less D . An excess of B results in the production of more D and less M .

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 (Cantrell et al., 1995). Changes in

feed flowrate or feed composition have less of a dynamic effect on distillate composition than they do on bottoms composition if the feed is saturated liquid. The reverse is true if the feed is saturated vapor: bottom is less affected than distillate. If our primary goal is to achieve tight product quality control, the basic column design should consider the dynamic implications of feed thermal conditions.

Richardson rule

Richardson (1995) suggested the heuristic that the largest stream should 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. An analogy is that it is much easier to maneuver a large barge with a tugboat than with a life raft. The point is that the bigger the handle you have you have to affect a process, the better you can control it.

Tyreus tuning

Tyreus and Luyben (1997) suggested 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. They strongly recommend the use of P-only controllers for liquid levels (even in some liquid reactor applications) and the use PI controller for other control loops. 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:

$$K_{ZN} = K_u/2.2 \quad \tau_{ZN} = P_u/1.2$$

$$K_{TL} = K_u/3.2 \quad \tau_{TL} = 2.2P_u$$

3.7 New Plantwide Control Structure Design Procedure

New design procedure of Wongsri (2009) presented plantwide control structure design procedure based on heuristics 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 proposed plantwide control structure design procedure for selection of 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 sum of its parts. These 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:

Step 1: Establishment of control objective.

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

Step 3: Selection of manipulated variables and measurements via DOF analysis.

Step 4: Energy management via heat exchanger networks.

Step 5: Selection of control configuration using various tools available.

Step 6: Completing control structure design by checking the component balance.

Step 7: Selection of controller type: single loop or multiloop.

Step 8: Validation via rigorous dynamic simulation.

3.7.1 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).

CHAPTER IV

STYRENE PROCESS

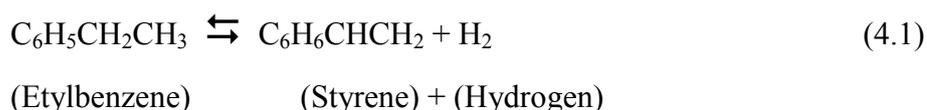
4.1 Introduction

Styrene is one of the most important aromatic monomer in the world used for the manufacture of plastics. Then 65% of styrene is used to produce polystyrene (Woodle, 2006). In styrene process contain multi-unit operations: furnace, heat exchanger, heater, cooler, two adiabatic plug flow reactors in series, three-phases separator and two distillation columns. Two raw materials, ethylbenzene (EB) and low-pressure steam (LPs) converted into the styrene product, with the most hydrogen as by-product.

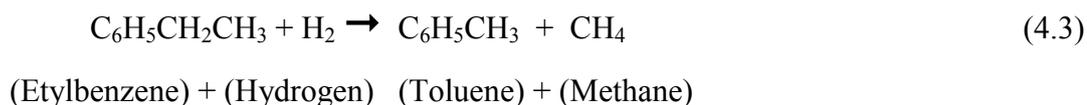
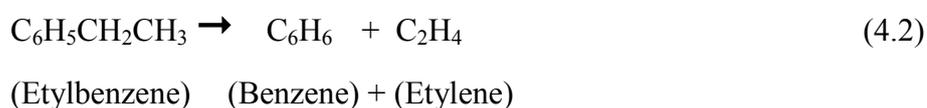
4.2 Reaction Kinetics.

Styrene is usually produced from dehydrogenation of ethylbenzene in highly endothermic vapor-phase reaction, which required high temperature and low pressure. The main reaction is reversible.

Styrene production



There are several other side reactions such as the dealkylations from ethylbenzene to benzene and ethylene or toluene and methane.



Both methane and ethylene pass steam-reforming reactions according to the following equations:



(Water) + (Etylene) (Caponmonoxide) + (Hydrogen)

The water-gas shift reaction occurs together and is commonly near equilibrium at the reaction temperature.



(Water) + (Etylene) (Caponmonoxide) + (Hydrogen)



(Water) + (Etylene) (Caponmonoxide) + (Hydrogen)

Model formulation this research, kinetic parameters for styrene following as Table 4.1.

Table 4.1: Reaction Kinetics. (Luyben, 2011)

reactions	<i>k</i>	E (kJ/kmol)	concentration
(1) forward	0.044	90,981	P_{EB}
(1) reverse	6×10^{-8}	61,127	$P_{\text{S}}P_{\text{H}}$
(2)	27,100	207,989	P_{EB}
(3)	6.484×10^{-7}	91,515	$P_{\text{EB}}P_{\text{H}}$
(4)	4.487×10^{-7}	103,997	$(P_{\text{W}})^2P_{\text{E}}$
(5)	2.56410^{-6}	62,723	$P_{\text{W}}P_{\text{M}}$
(6)	1,779	73,638	$P_{\text{W}}P_{\text{CO}}$

Overall reaction rates have units of $\text{kmol s}^{-1} \text{m}^{-3}$ and concentration units are partial pressure is Pascals.

4.3 Process Description.

In the styrene process, fresh EB and a part of the low-pressure steam (LPs) are initially mixed to reduced partial pressure and then preheated in a feed effluent heat exchanger (FEHE) using the reactant effluent stream. Before entering the reactor both streams are initially mixed. The remaining LP stream is superheated in a furnace and mixed with the preheated mixture. Then, mixture fed to adiabatic PFRs in series with a heater in between, for the production of styrene. That generates high-temperature and low-pressure steam from the endothermic reactions.

The reactor effluent is cooled in the FEHE and further cooled in a cooler to remain at 40 °C and pressure at 120 kPa before sent to the three-phase separator, where the light gases are removed as the light product and water is removed as the heavy product. The intermediate organic layer is sent to a set of distillation columns for styrene separation from the other components.

In the product column, the product column is operating under vacuum to prevent styrene polymerization; styrene is removed as the bottom product, and the distillate D1 is ethyl benzene and light components are separated. This produces a vapor product of mostly light components that are removed through a compressor and a liquid distillate that is sent to the recycle column.

The liquid distillate from the product column D1 is the feed to the recycle column. This column removes the light components (mostly benzene and toluene) in the distillate and recovers the ethyl benzene in the bottoms for recycle back to the reaction section.

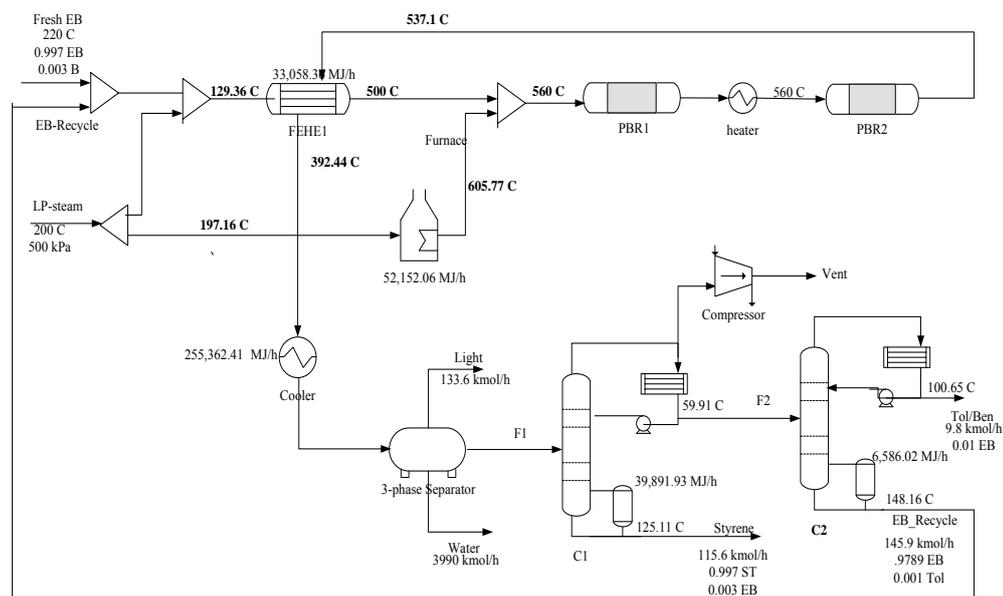


Figure 4.1 Flowsheet for styrene process (base case).

4.4 Design of Heat Exchanger Networks

At this point, the heat exchanger networks design method provide by Wongsri (1990) for styrene process. The design procedures and definitions from previous chapters will be methods to design and compare with the preliminary stage of a

process design without energy integration. The information for design is shown in the following table 4.2.

Table 4.2 stream data conditions for styrene process.

Stream Name	Tin (°C)	Tout (°C)	W (MJ/h°C)	Q duty(MJ/h)
H1:PFR 2 outlet to E-2	537.06	40.00	580.28	288433.41
C1:mixed stream to FEHE E2	129.36	500.00	89.23	33072.04
C2:steam to Furnace E1	197.16	605.77	127.63	52151.08
C3: To Reboiler @ C1	125.10	125.11	4,051,565.46	39,891.93
C4: To Reboiler @C2	147.98	148.16	36,485.70	6,586.02

4.4.1 HEN Base Case.

According to table 4.2, it can be simply translated to a heat exchanger network for styrene process (Base Case) in Figure 4.2.

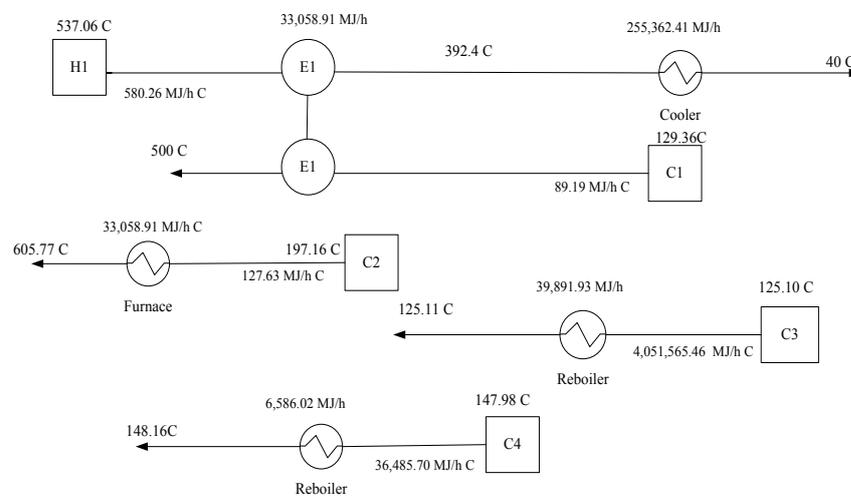


Figure 4.2 The Heat exchanger network, Base Case for styrene process.

4.4.2. HEN using the disturbance propagation method.

There are one hot and two cold streams in the system. Therefore, we can find Pinch temperature using the Problem table method occurs at 537.06/527.06 °C. The synthesis procedure using the disturbance propagation method and math pattern are shows in the Table 4.3. Figure 4.3 shows a design of heat exchanger network and

Figure 4.4 shows flowsheet for styrene processes. In this design, the minimum temperature difference in the process- to- process-heat exchanger is set 10 °C.

Figure 4.3 The Heat exchanger network, HEN for styrene process.

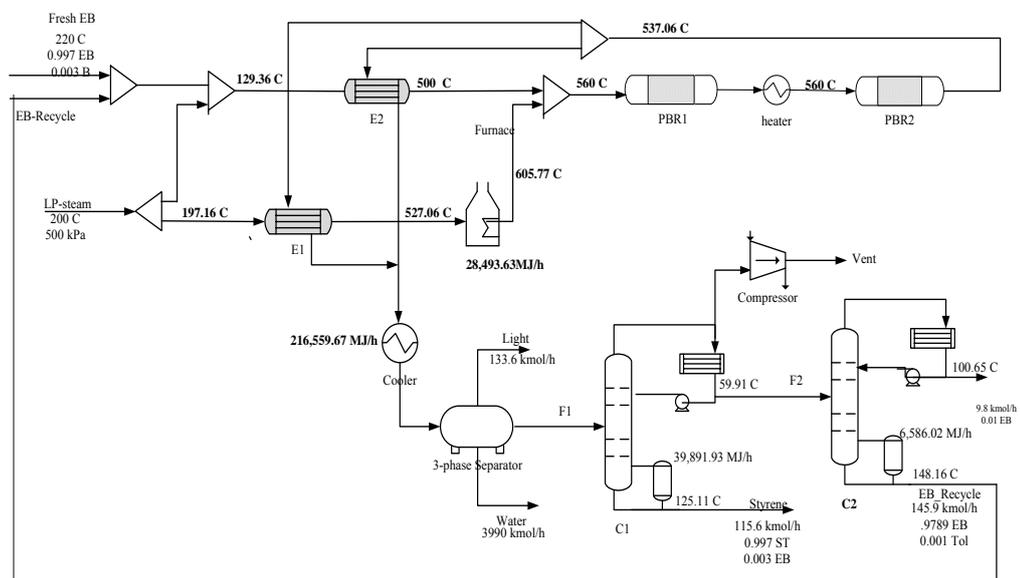


Figure 4.4 Flowsheet for Heat exchanger network for styrene process.

Table 4.3 Synthesis table for HEN.

Hot end synthesis table							
Stream	Load	W	T1	T2	D1	D2	Action
a) State 1							
H1	0.00	580.28	537.06	537.06	5802.84	5802.84	Selected A[C]
C2	10045.57	127.63	527.06	605.77	1276.31		Selected
b) State 2							
H1							Matched To C2
C2	10,045.57	127.63	527.06	605.77	1276.31	0.00	To Heater
Cold end synthesis table							
a) State 1							
H1	288433.41	580.28	537.06	40.00	5802.84	0.00	Split
H11	162,373.59	326.67	537.06	40.00			
H12	126,059.82	253.61	537.06	40.00			
C1	30406.38	83.16	134.36	500.00	831.59	0.00	Select
C2	41,467.35	127.63	202.16	527.06	1276.31	1276.31	Select
b) State 2							
H1	216559.67	580.28	333.20	40.00	0.00	0.00	Cooler
H11	120,906.23	326.67	224.48	40.00			
H12	95,653.44	253.61	195.09	40.00			
C1							H11
C2							H12

Table 4.4 Utility costs. (Turtun R el at, 2007)

Utility	Description	Cost (\$/GJ)
Steam from boilers	Process steam: latent heat only	
	a. Low pressure (5 barg, 160 °C) from HP steam	6.08
	b. Medium pressure (10 barg, 184 °C) from HP steam	6.87
	c. High pressure (41 barg, 254 °C) from HP steam	9.83
	Cooling water	0.345
	Electric Distribution	
	a. 110 V	16.8
b. 220 V		
c. 440 V		

4.5 Evaluation of cost.

This section discusses the economic cost comparison of styrene process with the base case control structure and designed control structures. The expenses associated with capital costs (furnace, reactors, heat exchangers, cooler, decanter, and distillation columns) and energy costs, which consists of furnace fuel, cooling water, and column reboiler energy. The costs associated with supplying a given utility are then obtained by calculating the operating costs to generate the utility. These costs have been presented in Table 4.4.

The increase in heat exchanger capital investment for the designed control structure is 458.66 K\$. In addition, the total capital costs decreased from 8,679.48 K\$ to 8,567.58 K\$ and the total utility costs decreased from 8,642.87 K\$ to 5,190.52 K\$ per year when compare with the base case control structure.

The economic evaluation is based on total annual costs (TAC), which assume the operating day as 340 day per year and payback period equal to 3 years. The result total annual costs of the base case and design control structures are 11,536.03 K\$ and 8,046.37 K\$, respectively show in Table 4.5.

Table 4.5 Economic costs for styrene process.

1. Utilities usage (GJ/h)	Base case	New designed
furnace	52.15	10.56
heater	8.66	8.66
cooler	255.46	213.77
Reboiler	45.89	45.89
condenser	40.69	40.69
compressor	0.05	0.05
Total hot utilities	106.75	77.30
Total cold utilities	296.16	269.32
2. Utilities costs (K\$/year)	8,642.87	5,190.52
Cost saving (%)		39.94
3. Total capital cost (K\$)	8,679.48	8,567.58
Investment (%)		-1.29
4. TAC	11,536.03	8,046.37
TAC saving (%)		30.25

4.6 Steady state simulation

The steady state model is built in HYSYS. The flowsheet information is obtained from Luyben (2011). It is highly important to select the most suitable fluid package for realistic simulation. The Peng-Robinson equation of state is chosen, as it is very reliable for predicting the properties of hydrocarbon components over a wide range of conditions and is appropriate for the components in the styrene production process.

All of two columns are simulated using the distillation column module. In steady state, the specification of inlet stream, pressure profiles, number of trays and feed tray need for model. Besides, two additional variables should be additionally specified for column with condenser or reboiler. These could be duties, reflux rate, draw stream rates, composition fraction, etc. two compositions of light key ($x_{B(LK)}$) and heavy key ($x_{D(HK)}$) is specified at the columns. Figure 4.5 presents the steady state flowsheet with heat exchanger network built in HYSYS.

CHAPTER V

CONTROL STRUCTURES DESIGN

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 styrene process. Heat exchanger networks (HEN) are designed for styrene and compared between base case (Luyben, 2011) on rigorous dynamic simulation via HYSYS.

5.1 New Plantwide Control Strategies

In this research, the plantwide control structure design procedures in the styrene 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

The steady-state design of the process is assessed and the dynamic control objectives are formulated. The objectives typically include reaction and separation yields, product quality specifications, environmental restrictions, and safety concerns. The control objectives for the styrene process are as follows:

Plantwide level:

1. Product quality: styrene composition at 99.7 mol%
2. Production capacity: more than 110 kmol/h of styrene

Unit level: stabilization and smooth operations.

Process constrains during operation:

1. The split ratio of the fresh steam is 0.18. This means 18% of the fresh steam is mixed with the total EB (feed plus recycle), while the rest is sent to the furnace for superheating. (Vasudevan et al., 2009)

2. The ratio of steam to EB molar flow rates entering the first reactor is set between 14 to 15.
3. The temperature at both the reactor inlets should be 560 °C to minimize the side reactions.
4. The temperature of the three-phase separator should be at or below 40 °C in order to maximize organic product recovery.
5. The first distillation column should operate under vacuum due to the risk of styrene polymerization in the column bottom section. Therefore, the column operating pressure is set at 10 kPa. (Lee and Hubbell, 1982)

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

Plantwide Level: Consider material recycle loop because it cause a system to be born “snowball effect”. Styrene process has a large recycle stream that is EB. Therefore, to avoid snowball effect, EB fresh feed valve is manipulated to control total EB molar flow rate (fresh feed + recycle flow) in the process.

Unit Level: The Fixture Point theorem is used to rank the selected CVs to maintain product quality, to satisfy safety operational and to environmental constraints. 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 IAE summation (this value is the controlled variables deviation from set point value, when it is tested disturbances) results of feed streams from Fixture Point method to select the appropriate controlled variables to setting the production rate from a candidate output deviation. From this table, total EB molar flow rate and LP steam molar flow are selected as controlled variables because they are more sensitive than the others, and Table 5.2 shows the ranked CVs and the IAE summation results.

Table 5.1 IAE summation results of feed stream.

Rank	Variables		SUM IAE
1	Total-EB	Molar flow rate	1.5402
2	LP-steam	Molar flow rate	1.7595
3	Total-EB	Temperature	0.2268
4	LP-steam	Temperature	0.0000
6	Total-EB	Pressure	0.4792
7	LP-steam	Pressure	0.0000

Table 5.2 The ranked CVs and the IAE summation results.

Rank	Variables		SUM IAE
1	3-phase separator	%level	10.7134
2	Reboiler of product column	%level	9.9573
3	Condenser of recycle column	Pressure	9.2424
4	3-phase separator	temperature	7.7867
5	R1_inlet stream	temperature	7.7973
6	L1 stream of product column	temperature	7.3706
7	Reboiler of recycle column	%level	6.7595
8	R2_inlet stream	temperature	4.5403
9	Condenser of product column	%level	3.1570
10	Condenser of recycle column	%level	3.0037
11	EB recycle	molar flow rate	2.6019
12	Total EB stream	molar flow rate	2.4813
13	3-phase separator	pressure	1.87369
14	LPS2_stream	molar flow rate	1.4772
15	Condenser of product column	Pressure	0.7733
16	LPS1_stream	molar flow rate	0.4809
17	Bottom of recycle column	Toluene composition	0.3387
18	Distillate of recycle column	pressure	0.2446
19	Bottom of recycle column	EB composition	0.2402
20	3-phase separator	% aqueous level	0.0831

The product column operates under vacuum to prevent styrene polymerization. There is no temperature control because the boiling points of the key components are

so close. Temperatures are affected more by changes in pressure than by changes in compositions. Therefore, a composition measurement is required.

Table 5.3 shows IAE summation result of tray EB composition for product column. Found that, tray 54th has maximum deviation is selected to control the operation of product column. This selected tray is the same tray that changes significantly from tray to tray as Figure 5.1 show that tray 54th give quite high gradient too.

Table 5.3 IAE summation result of tray EB composition deviation for product column.

Tray	Sum IAE						
1	1.2181	21	5.7253	41	5.0354	61	10.6738
2	1.4235	22	5.5724	42	5.5905	62	9.6188
3	1.9849	23	5.3969	43	6.2700	63	8.5762
4	2.4150	24	5.2094	44	7.0454	64	7.5744
5	2.7786	25	5.0192	45	7.9099	65	6.6327
6	3.1226	26	4.8373	46	8.8473	66	5.7628
7	3.4796	27	4.6718	47	9.8341	67	4.9762
8	3.8389	28	4.5319	48	10.8409	68	4.2736
9	4.1879	29	4.4239	49	11.8291	69	3.6523
10	4.5228	30	4.3499	50	12.7552	70	3.1096
11	4.8375	31	4.3427	51	13.5724	71	2.6340
12	5.1246	32	4.4084	52	14.2350	72	2.2216
13	5.3782	33	4.5424	53	14.7025	73	1.8643
14	5.5919	34	4.7491	54	14.9424	74	1.5616
15	5.7604	35	4.2489	55	14.9363	75	1.3022
16	5.8795	36	4.0418	56	14.6836	76	1.0794
17	5.9469	37	3.9726	57	14.2000	77	0.8872
18	5.9622	38	4.0474	58	13.5144	78	0.7247
19	5.9274	39	4.2754	59	12.6674	79	0.5858
20	5.8464	40	4.6047	60	11.7050	80	0.4661

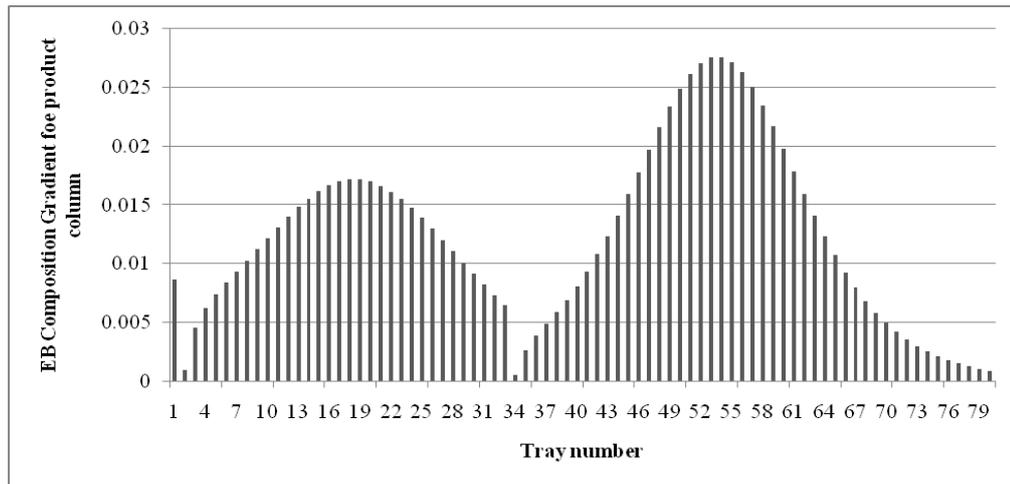


Figure 5.1 EB Composition Gradient of product column.

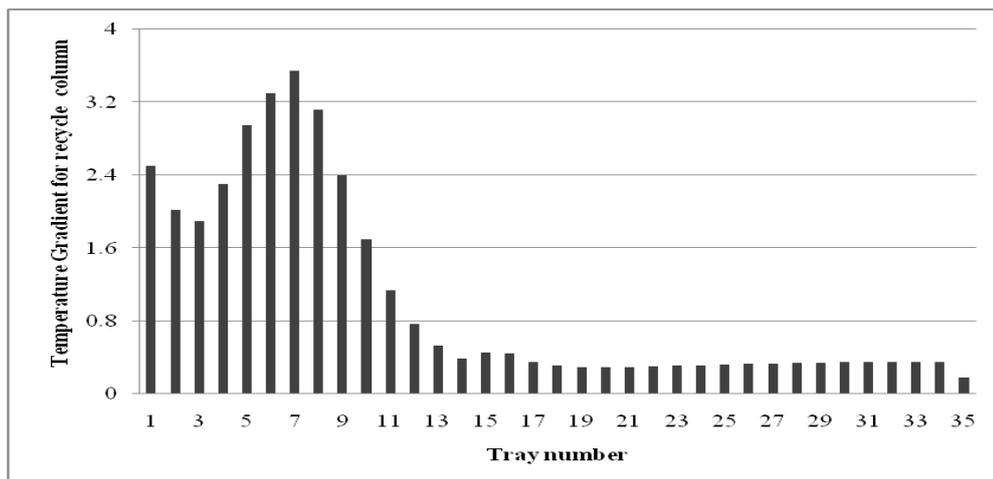


Figure 5.2 Temperature Gradient of recycle column

The slope value from temperature profile in the recycle column is shown in Figure 5.2. There is little temperature change from tray to tray in the stripping section but there is a section in the rectifying section that can be used for temperature control. This column is controlled with a dual structure. Due to one temperature can be controlled, the other controller must control a composition. The results in significant temperature profile are affected on composition compared with the IAE summation result of toluene composition for recycle column from the Fixture Point method in Table 5.4. It is found that, tray 7th has maximum deviation in the stripping section to control temperature and tray 28th has maximum deviation is selected to control the toluene composition in

the stripping section, it should do an effective job of maintaining the toluene impurity in bottom of recycle column.

Table 5.4 IAE summation result of tray temperature deviation for recycle column.

Tray	sum IAE	Tray	sum IAE	Tray	sum IAE
1	1.3401	13	3.6157	25	3.0930
2	1.5061	14	3.3852	26	3.2450
3	2.8448	15	3.2676	27	3.3431
4	4.6086	16	2.7812	28	3.3777
5	6.6592	17	2.4686	29	3.3397
6	8.4284	18	2.1699	30	3.2270
7	9.1522	19	1.8817	31	3.0387
8	8.6215	20	1.8351	32	2.7762
9	7.3539	21	2.0853	33	2.4415
10	5.9756	22	2.3839	34	2.0378
11	4.8500	23	2.6554	35	1.5738
12	4.0614	24	2.8941		

Step 3: Selection of manipulates variables and measurements via degree of freedom (DOF) analysis.

The restraining number method is used to determine the DOF of this process. All the streams (including energy and utility streams) are numbered in the styrene process. The DOF of the styrene process is 22.

Step 4: Energy management via heat exchanger network

The energy management via heat exchanger network is described in Chapter IV of design of heat exchanger network section.

Step 5: Selection of control configuration using various tools available.

Selection of control configuration use process 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 remainder of the plant.

For dehydrogenation reaction: EB concentration, temperature and pressure. There are parameters to influence the reactor productivity. Therefore, the total EB flow (fresh feed + recycle) is used for setting the production rate. The most suitable manipulator to control total EB flow is the EB fresh feed flow (V-1).

The inlet temperature of both the reactors needs to be controlled at the desired values. While the manipulator that most appropriate the first and second reactors, furnace, and intermediate heater duty respectively.

From process constraint, two ratio controllers are installed to maintain the split steam to EB ratio by manipulating split flow (V-3) and steam flow feeds to furnace (V-4) is manipulated to control the rest steam to EB ratio.

The reflux- to- feed ratio (R/F) controller is used to increase the reflux flowrate in product column. The feed flowrate is measured, and the signal is sent to a multiplier whose output is the desired reflux flow. This signal becomes the set point of the reflux flow controller, which is on cascade. This controller is installed in all design control structures. Whereas, reflux flowrate of the designed control structure III (CS3) is fixed.

The designed control structure IV (CS4), from the product column has the boilup ratio is high (V/B). Therefore, this control structure used to the bottom flow to control tray EB composition and reboiler duty to control base level.

For temperature on tray in recycle column is controlled by manipulating reflux ratio. This signal becomes the setpoint of the reflux flow controller, which is on cascade. While the designed control structure (CS1), (CS2) and (CS4), these are control temperature on tray by manipulating direct reflux flowrate as show in Table 5.5.

Table 5.5 Matching CVs with MVs.

CVs		MVs			
		CS1	CS2	CS3	CS4
3-phase separator	%level	V-6	V-6	V-6	V-6
Reboiler of product column	%level	V-9	V-9	V-9	V-9
Condenser of recycle column	P	qc2	qc2	qc2	qc2
3-phase separator	T	q4	q4	q4	q4
R1_inlet stream	T	q1	q1	q1	q1
L1 stream of product column	T	kp1	kp1	kp1	kp1
Reboiler of recycle column	%level	V-11	V-11	V-11	qr1
R2_inlet stream	T	q3	q3	q3	q3
Condenser of product column	%level	V-8	V-8	V-8	V-8
Condenser of recycle column	%level	V-10	Reflux	V-10	V-10
Total EB stream	F	V-1	V-1	V-1	V-1
3-phase separator	P	V-5	V-5	V-5	V-5
LPs2_stream	F	V-4	V-4	V-4	V-4
Condenser of product column	P	qc1	qc1	qc1	qc1
LPs1_stream	F	V-3	V-3	V-3	V-3
Tray 28 of recycle column	Tol comp	qr2	qr2	qr2	qr2
Tray 7 of recycle column	T	Reflux	V-10	Reflux	Reflux
Tray 54 of recycle column	EB comp	qr1	qr1	qr1	V-11
3-phase separator	% aq	V-7	V-7	V-7	V-7

Note: F refers to molar flow rate, T refers to temperature, P refers to pressure, and CC refers to composition.

Step 6: 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 buildup of any chemical component within the process (Downs, 1992). 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 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 (Vasudevan et. al, 2009).

Component	input	generation	output	consumption	Accumulation (= inventory) controlled by
EB	fresh feed	0	0	eqs. 4.1-4.3	total EB flow control
Steam	fresh feed	0	water	eqs. 4.4-4.6	LPs/total EB ratio control
styrene	0	eq. 1	SM	0	1 st column composition control
benzene	fresh feed	eq. 2	tol/ben	0	2 nd column temperature control
toluene	0	eq. 3	tol/ben	0	
hydrogen	0	eqs. 4.1,4.4-4.6	lights	eq. 4.3	3- phase separator pressure control
ethylene	0	eq. 4.2	lights	eq. 4.4	
methane	0	eq. 4.3	lights	eq. 4.5	
CO ₂	0	eq. 4.6	lights	0	
CO	0	eqs. 4.4-4.5	lights	eq. 4.6	

Step 7: Selection of controller type: single loops or multiple loops.

In this research, controller type is single loop and multiple loops. There are temperature controller, composition controller, pressure controller, flow controller, and level controller. Temperature controllers and composition controllers are PIDs which are tuned using relay feedback. Pressure controllers and flow controllers are PIs and their parameters are heuristics values. Proportional-only level controllers are used and their parameters are heuristic values. All control valves are half-open at nominal operating condition.

Step8: Validation via rigorous dynamic simulation.

Using software HYSYS to evaluate performance for styrene process of all designed control structures and compare with base case control structure (Luyben, 2011) at dynamic simulation.

5.2 Design of Plantwide control structures.

In this research, we designed the new plantwide control structures CS1 to CS4 of the styrene process. In all of these control structures, the same loops are used as follows:

Plantwide level

- Valve V-1 is manipulated to control the total ethylbenzene (fresh + recycle flow).

Unit level

Reaction section (Figure 5.3).

- Valve (V-3) is manipulated to control flow rate of the steam that is mixed with the total EB stream is ratioed to the flow rate of the total ethylbenzene.
- Valve (V-4) is manipulated to control flow rate of the steam that to furnace E2 is ratioed to the flow rate of the total ethylbenzene.
- Furnace duty (Q1) is manipulated to control the inlet temperature of the first reactor.
- Heater duty (Q3) is manipulated to control the inlet temperature of the second reactor.

Three-phase separator (Figure 5.4)

- Cooler duty (Q4) is manipulated to control the temperature separator.
- Valve V-5 is manipulated to control the pressure separator.
- Valve V-6 is manipulated to control organic level in the separator.
- Valve V-7 is manipulated to control water level in the separator.

Product column

- Condenser duty (qc1) is manipulated to control the condenser pressure.
- Distillate flow valve (V-8) is manipulated to control reflux drum level.
- Compressor duty (kp1) is manipulated to control the reflux drum temperature.

Recycle column

- Condenser duty (qc2) is manipulated to control the condenser pressure.
- Bottom flow valve (V-11) is manipulates to control the base level.

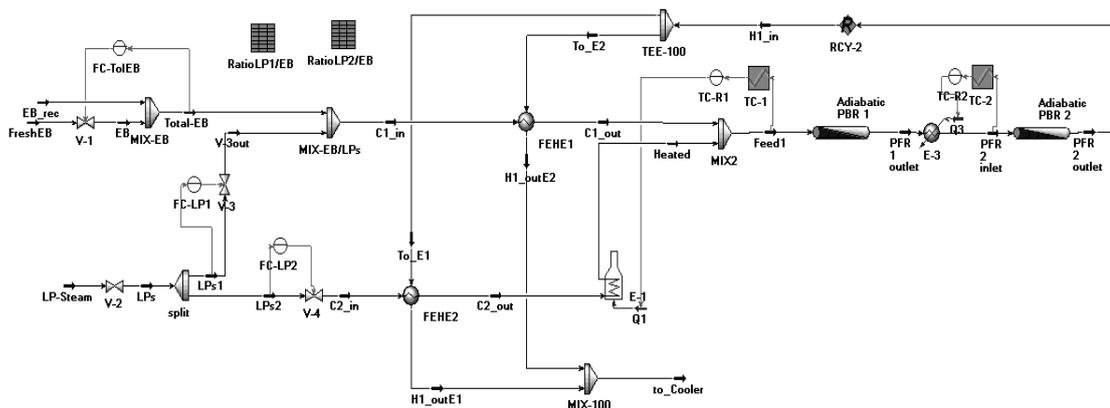


Figure 5.3 The control structure in reaction section of all designed structures (CS1 to CS4).

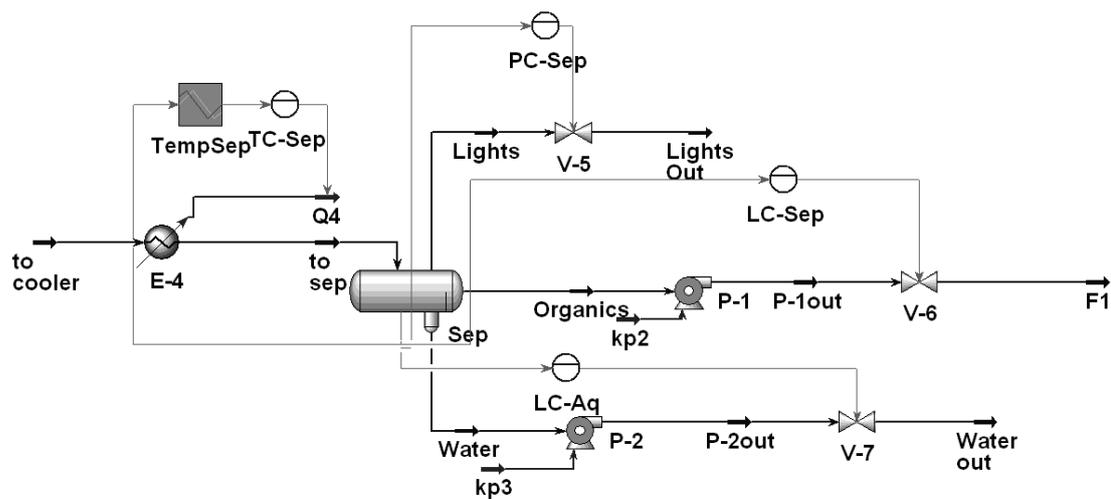


Figure 5.4 The control structure in three phase separator section of all designed structures (CS1 to CS4).

The main differences between CS1 to CS4 are explained below:

5.2.1 Design of control structure I (CS1)

The reaction section, inlet temperature for the first reactor (PBR1) is controlled by manipulating furnace duty (Q1). In product column, using a reflux-to-feed single-end control structure and it controlled the ethylbenzene impurity on tray 54th by manipulating reboiler duty (qr1) as shows in Figure 5.5.

In the recycle column used reflux flow for directly control temperature in tray 7th. Moreover, the toluene impurity on tray 28th is controlled by manipulating reboiler duty as shows in Figure 5.8.

5.2.2 Design of control structure II (CS2)

This control structure develops from CS1. Reflux flow rate is ratioed to product column feed flow rate is controlled by reflux flow rate. The ethylbenzene impurity on Tray 54th is controlled by manipulating direct reboiler duty (qr1) as shows in Figure 5.5.

For the recycle column, the toluene impurity on tray 28th is controlled by manipulating reboiler duty (qr2). By the way, temperature on tray 7th is controlled by manipulating reflux ratio, distillate flow valve (V-10) to keep a specified reflux ratio and the reflux- drum level is controlled by reflux flow as shows in Figure 5.9.

5.2.3 Design of control structure II (CS3)

This control structure develops from CS1, reflux flow in the product column is fixed as shows in Figure 5.6.

5.2.4 Design of control structure II (CS4)

This control structure develops from CS1. For the product column, using a reflux-to-feed single-end control structure, the ethylbenzene impurity on tray 54th is controlled by manipulating bottom flow valve (V-9) and base level is controlled by manipulating reboiler duty (qr1) as shows in Figure 5.7.

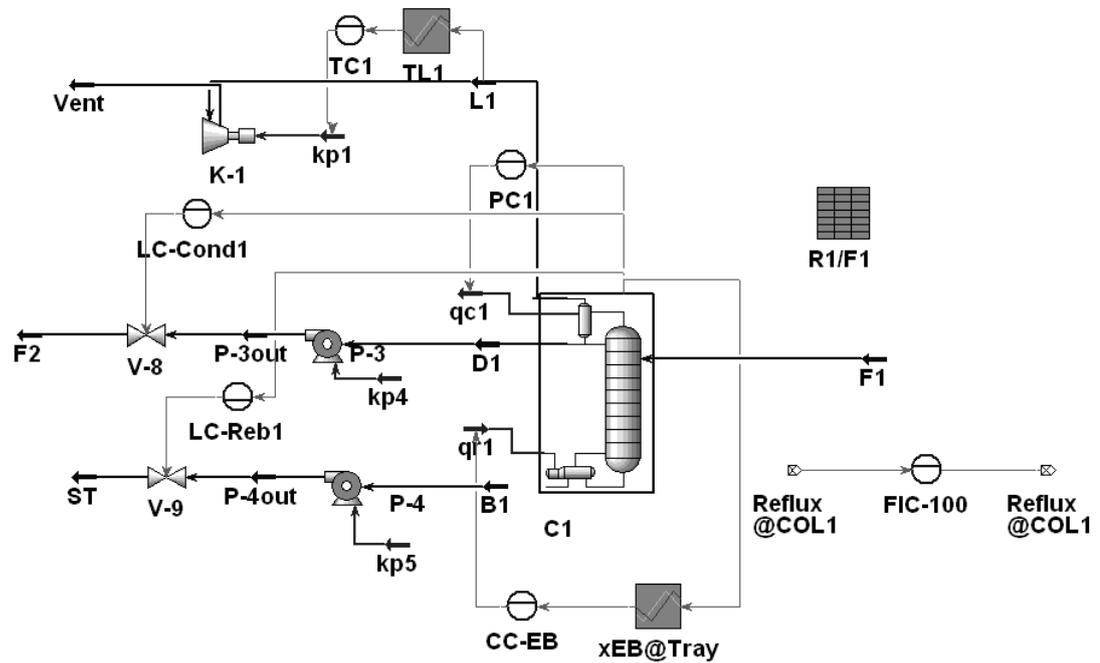


Figure 5.5 The control structure in product column of designed structures I, II (CS1 and CS2).

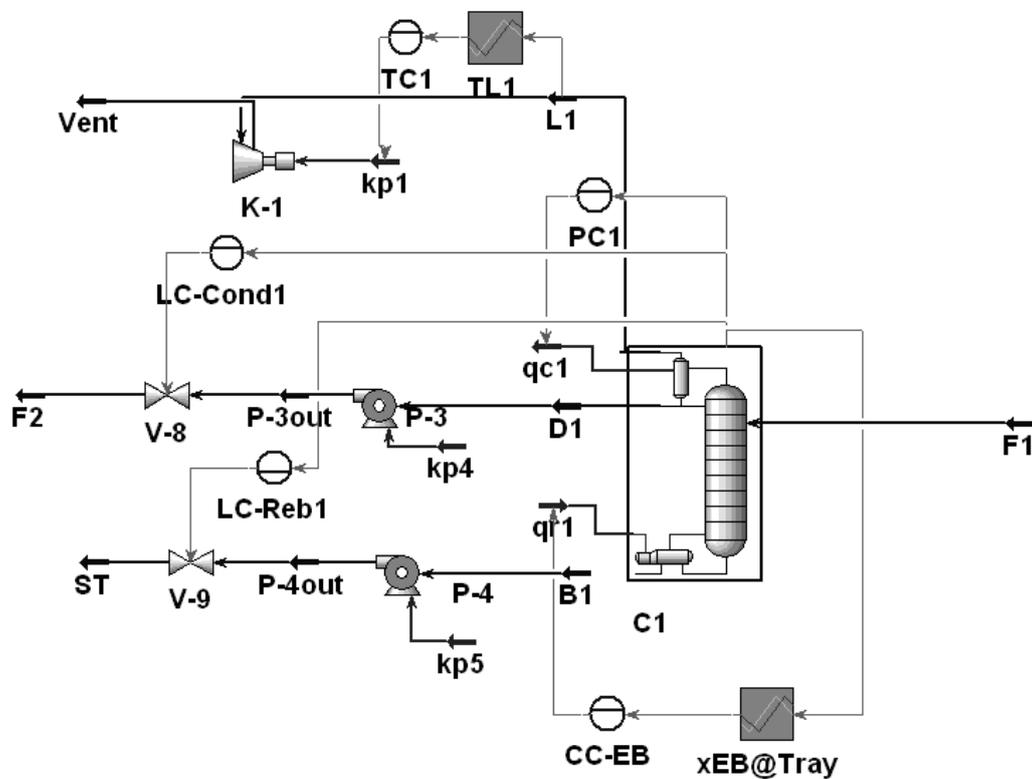


Figure 5.6 The control structure in product column of designed structures III (CS3).

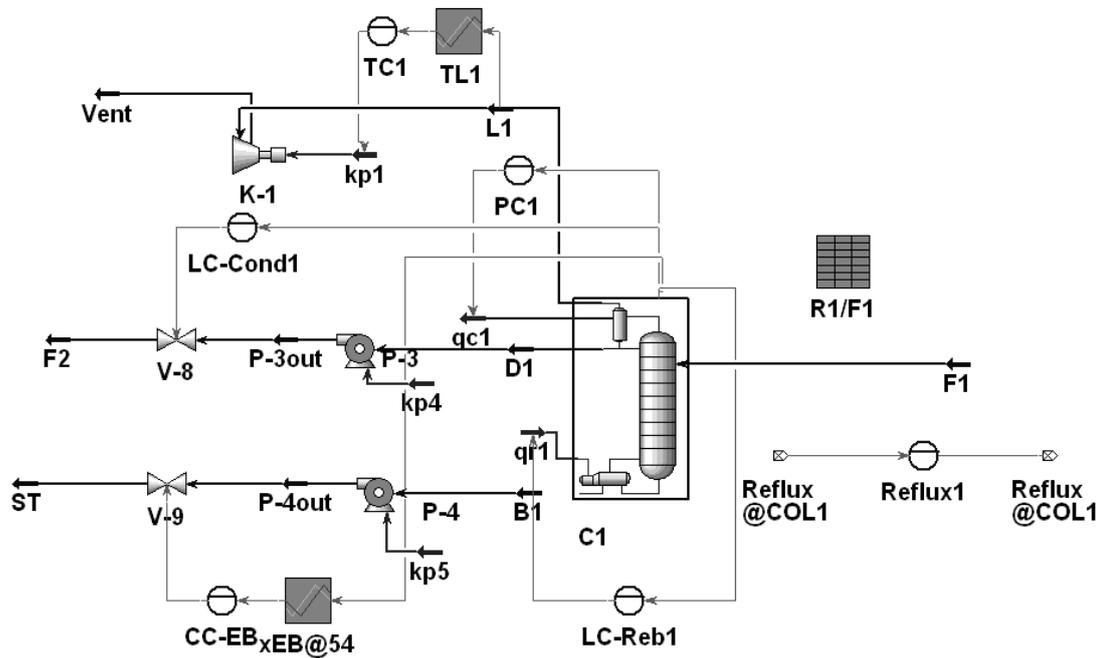


Figure 5.7 The control structure in product column of designed structure IV (CS4).

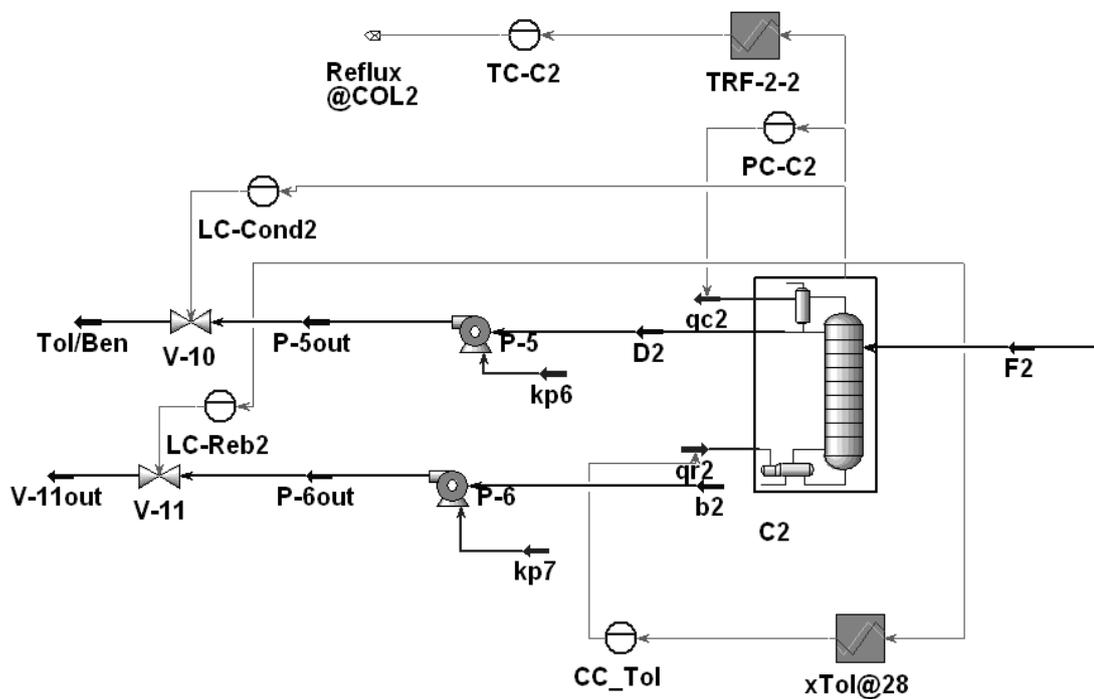


Figure 5.8 The control structure in recycle column of designed structures I, III, IV (CS1, CS3 and CS4).

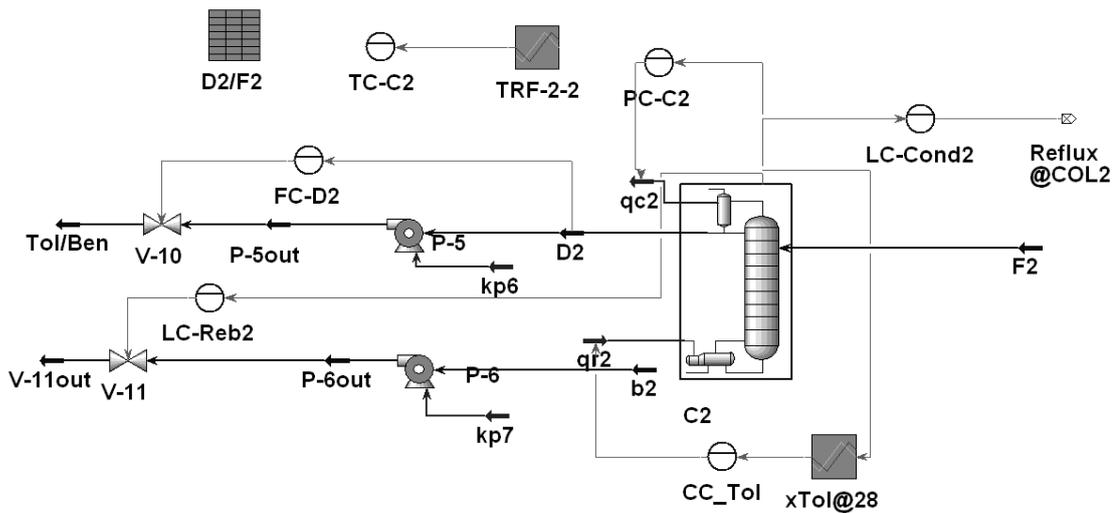


Figure 5.9 The control structure in recycle column of designed structure II (CS2).

5.3 Dynamic simulation results.

In order to illustrate the dynamic of base case control structure by Luyben (2011) and all designed control structures using new design procedure of Wongsri (2009), two types of disturbance: thermal and material disturbances are used to test the response of the system. Temperature controllers and composition controllers are PIDs which are tuned by relay-feedback test. The temperature loops have 1 min dead time and the composition loop has 3 min dead times. Flow and pressure controllers are PIs. Level loops are proportional-only. These parameters are heuristic values. All control valves are half-open at nominal operating conditions. The dynamic results are explained in this part.

5.3.1 Changes in material disturbances of the total ethylbenzene (fresh + recycle) flow rate for all control structures (base case (CS0), designed control structure (CS1-CS4))

Figure 5.10 shows the dynamic response of the styrene process by a step change in molar flow rate $\pm 10\%$ in the total ethylbenzene, increasing from 278.72 kmol/h to 306.59 kmol/h at 0.5 h to 10 h, decreasing from 306.59 kmol/h to 250.85 kmol/h at 10 h to 20 h, and then returning to its set point at time equals 20 h to 30 h. The graph of

dynamic responses of base case control structure (CS0), designed control structures (CS1-CS4) in each section shown in Figure 5.11.

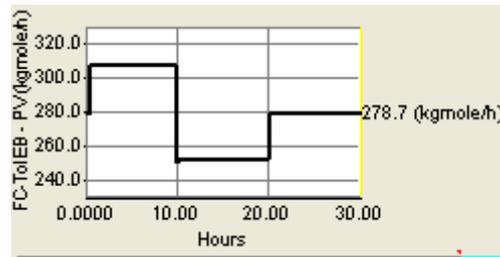


Figure 5.10 dynamic responses for molar flow rate $\pm 10\%$ in the total EB.

The change in the total EB immediately changes the two streams addition rate through the action of the ratio elements. Notice that there are large transient changes in the flow of the fresh EB feed. In all structures, the results give same trend to reject material disturbance and return to their set point as shown in the figures 5.11A (a-d).

In Figure 5.11A (e) shows dynamic response of temperature inlet to the first reactor, when total EB flow is changed. The transient deviations increase for these disturbances, but stable control can be achieved the desired values of reactors. The temperature control by adjusting the QE1/TotalEB ratio in CS0 takes more time to return to specified value than adjusting the temperature with direct furnace duty. This also affects the production rate. Notice for lights, and water flow rates from three-phase separator are increased when total EB is increased for all structure. As shown in Figure 5.11A (g,h)

Figure 5.11B shows how variable in product column changes for this disturbance, the dynamic responses of the purity of styrene product $x_{B1(ST)}$ of the designed control structure CS4 is more oscillatory than other control structures. However, all the control structures to maintain the quality of the product as specified show in Figure 5.11B (j). The feed to column F1 increased as the distillate and bottom flowrate increased. The reflux is ratioed to the feed flowrate ($R1/F1$) in control structures (CS0, CS1, CS2, and CS4), while CS3 reflux flow is fixed. Result shows that a change in flow rate at the bottom of this column would be less than others

control structures as show in figure 5.11B (i). The temperature and pressure at reflux drum are well controlled.

Figure 15.11C shows how variable in recycle column changes for this disturbance, the pressure condenser of all control structure (CS0 to CS4) are less deviation as show in Figure 5.11C (p). The changes in the flowrate of distillate (D2) are very slow and it comes to new steady state. These changes affect the temperature controller of this column, that is controlled by adjust reflux ratio ($R2/D2$) in base case control structure (CS0). Notices from Figure 5.11C(o), the temperature response in CS0 is more oscillatory than the designed control structures CS1, CS3, and CS4 which are controlled by adjusting reflux flow directly. As for the designed control structure II (CS2) that the temperature controller changes the distillate flow rate through the $D2/R2$ ratio and reflux is manipulated to control reflux- drum level. The tuning of refulux-drum level controller in this column impact the time it takes to come to setpoint of temperature controller from Figure 5.11C(o) can see the effective temperature controller as well.

I Figure 5.11C (r) found that the impurity of toluene in the bottom of recycle column shows a large transient deviation for changes in total EB flowrate but is brought back to very close to its setpoint.

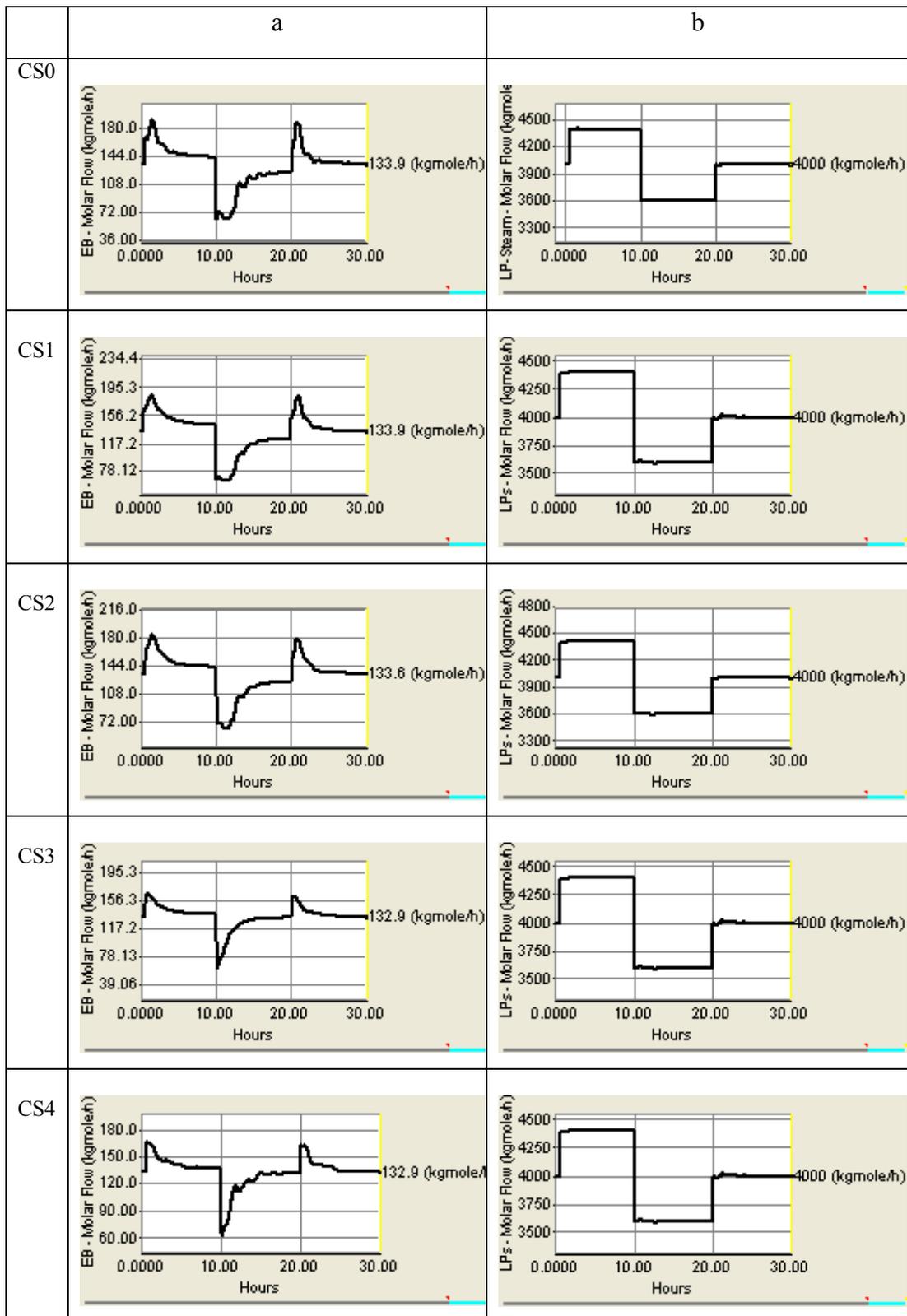


Figure 5.11A Dynamic responses of reaction section when changes in total EB (fresh + recycle flow): (a) EB fresh flow, (b) low pressure steam (LPs) feed flow.

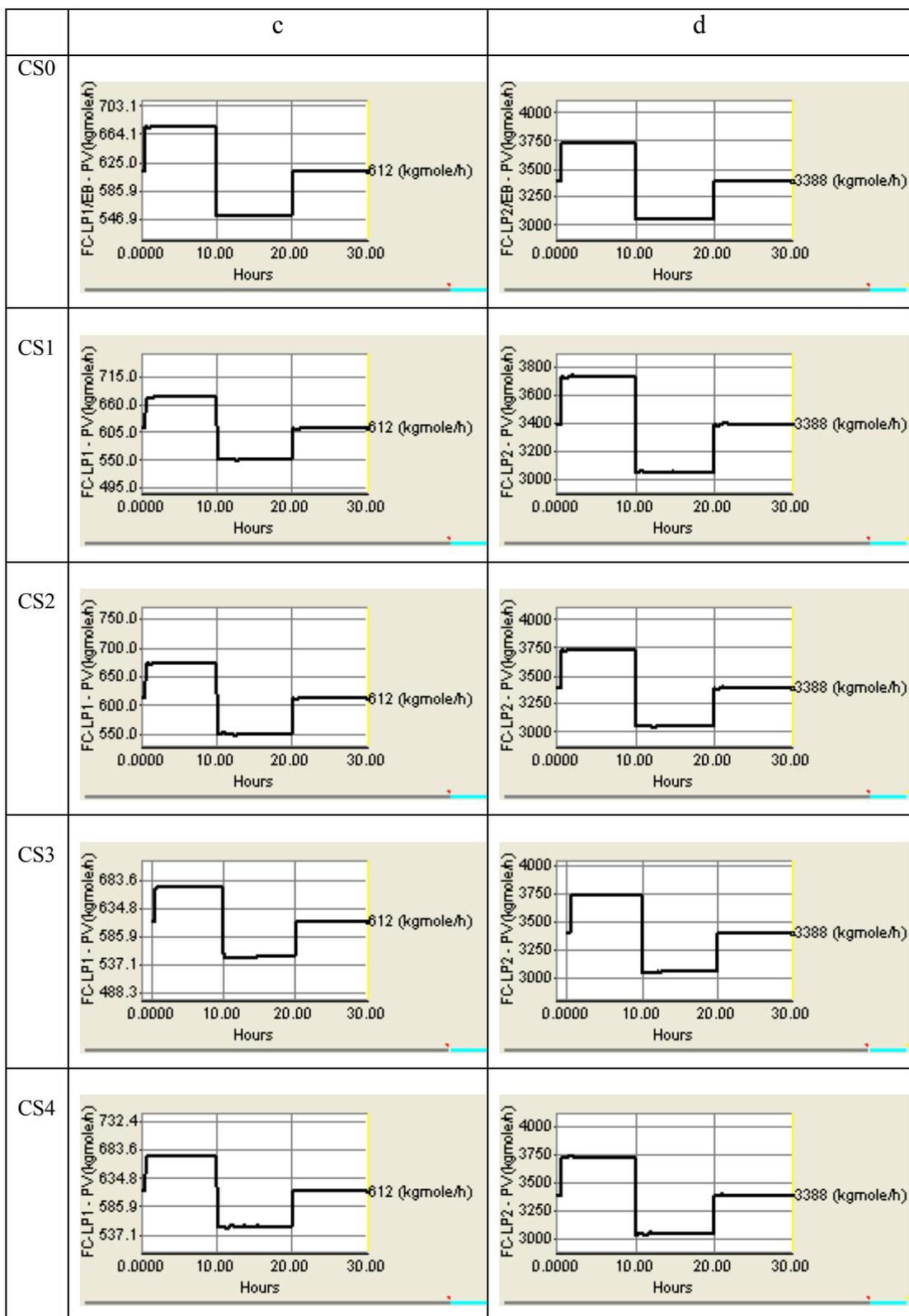


Figure 5.11A (Continued) Dynamic responses of reaction section when changes in total EB (fresh + recycle flow): (c) LPs1 split flow to mix the total EB flow, (d) LPs2 flow fed to furnace.

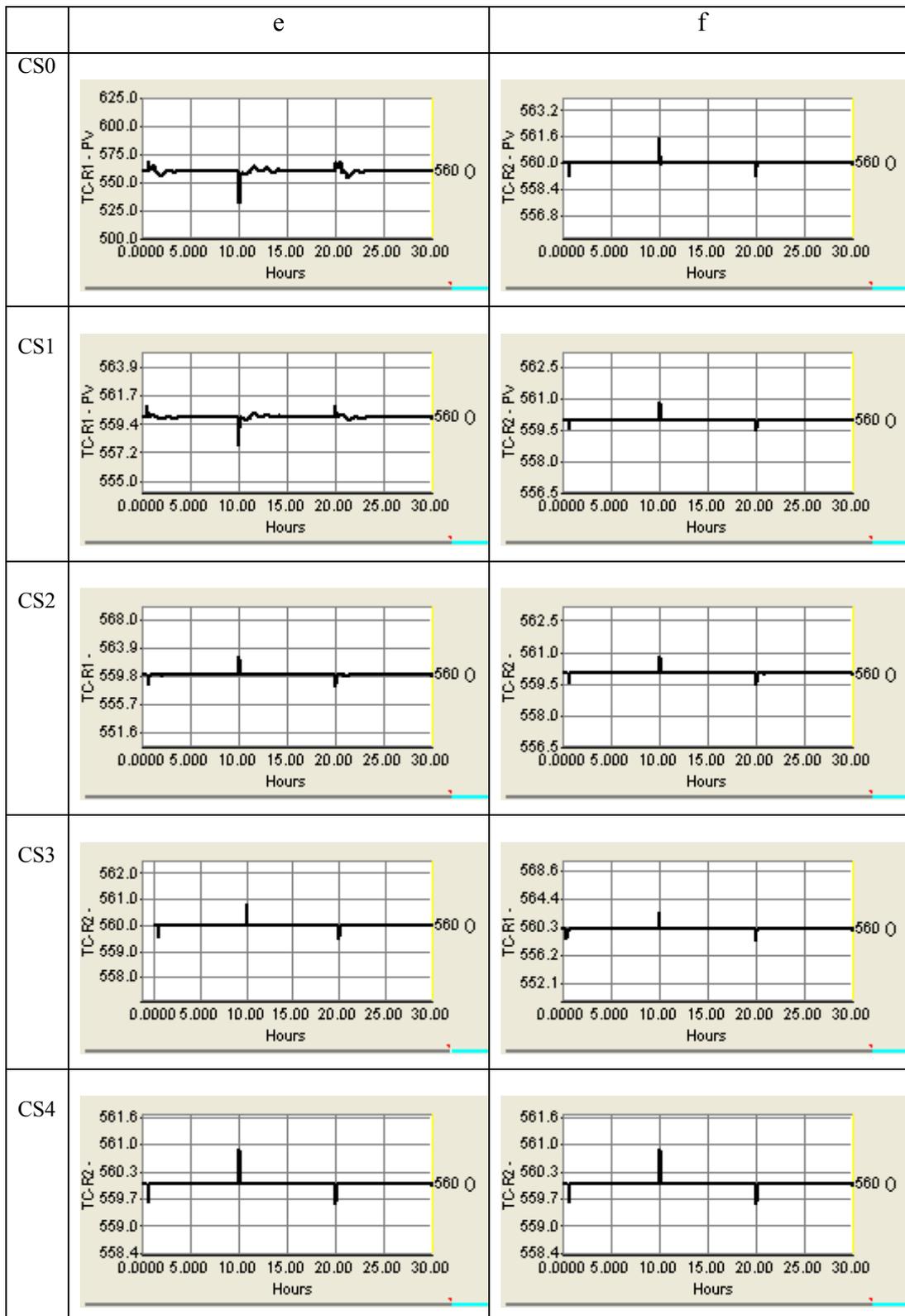


Figure 5.11A (Continued) Dynamic responses of reaction section when changes in total EB (fresh + recycle flow): (e) temperature inlet to PBR1 reactor, (f) temperature inlet to PBR2 reactor.

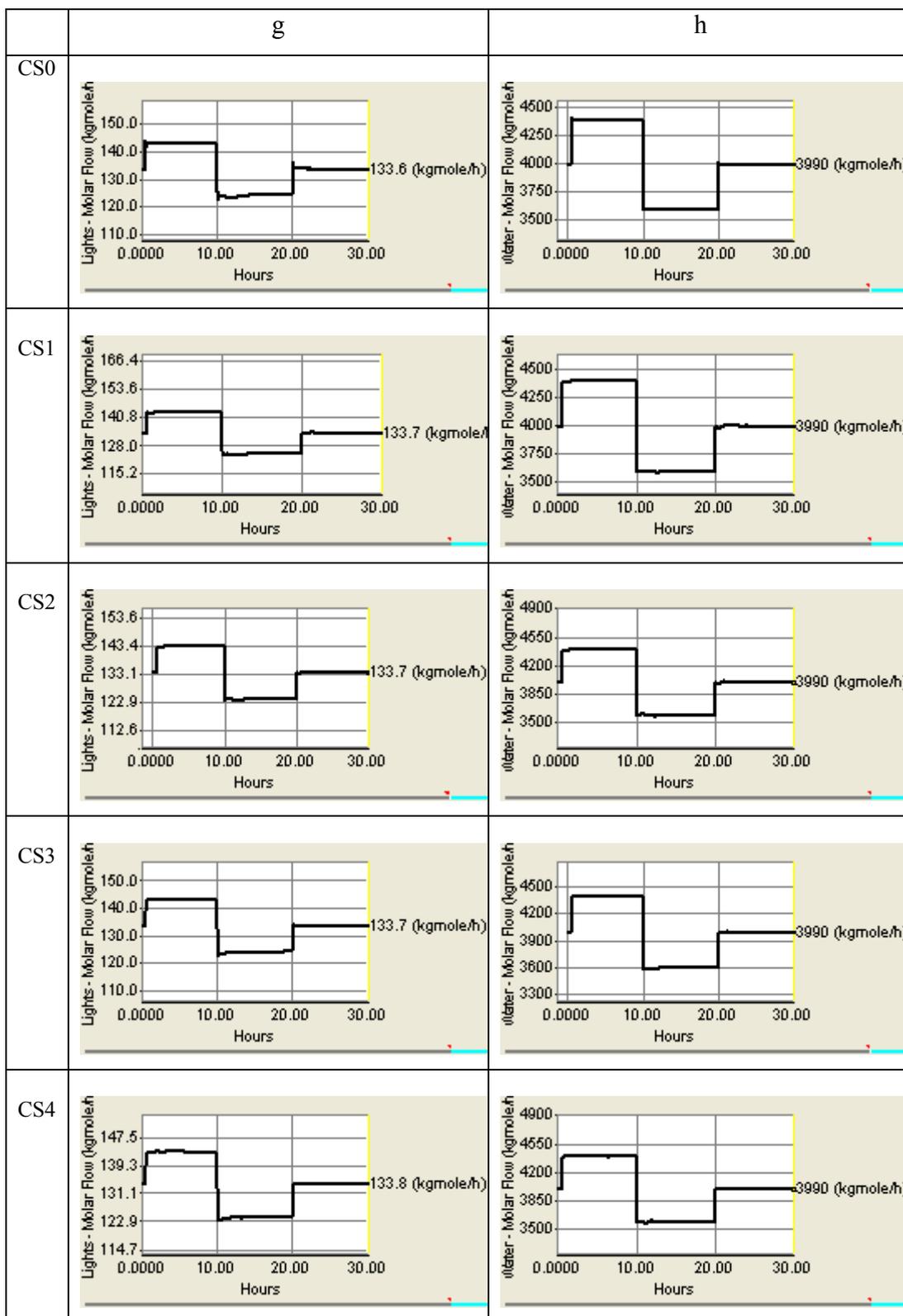


Figure 5.11A (Continued) Dynamic responses of reaction section when changes in total EB (fresh + recycle flow): (g) light molar flow, and (h) water molar flow from 3-phase separator.

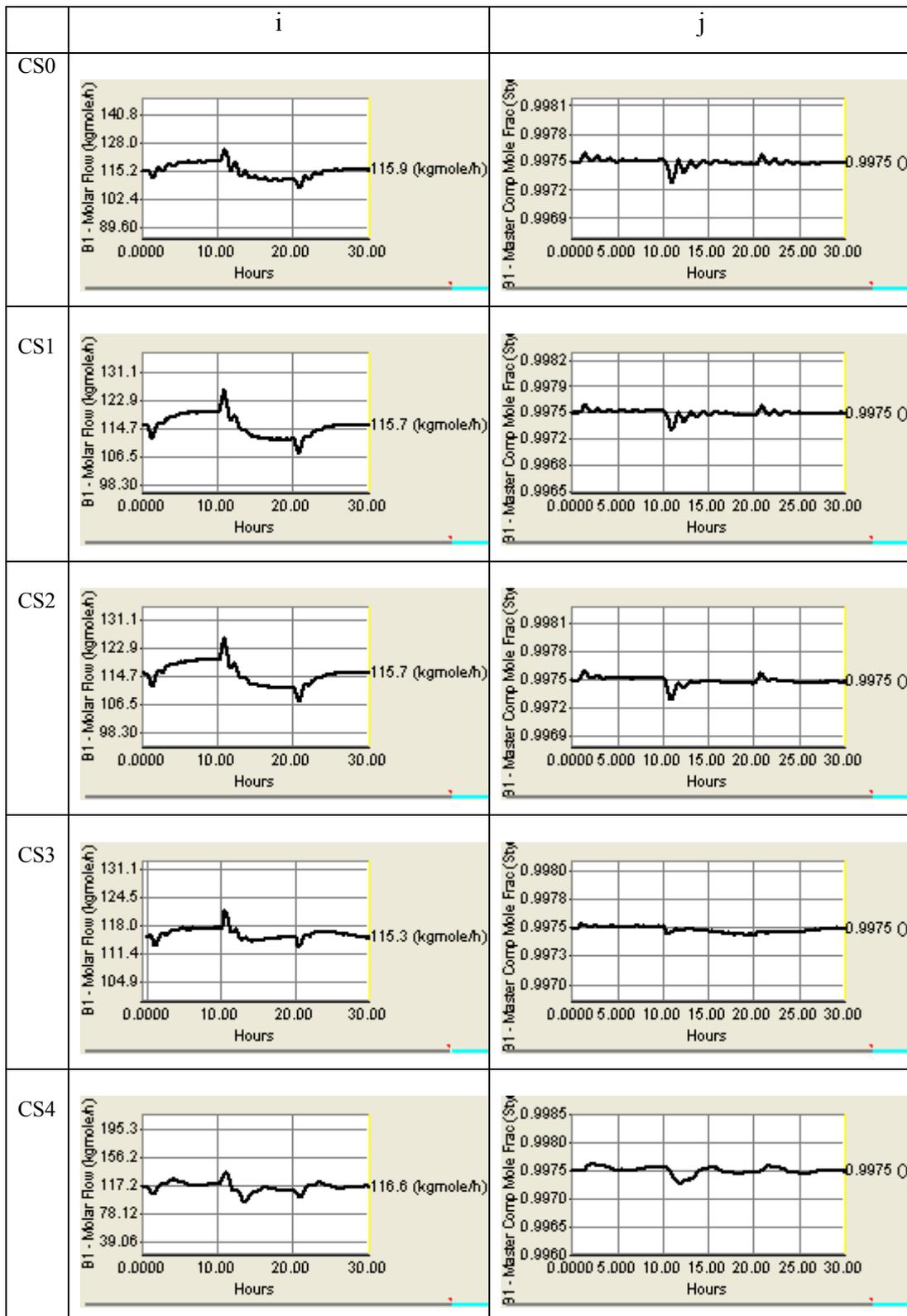


Figure 5.11B Dynamic responses of product column when changes in total EB (fresh + recycle flow): (i) bottom molar flow, and (j) styrene composition at bottom of product column.

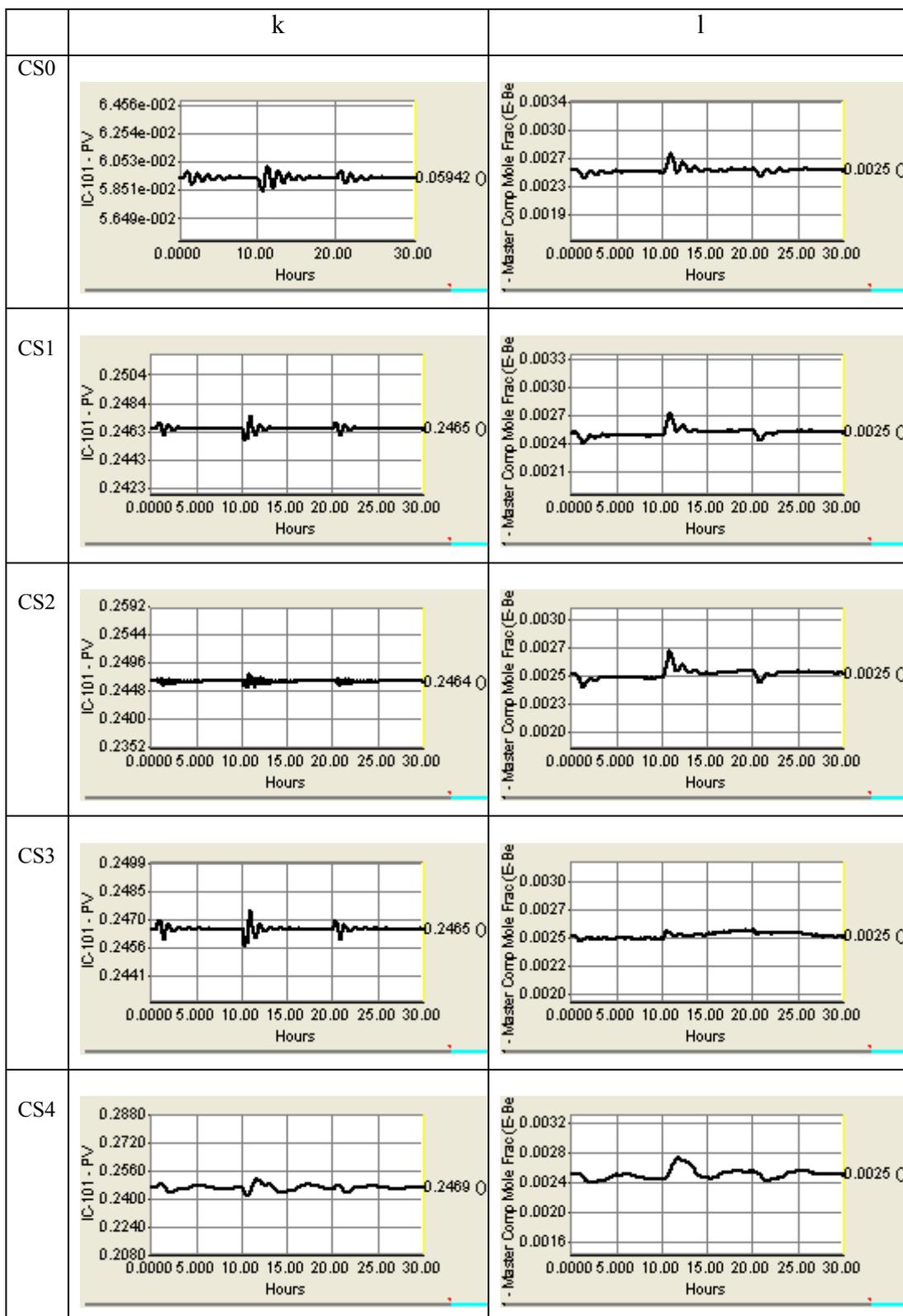


Figure 5.11B (Continued) Dynamic responses of product column when changes in total EB (fresh + recycle flow): (k) EB composition on tray 54th and (l) EB composition at bottom of product column.

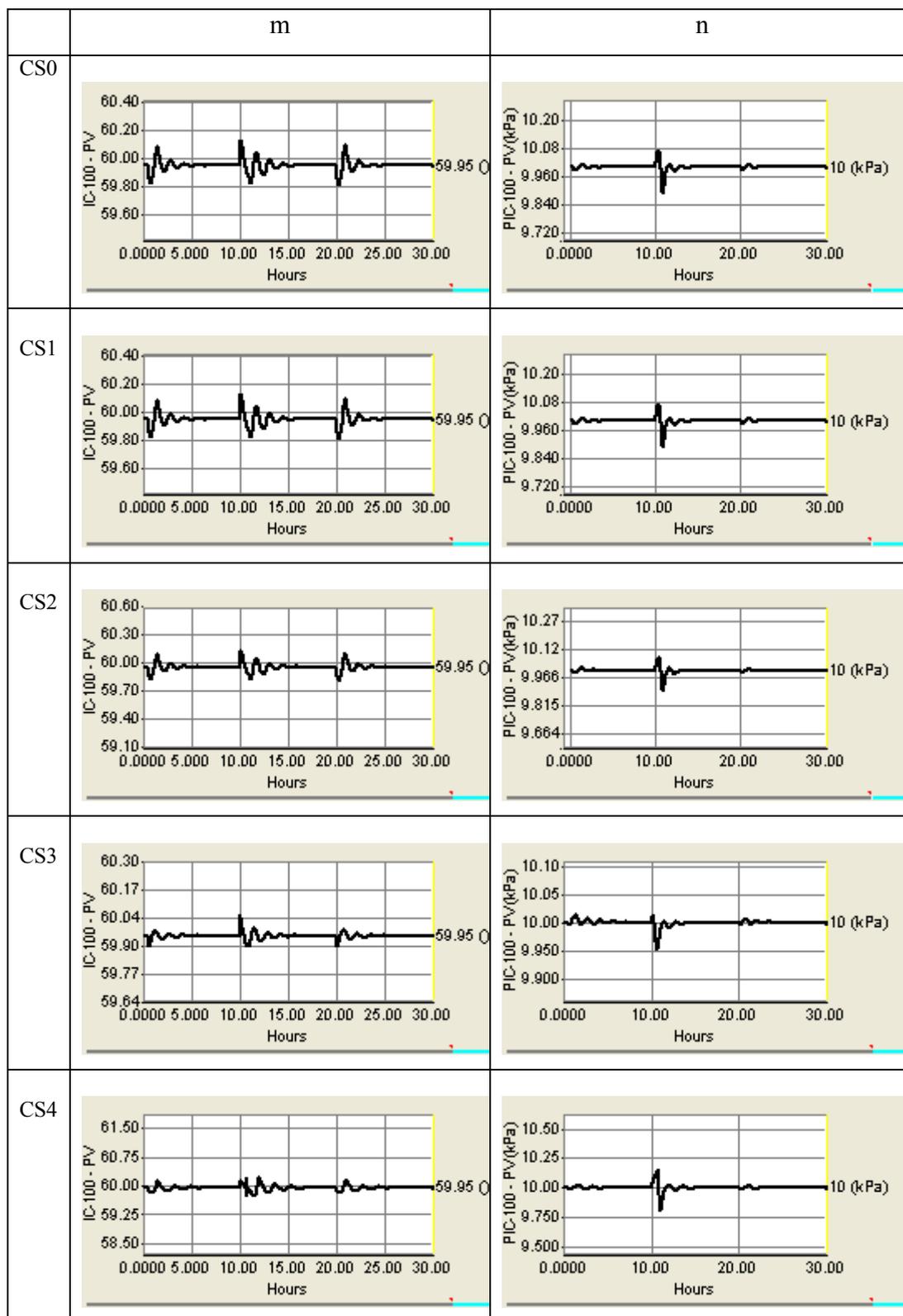


Figure 5.11B (Continued) Dynamic responses of product column when changes in total EB (fresh + recycle flow): (m) top product column temperature, (n) product column- condenser pressure.

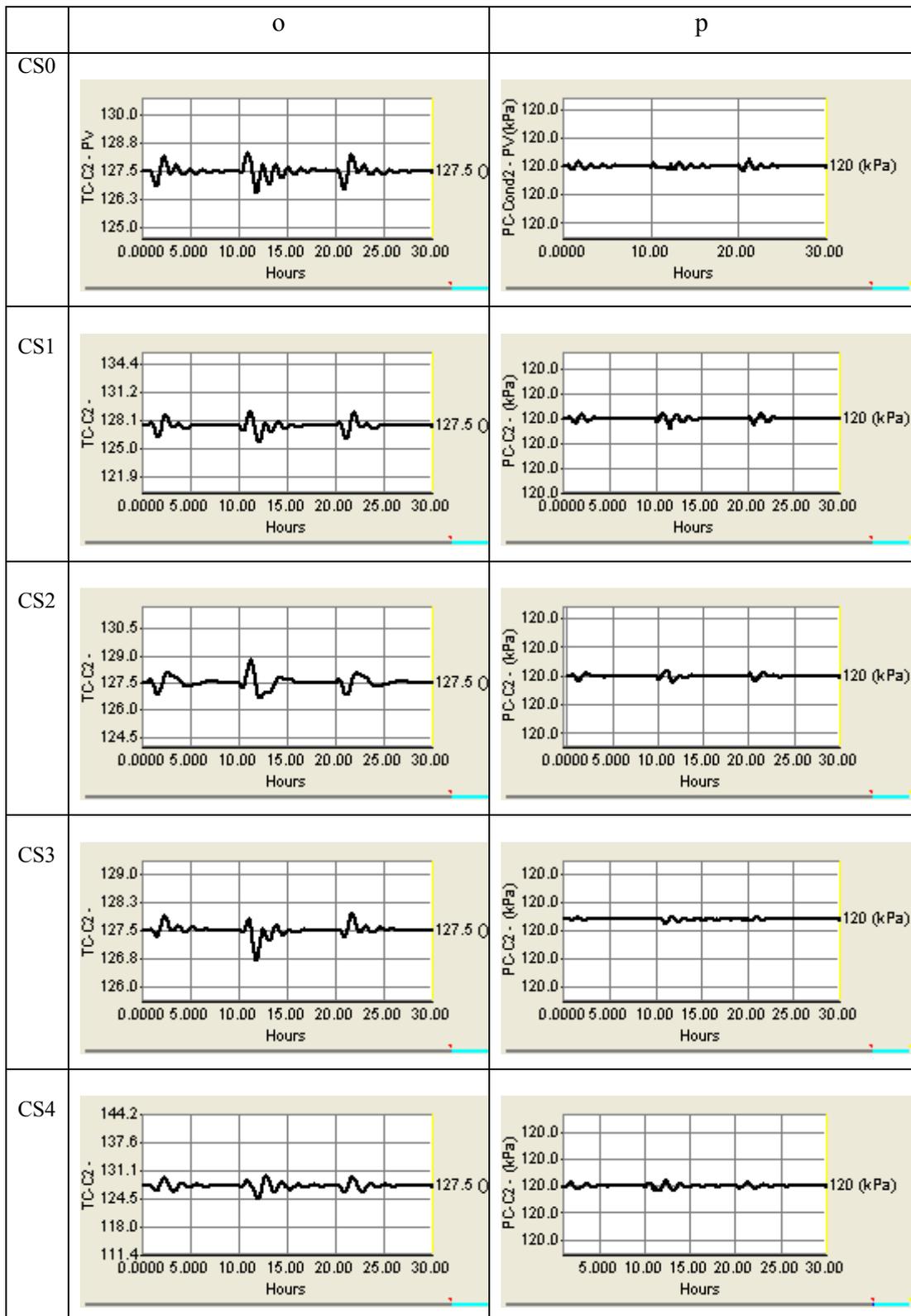


Figure 5.11C Dynamic responses of recycle column when changes in total EB (fresh + recycle flow): (o) temperature on Tray 7th in recycle column, (p) recycle column-condenser pressure.

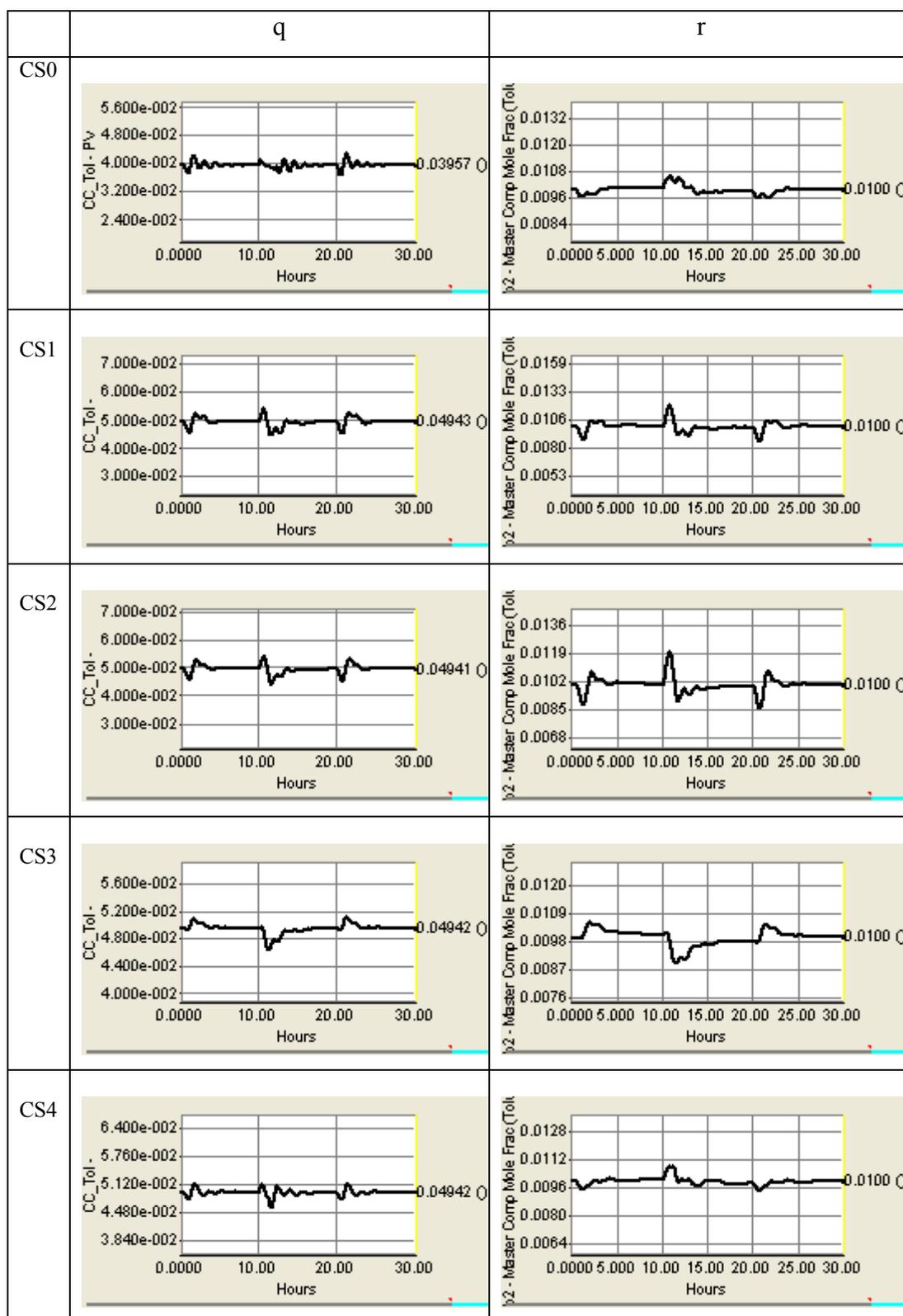


Figure 5.11C (Continued) Dynamic responses of recycle column when changes in total EB (fresh + recycle flow): (q) toluene composition on tray and (r) toluene composition at bottom of recycle column.

5.3.2 Changes in the thermal disturbances of two inlet temperature of PBR1 and PBR2 reactors for all control structure (base case control structure (CS0), designed control structures (CS1-CS4)).

Figure 5.12 shows dynamic responses for styrene process by step change temperature ± 10 ($^{\circ}\text{C}$) in inlet temperature of PBR1 and PBR2 reactors increase from 560°C to 570°C at 0.5 h to 10 h and decrease from 570°C to 550°C at 10 h to 20 h and then return to its set point at time equals 20 h to 30 h. Graph of dynamic responses of base case control structure (CS0), designed control structures (CS1-CS4) in each section shown in Figure 5.13.

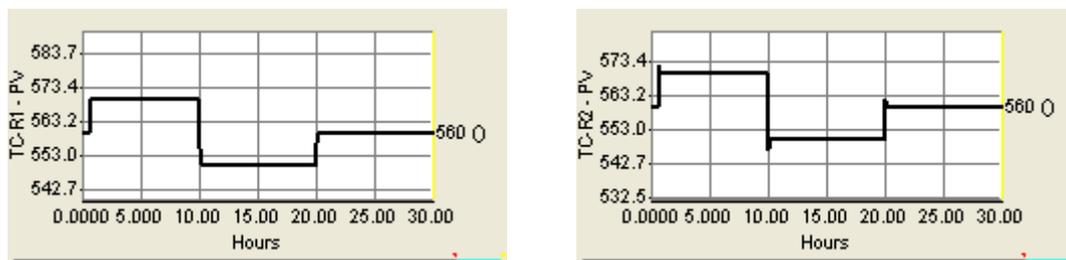


Figure 5.12 dynamic responses for change temperature ± 10 ($^{\circ}\text{C}$) of two reactors.

When changes setpoint of the two reactor inlet temperature controllers. There are no impact on the feed flow rate of the EB and steam. The total EB and steam are held constant. However, this disturbance affects the reaction rate of byproduct. The higher temperature leads to increases in lights, gas, and byproduct flow rates.

The control loops in the both column are affected very slightly when temperature changes, the purity of styrene product is maintained very close to its specification show in Figure 5.12B (e). In all control structure, results show the same trend to reject thermal disturbance that can reject disturbance and return to their set point.

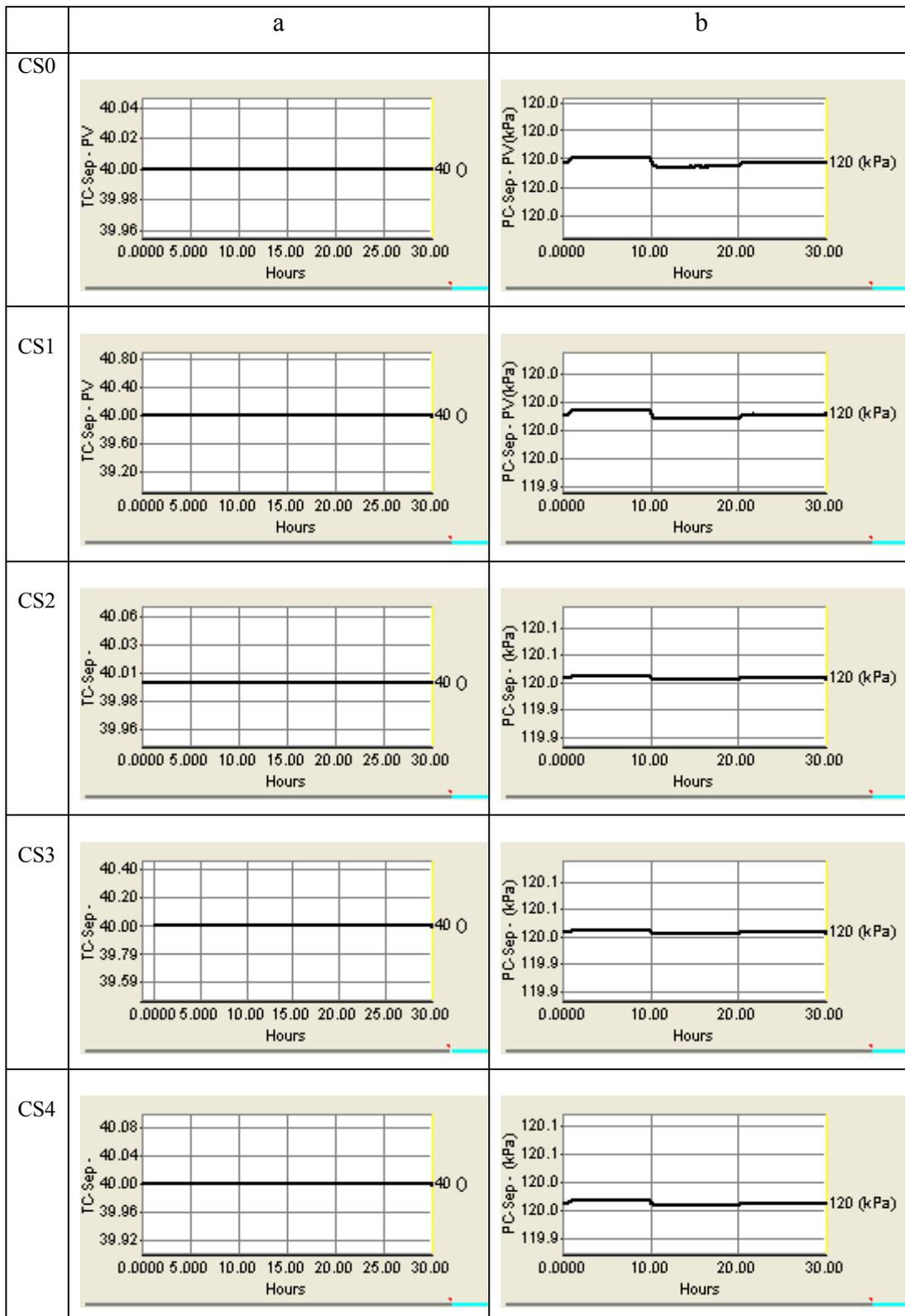


Figure 5.13A Dynamic responses of reaction section when changes in temperature inlet of two reactors: (a) 3-phase separator temperature, (b) 3-phase separator pressure.

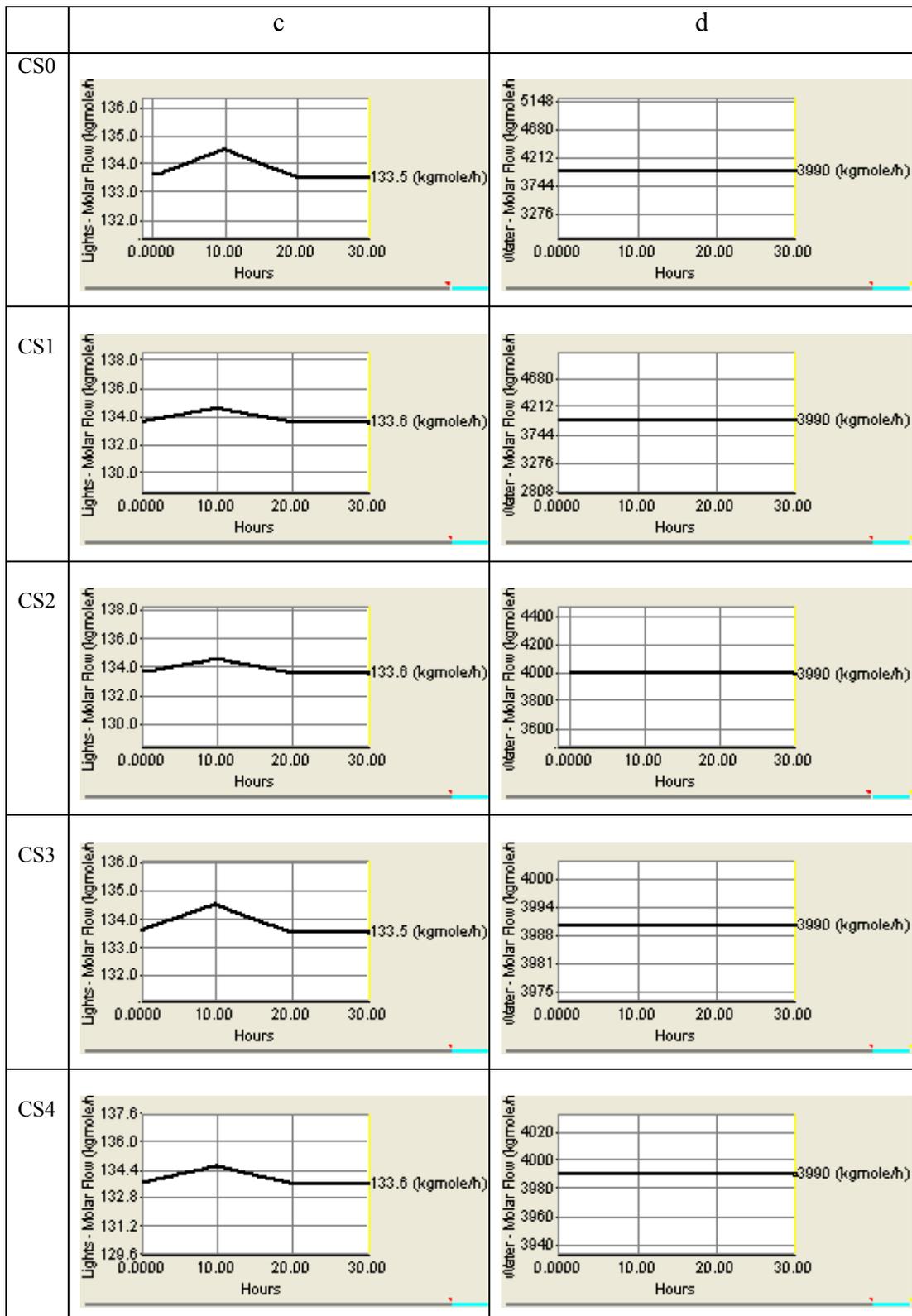


Figure 5.13A (Continued) Dynamic responses of reaction section when changes in temperature inlet of two reactors: (c) light molar flow, and (d) water molar flow at 3-phase separator.

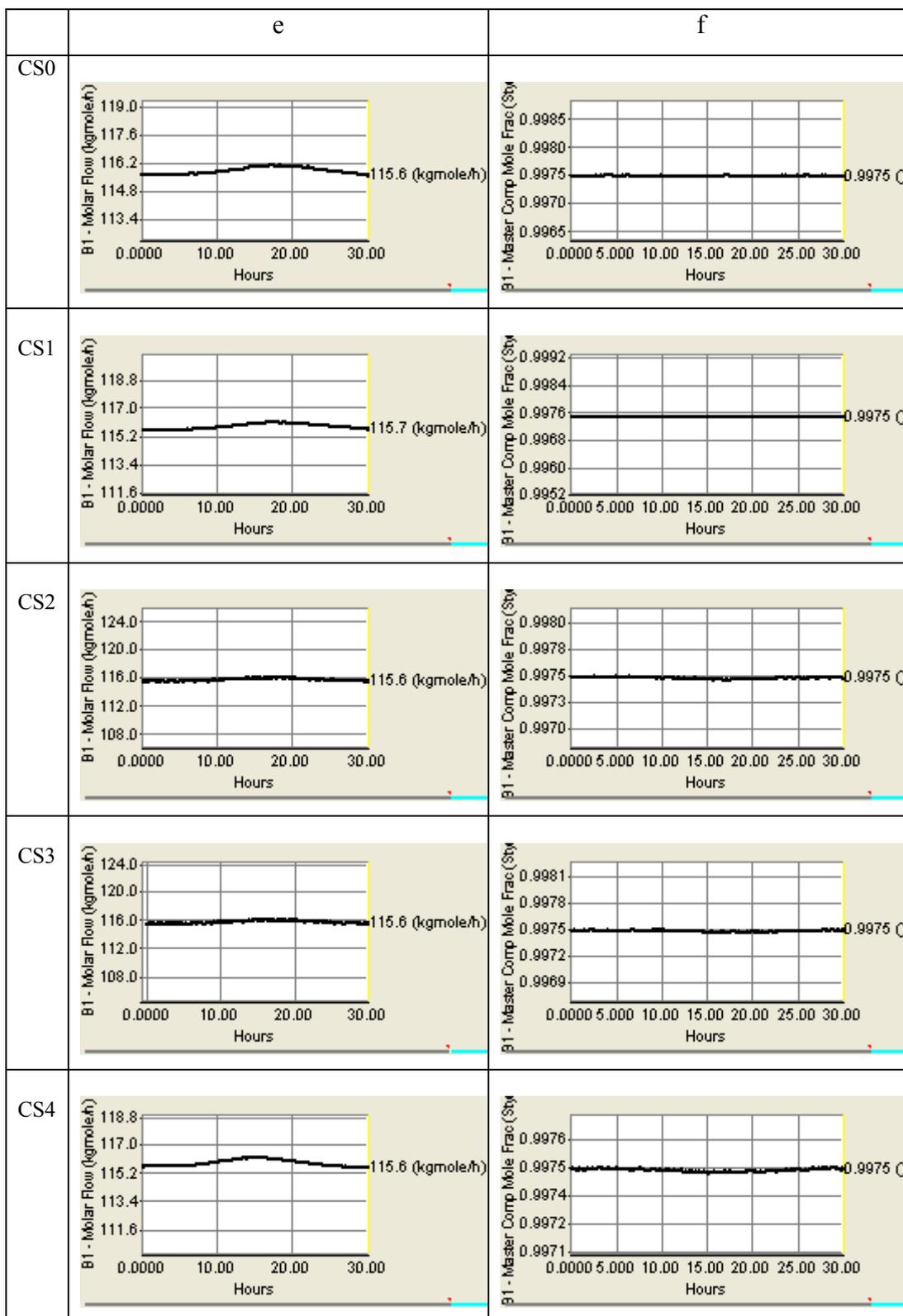


Figure 5.13B Dynamic responses of product column when changes in temperature inlet of two reactors: (e) bottom molar flow, and (f) styrene composition at bottom of product column.

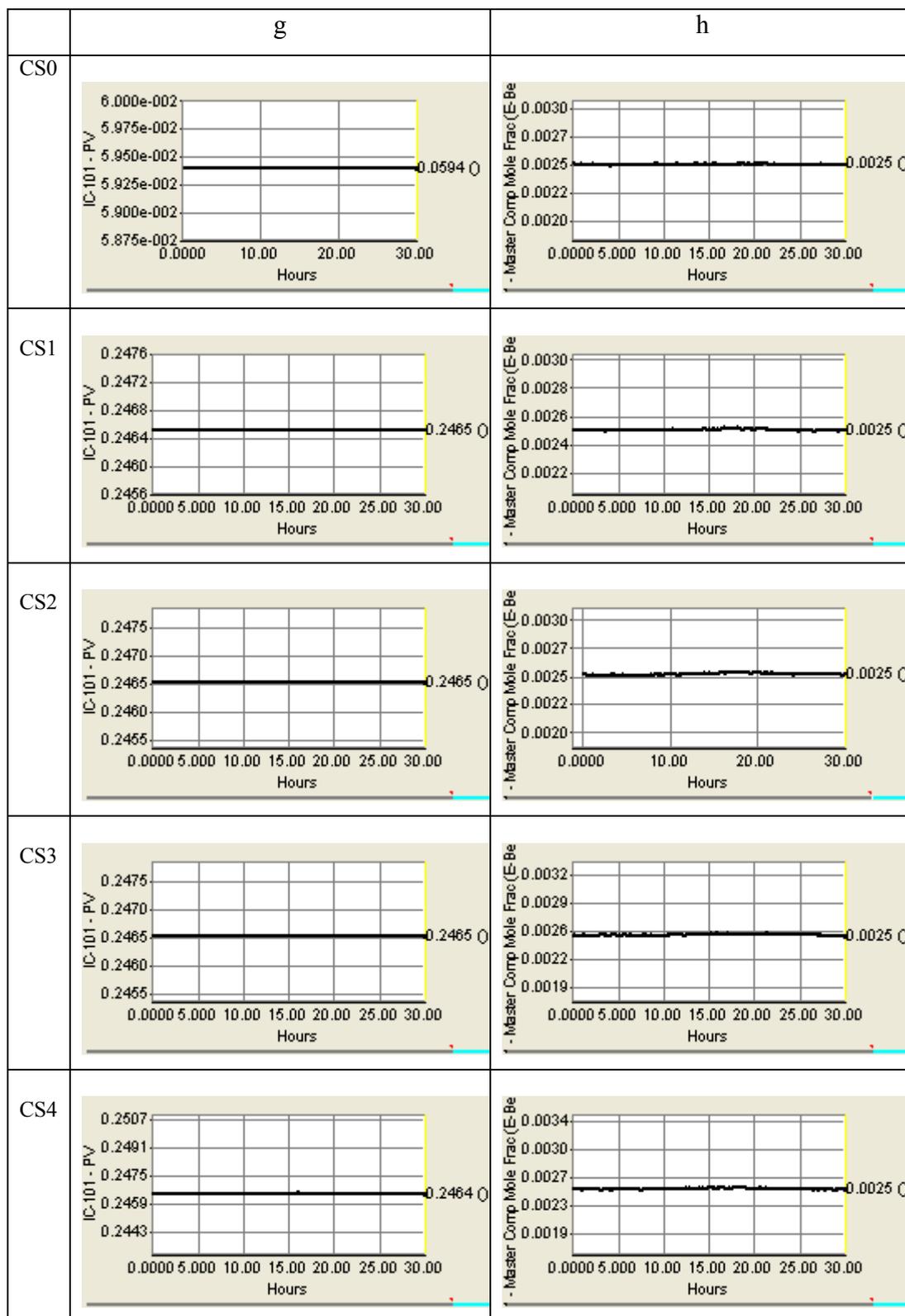


Figure 5.13B (Continued) Dynamic responses of product column when change in temperature inlet of two reactors: (g) EB composition on tray 54th, and (h) EB composition at bottom of product column.

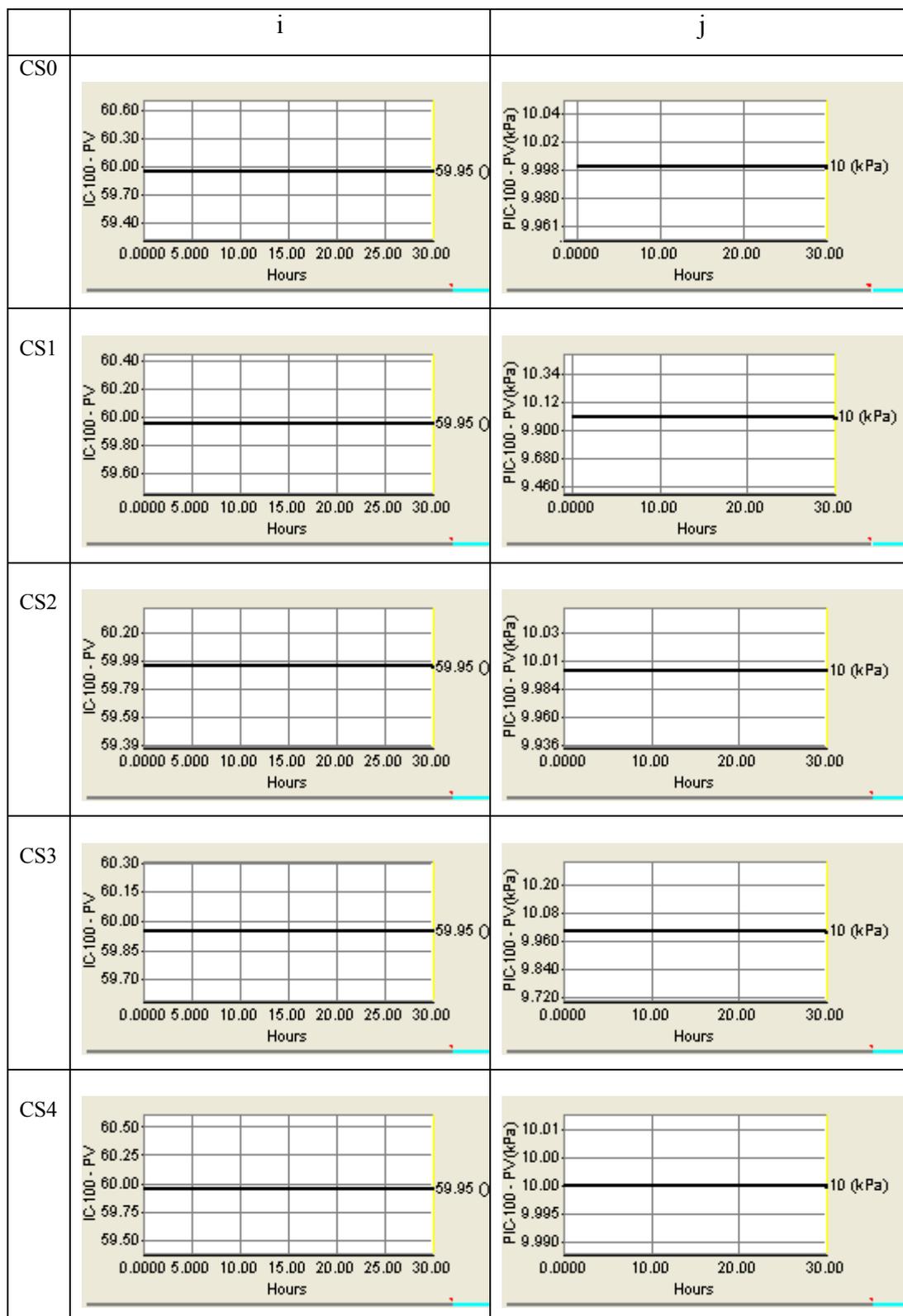


Figure 5.13B (Continued) Dynamic responses of product column when change in temperature inlet of two reactors: (i) top product column temperature, (j) product column- condenser pressure.

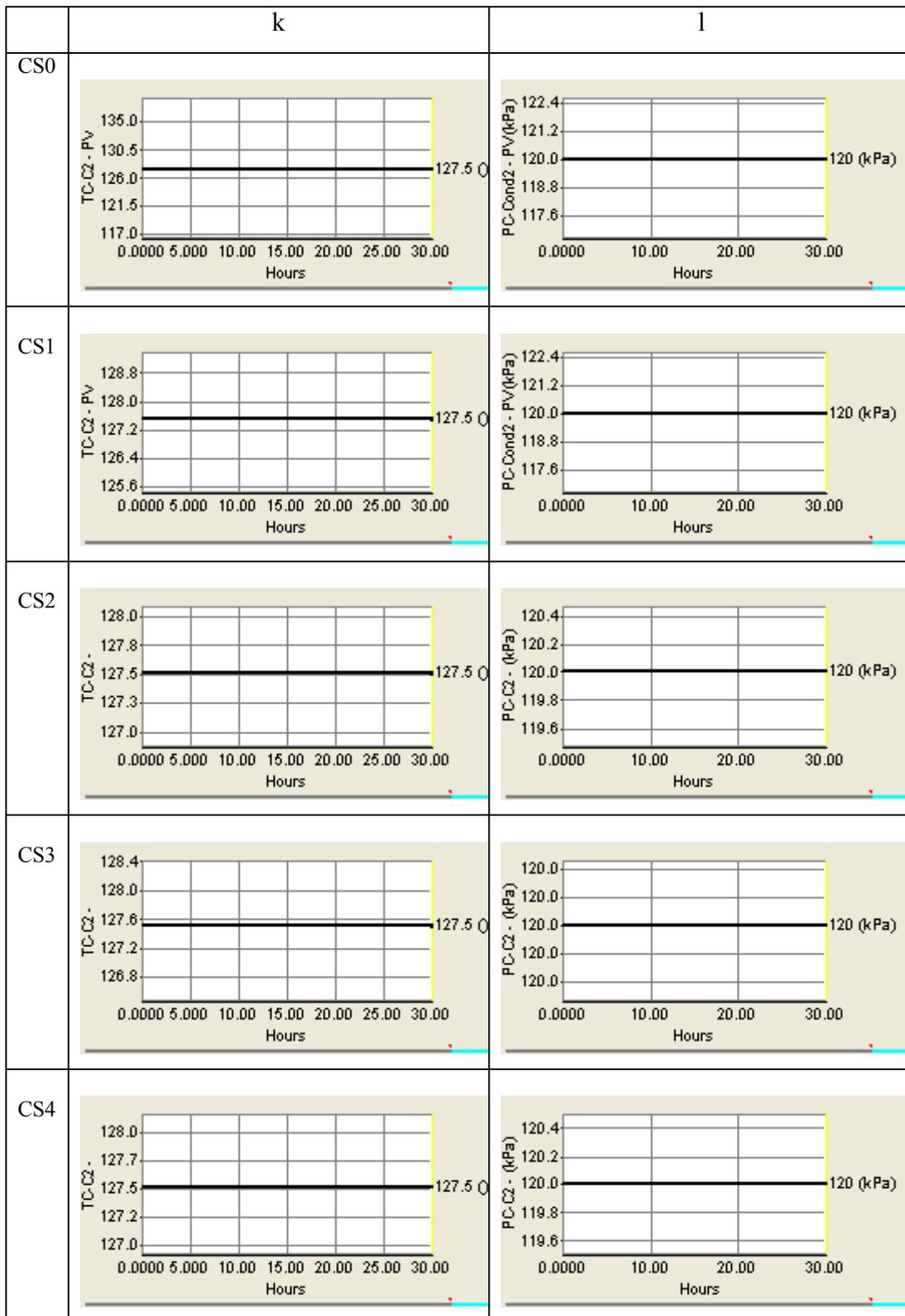


Figure 5.13C Dynamic responses of recycle column when change in total EB temperature inlet of two reactors: (k) temperature on Tray 7th in recycle column, (l) recycle column- condenser pressure.

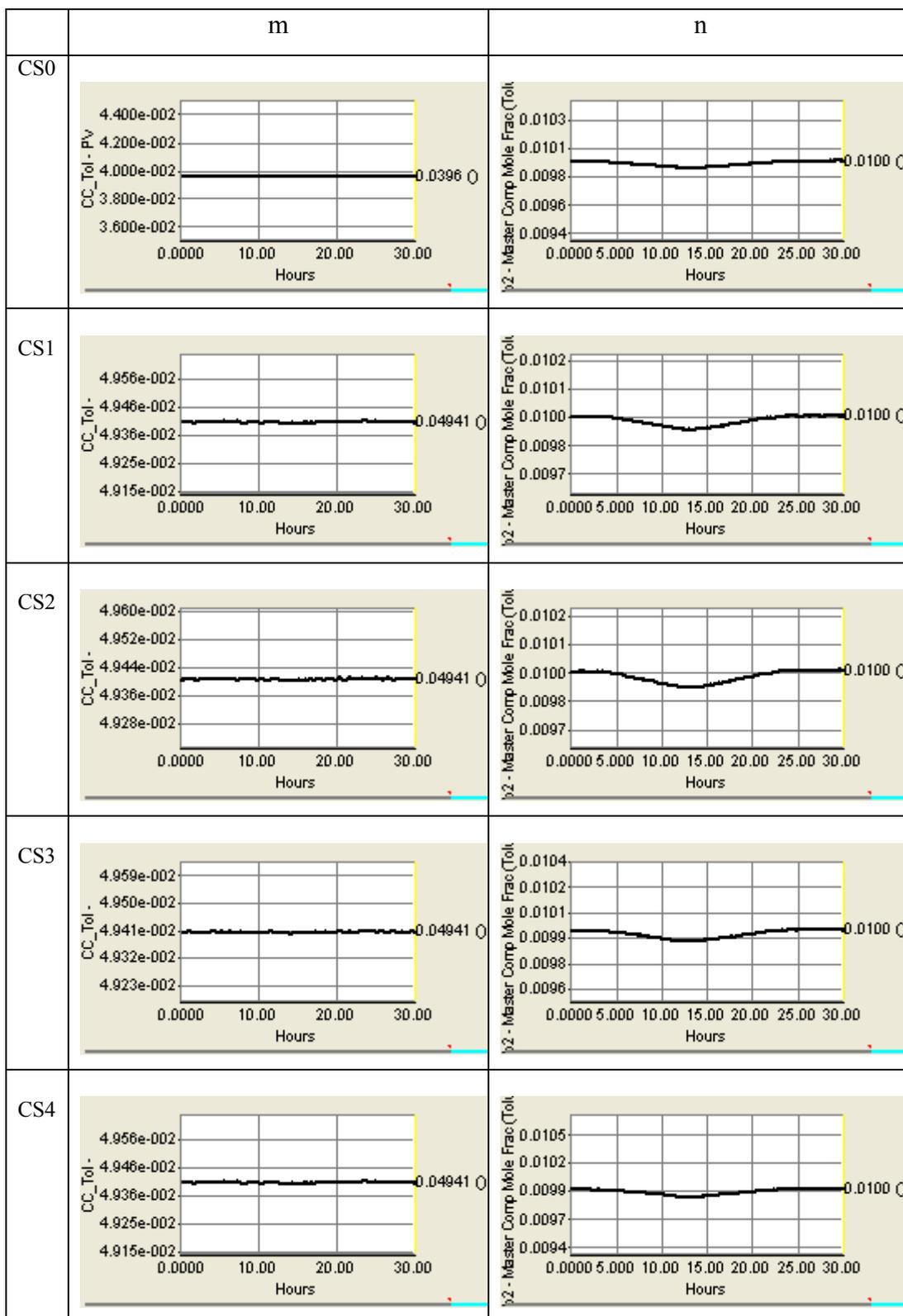


Figure 5.13C (Continued) Dynamic responses of recycle column when changes in temperature inlet of two reactors: (m) toluene composition on tray 28th and (n) toluene composition at bottom of recycle column.

5.3.3 Changes in ratio between the steam feds to furnace E2 and total EB for all control structures (base case (CS0)), designed control structure (CS1-CS4).

Figure 5.14 shows dynamic response of the styrene process by step change $\pm 10\%$ in ratio between the steam feds to furnace E2 and total EB increase from 12.156 to 13.371 at 0.5 h to 10 h, decrease from 13.371 to 10.940 at 10 h to 20 h. and then return to its set point at time equals 20 h to 30 h. The graph of dynamic responses of base case control structure (CS0), designed control structures (CS1-CS4) shown in Figure 5.15.

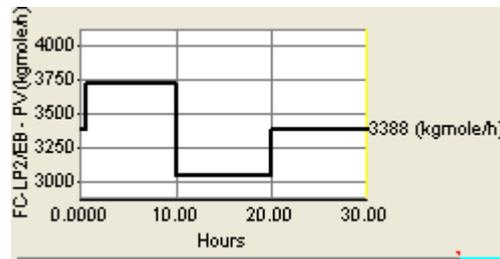


Figure 5.14 dynamic responses for $\pm 10\%$ in ratio between the steam feds to furnace E2 and total EB.

When the change ratio between the steam feds to furnace E2 and total EB. These disturbance affect to the production rate, notice from dynamic responses of Light, Gas, and styrene product rate. Such as the ratio increase leads to decrease in the product see in Figure 5.15B (g). Conversely, the EB recycle will increase.

For product column, the pressure condenser and temperature controller have good performance in all control structure show in Figure 5.13B (k, l).

Figure 5.15B (h) shows dynamic response of the purity of styrene in the bottom of product column in each structure that it a small transient deviation and can keep the quality of styrene product.

For recycle column, the temperature on tray 7th of designed control structures (CS1), (CS3), CS4) have a smaller deviation than the base case control structure (CS0), and designed control structure II (CS2) because the temperature on tray 7th is controlled by manipulating direct reflux flow show in Figure 5.15C (m). However all control structure, result the same trend to reject disturbance that can reject disturbance and return to their set point.

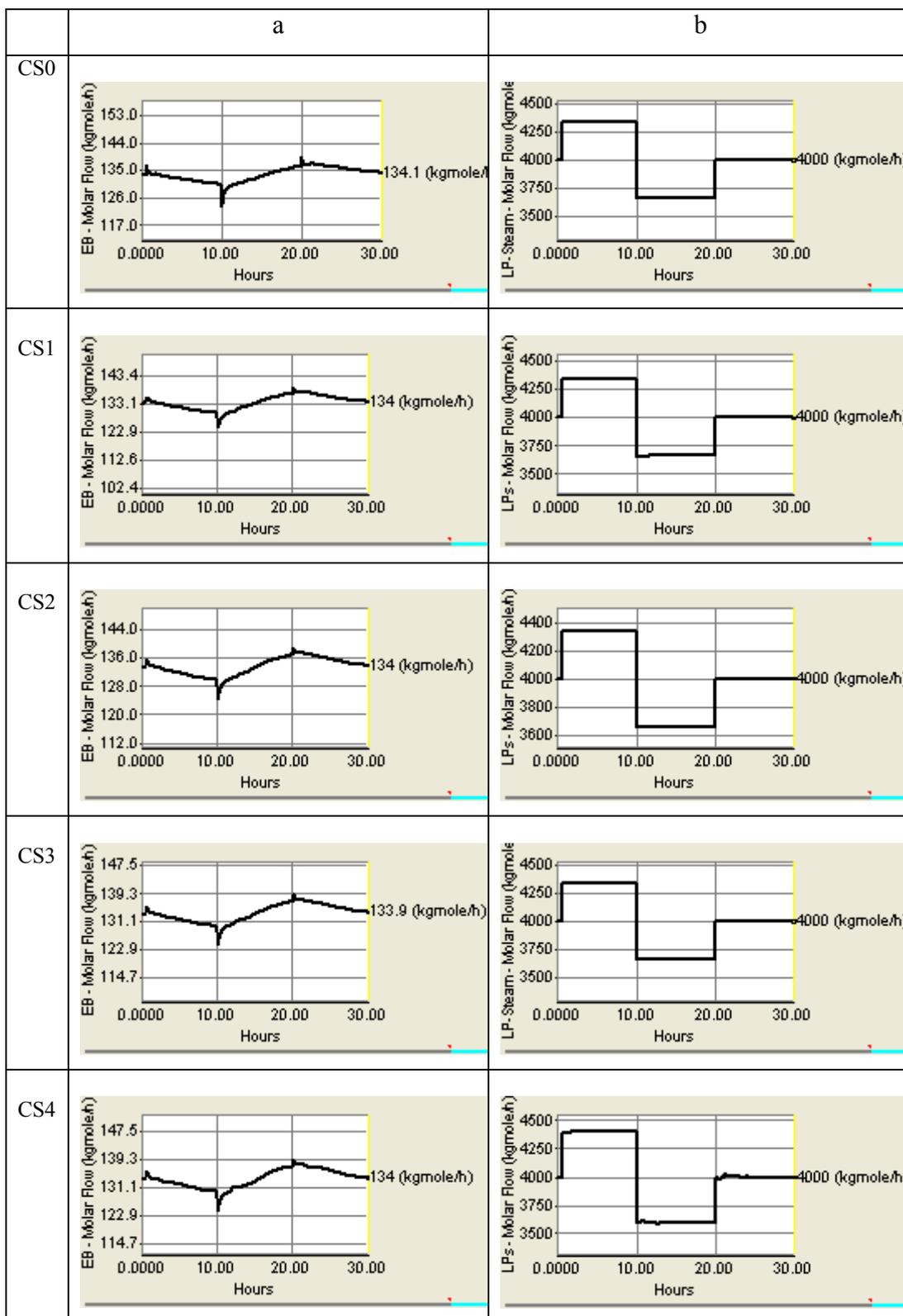


Figure 5.15A Dynamic responses of reaction section when change in ratio between the steam feeds to furnace E2 and total EB: (a) EB fresh flow, (b) low pressure steam (LPs) feed flow.

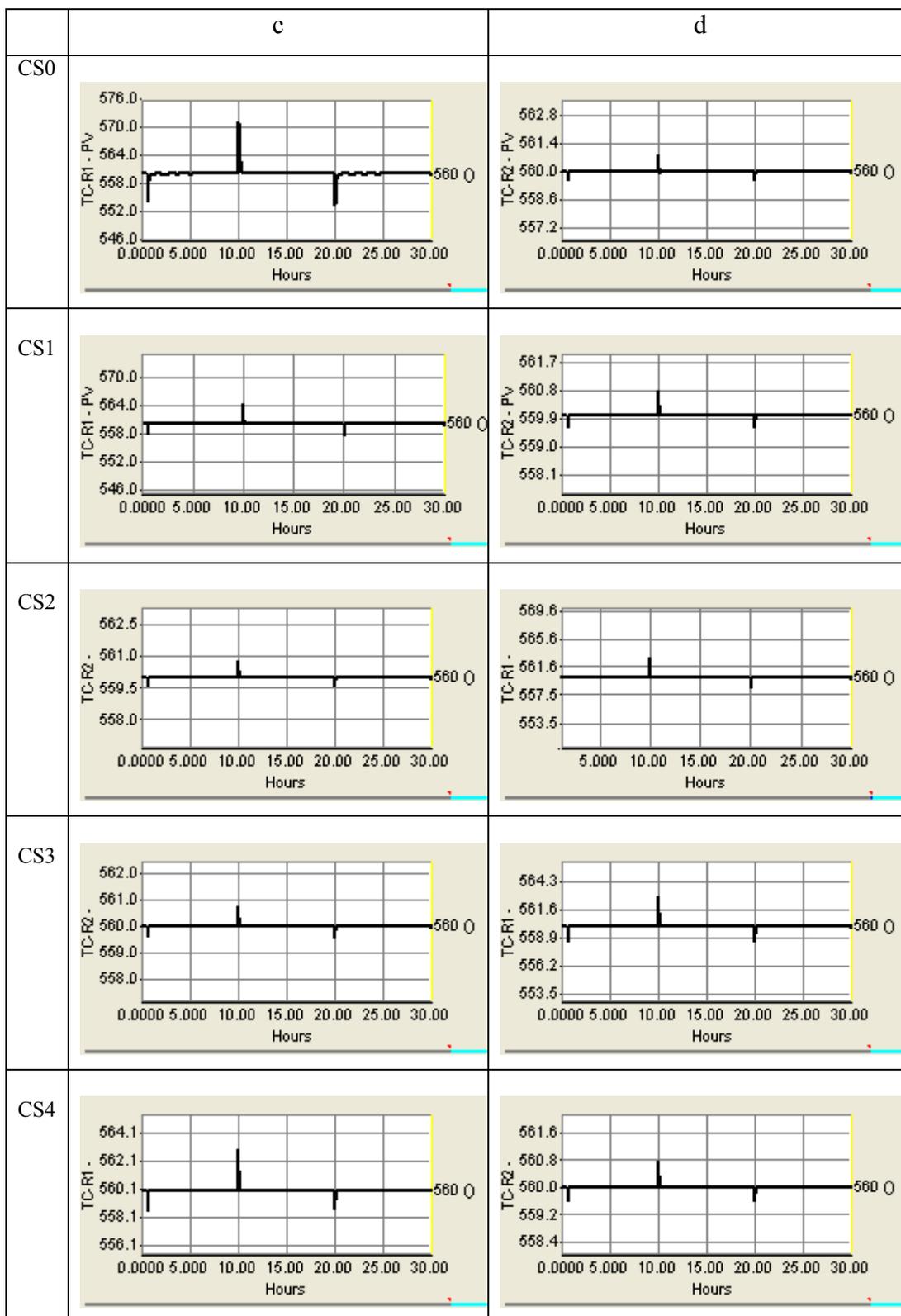


Figure 5.15A (Continued) Dynamic responses reaction section when changes in ratio between the steam feds to furnace E2 and total EB: (c) temperature inlet to PBR1 reactor, (d) temperature inlet to PBR2 reactor.

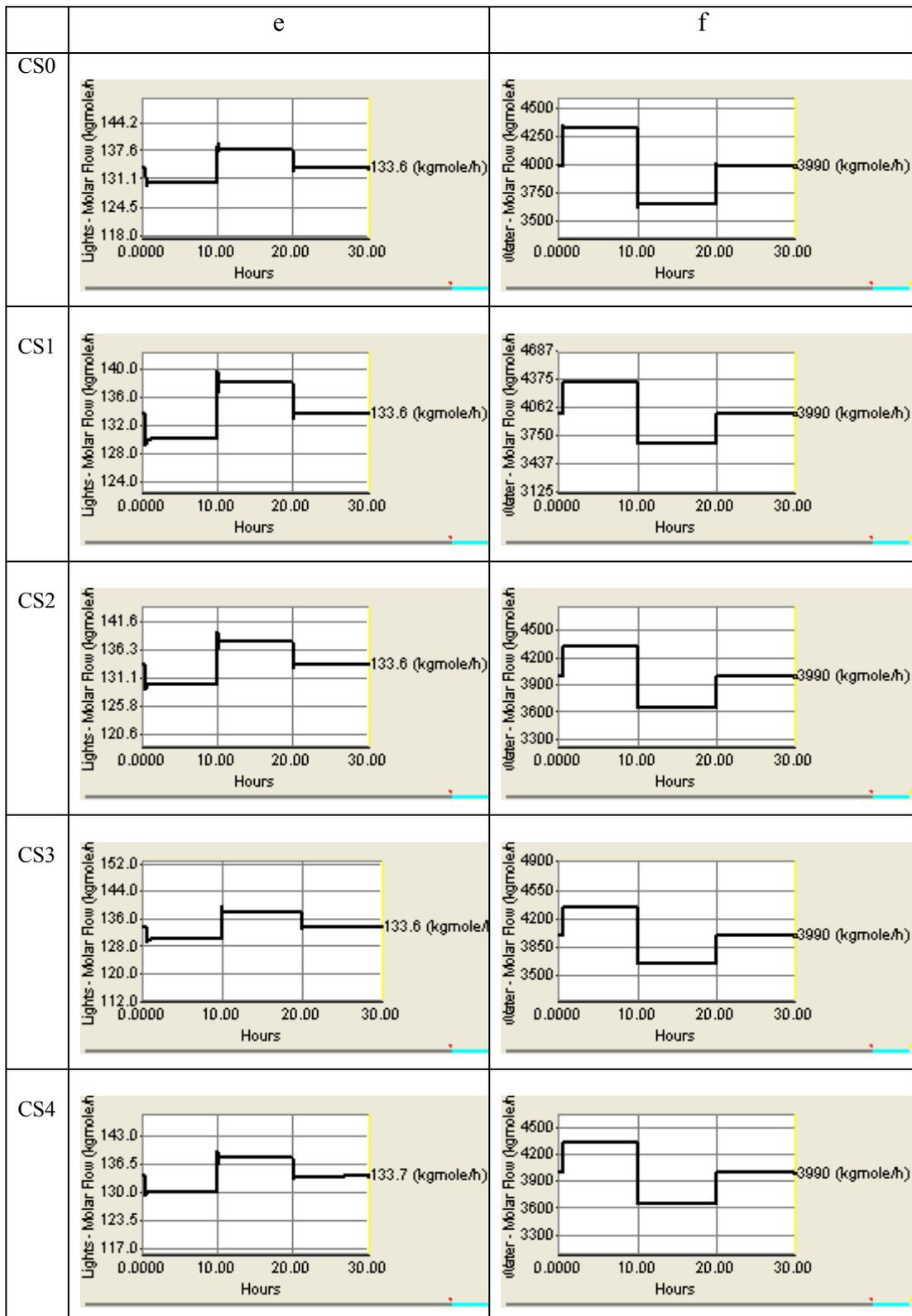


Figure 5.15A (Continued) Dynamic responses reaction section when changes in ratio between the steam feeds to furnace E2 and total EB: (e) light molar flow, and (f) water molar flow at 3-phase separator.

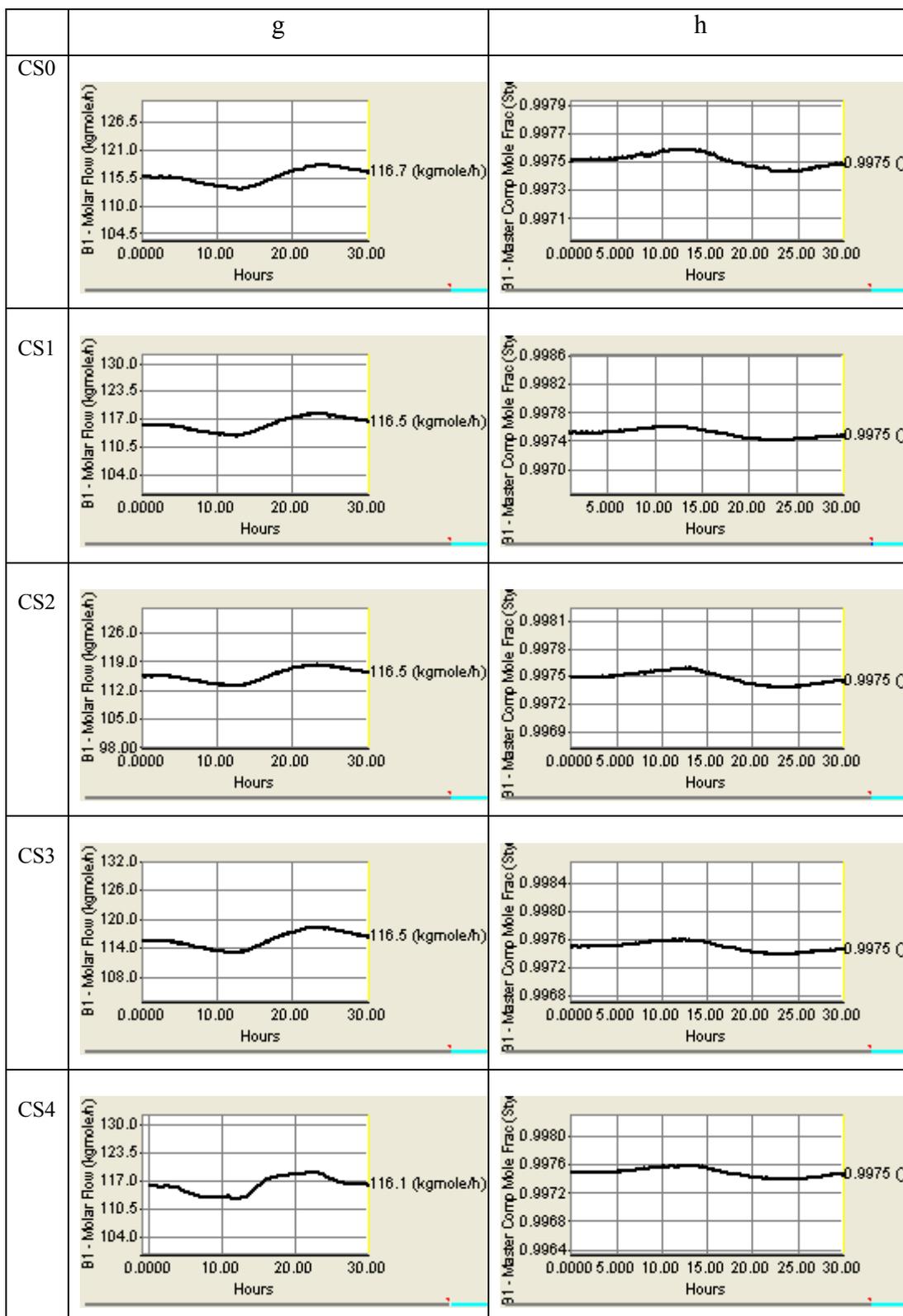


Figure 5.15B Dynamic responses of product column when changes in ratio between the steam feeds to furnace E2 and total EB: (g) bottom molar flow, and (h) styrene composition at bottom of product column.

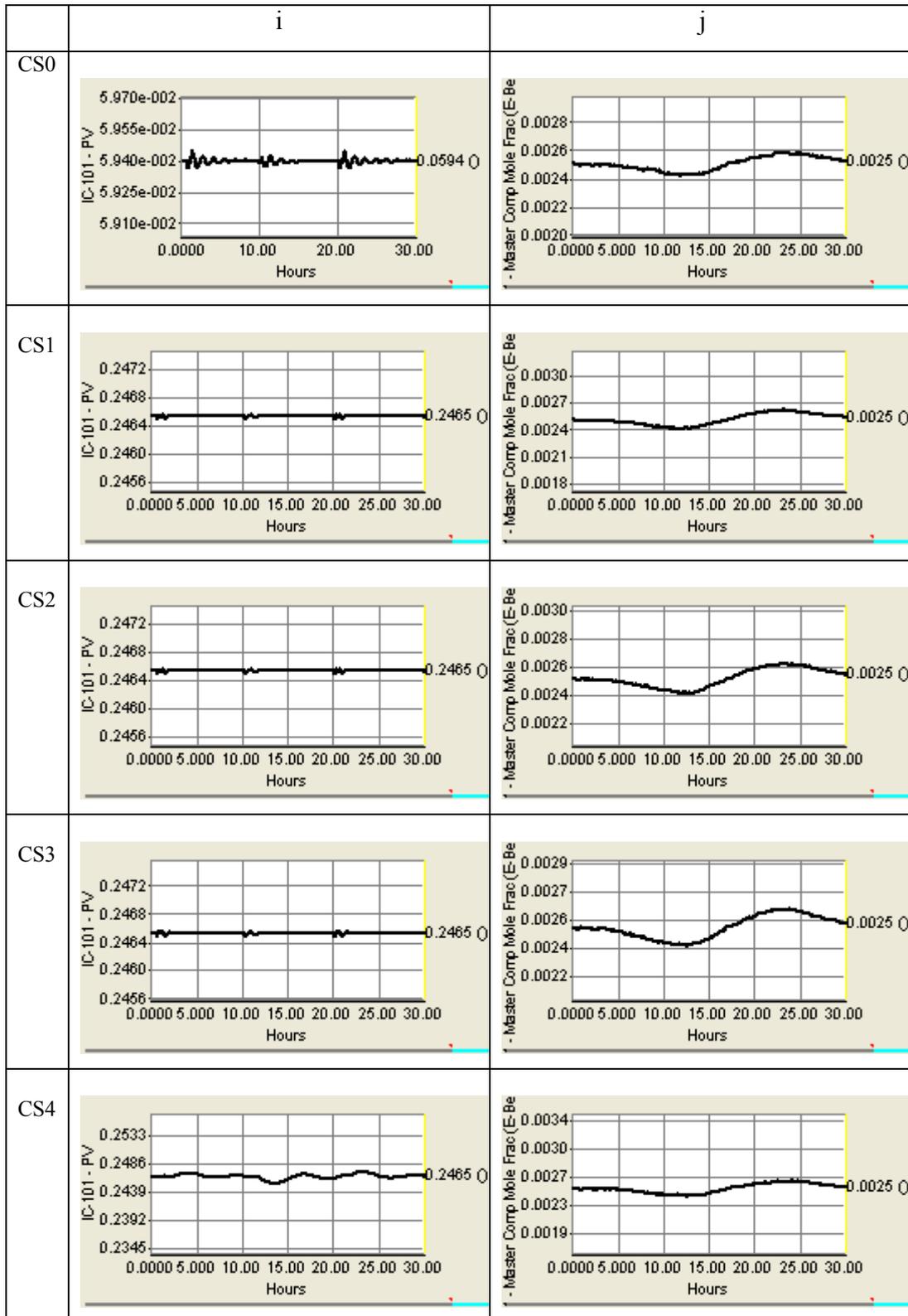


Figure 5.15B (Continued) Dynamic responses of product column when change in ratio between the steam feeds to furnace E2 and total EB: (i) EB composition on tray, and (j) EB composition at bottom of product column.

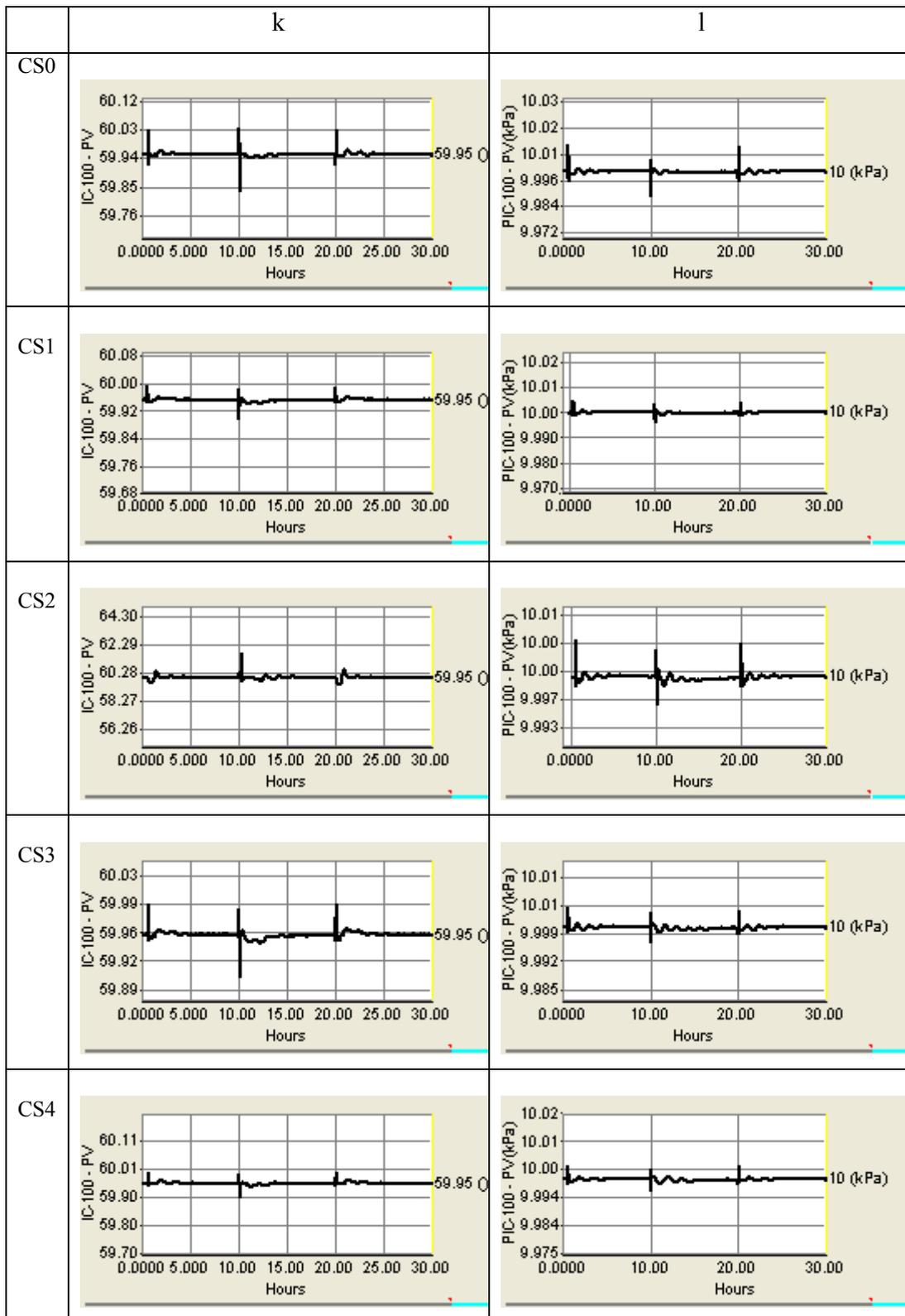


Figure 5.15B (Continued) Dynamic responses of product column when change in ratio between the steam feds to furnace E2 and total EB: (k) top product column temperature, (l) product column- condenser pressure.

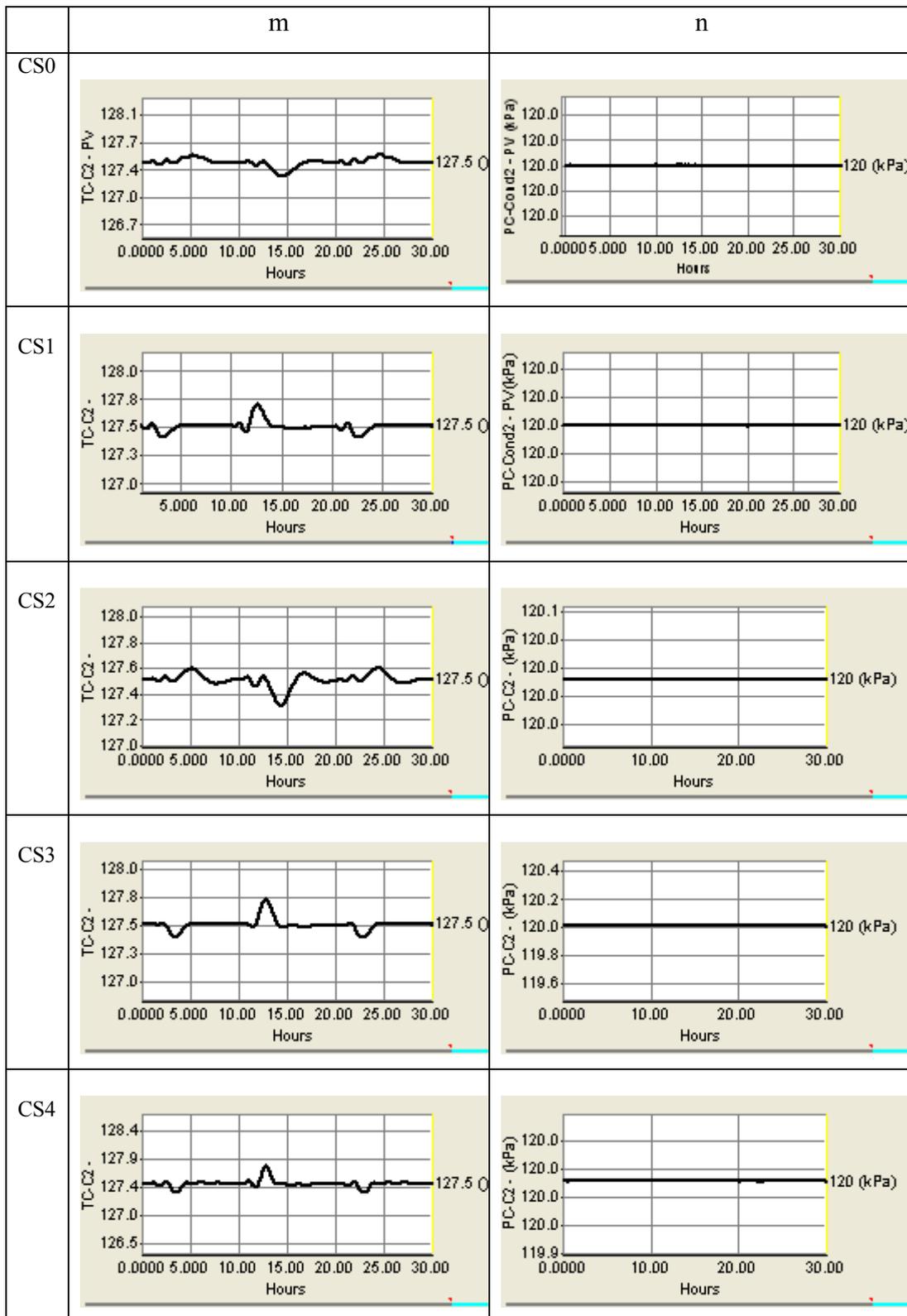


Figure 5.15C Dynamic responses of recycle column when change in ratio between the steam feds to furnace E2 and total EB: (m) temperature on Tray 7th in recycle column, (n) recycle column- condenser pressure.

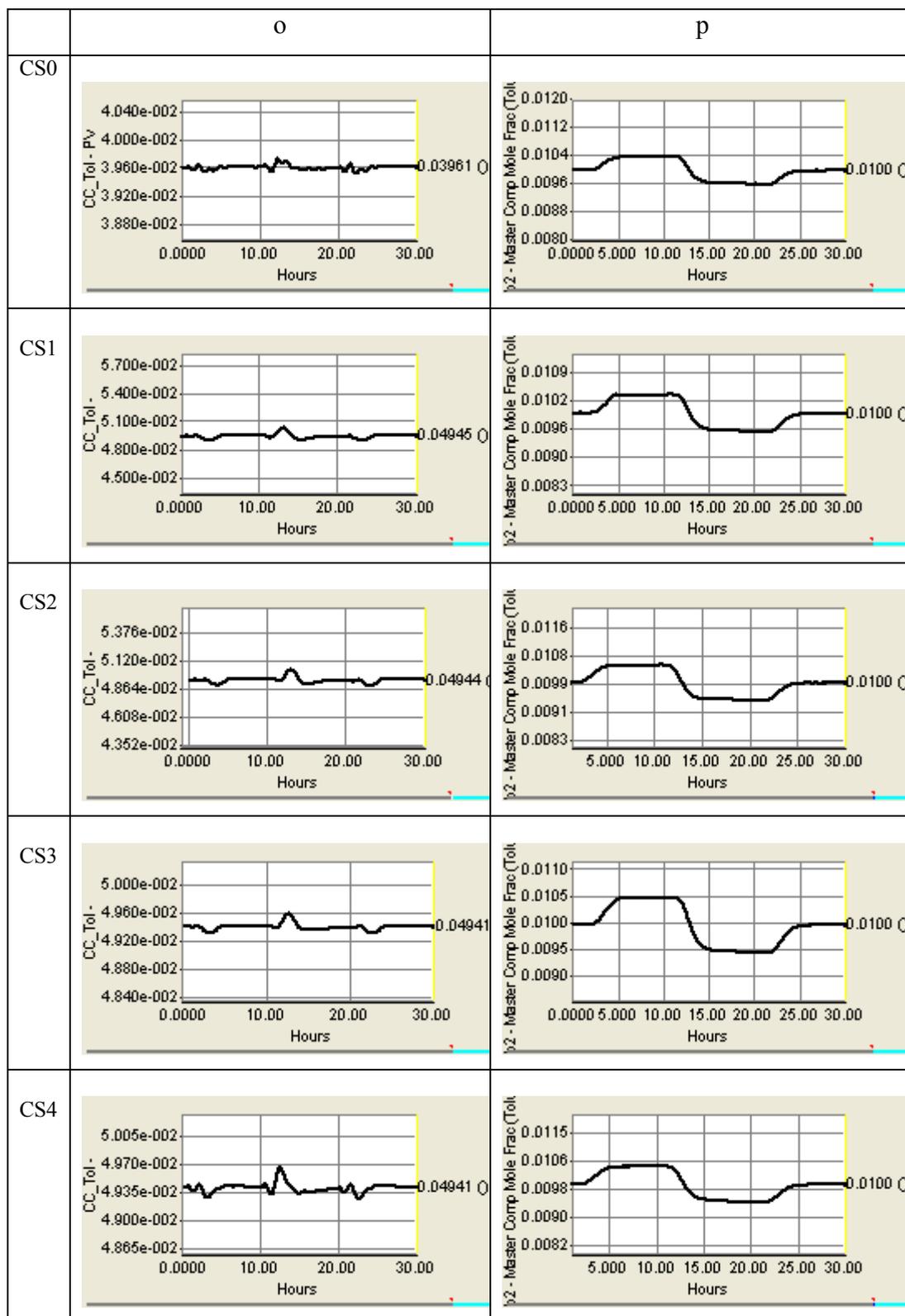


Figure 5.15C (Continued) Dynamic responses of recycle column when change in ratio between the steam feeds to furnace E2 and total EB: (o) toluene composition on tray, and (p) toluene composition at bottom of recycle column.

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:

$$\text{IAE} = \int |\varepsilon(t)| dt \quad (5.1)$$

Note that $\varepsilon(t) = y_{sp}(t) - y(t)$ is the deviation (error) of the response from the desired setpoint.

In this research, IAE method is used to evaluate the dynamic performance of the base case control structure and designed control structures. In the process, have many types of variables (temperature, pressure, and molar flow rate) 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 are used to evaluate the dynamic performance of the base case control structure and designed control structures.

For changing in material disturbances of total EB molar flow and ratio between steam feeds to furnace and total EB for base case control structure (CS0) and designed control structures (CS1-CS4). The IAE results for handle disturbances and maintain product quality are shown in Table 5.7, 5.8, 5.10, and Table 5.11, respectively.

For changing in thermal disturbances of temperature inlet to PBR1 and PBR2 reactors for base case control structure (CS0) and designed control structures (CS1-CS4). The IAE results for handle disturbances and maintain product quality are shown in Table 5.9 and Table 5.12.

For performance in used energy, the summation value of all energy that used in management disturbance is evaluated to find the designed control structure which is the most minimize energy use. Table 5.13 to Table 5.15 show the summation values of all energy use.

Table 5.7 The IAE result for handle disturbances to changes in total EB flow rate.

Controller	Integral Absolute Error (IAE)				
	CS0	CS1	CS2	CS3	CS4
TC-R1	0.0087	0.0009	0.0009	0.0007	0.0009
TC-R2	0.0002	0.0002	0.0002	0.0002	0.0002
TC-sep	0.0006	0.0004	0.0005	0.0004	0.0005
PC-sep	0.0030	0.0027	0.0027	0.0026	0.0028
TC-C1	0.0081	0.0041	0.0037	0.0012	0.0035
TC-C2	0.0163	0.0157	0.0136	0.0045	0.0196
PC-C1	0.0037	0.0017	0.0016	0.0006	0.0019
PC-C2	0.0000	0.0000	0.0000	0.0000	0.0000
CC-EB@C1	0.0019	0.0001	0.0001	0.0000	0.0955
CC-Tol@C2	0.0003	0.0014	0.0122	0.0005	0.0014
SUM	0.0427	0.0272	0.0355	0.0108	0.1262

Table 5.8 The IAE result for handle disturbances to changes in between LP steam feds to furnace and total EB ratio.

Controller	Integral Absolute Error (IAE)				
	CS0	CS1	CS2	CS3	CS4
TC-R1	0.0038	0.0006	0.0008	0.0006	0.0006
TC-R2	0.0002	0.0002	0.0002	0.0002	0.0002
TC-sep	0.0004	0.0005	0.0005	0.0005	0.0005
PC-sep	0.0025	0.0023	0.0023	0.0023	0.0023
TC-C1	0.00065	0.00058	0.00052	0.00051	0.00046
TC-C2	0.00378	0.00248	0.00241	0.00219	0.00235
PC-C1	0.0003	0.0002	0.0002	0.0001	0.0002
PC-C2	0.0000	0.0000	0.0000	0.0000	0.0000
CC-EB@C1	0.00013	0.00002	0.00003	0.00011	0.00071
CC-Tol@C2	0.0004	0.0002	0.0023	0.0002	0.0002
SUM	0.0121	0.0071	0.0091	0.0067	0.0075

Table 5.9 The IAE result for handle disturbances to changes in temperature inlet two reactors.

Controller	Integral Absolute Error (IAE)				
	CS0	CS1	CS2	CS3	CS4
TC-R1	1.2639	1.2647	1.2647	1.2647	1.2647
TC-R2	1.2646	1.2646	1.2646	1.2646	1.2646
TC-sep	0.0000	0.0000	0.0000	0.0000	0.0000
PC-sep	0.0006	0.0005	0.0005	0.0005	0.0005
TC-C1	0.00015	0.00026	0.00016	0.00021	0.00015
TC-C2	0.0004	0.0001	0.0004	0.0002	0.0002
PC-C1	0.0000	0.0001	0.0001	0.0000	0.0001
PC-C2	0.0000	0.0000	0.0000	0.0000	0.0000
CC-EB@C1	0.0000	0.0000	0.0000	0.0000	0.0001
CC-Tol@C2	0.0000	0.0000	0.0000	0.0000	0.0001
SUM	2.5297	2.5303	2.5306	2.5302	2.5304

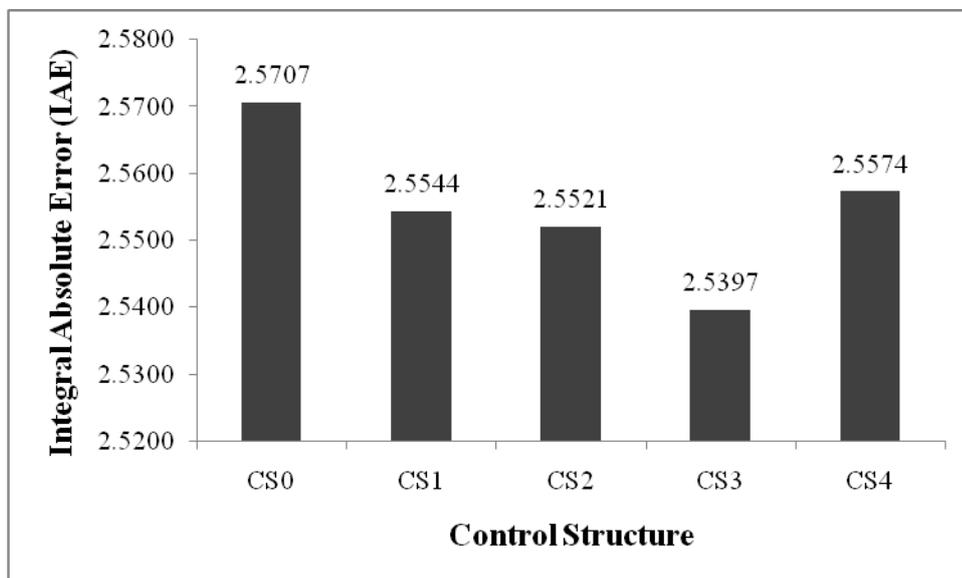


Figure 5.16 IAE for handle disturbances to changes in all disturbances testing of the temperature controllers.

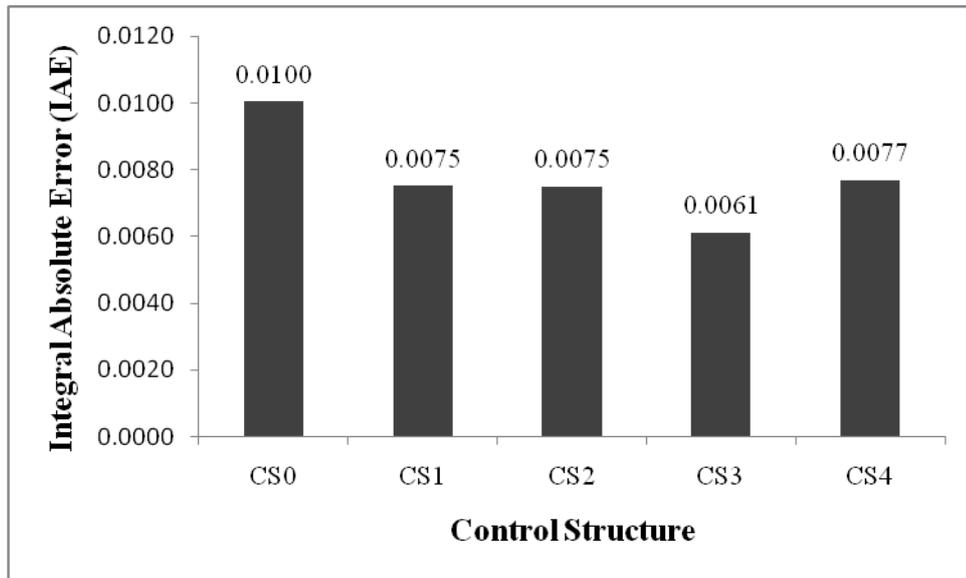


Figure 5.17 IAE for handle disturbances to changes in all disturbances testing of the pressure controllers.

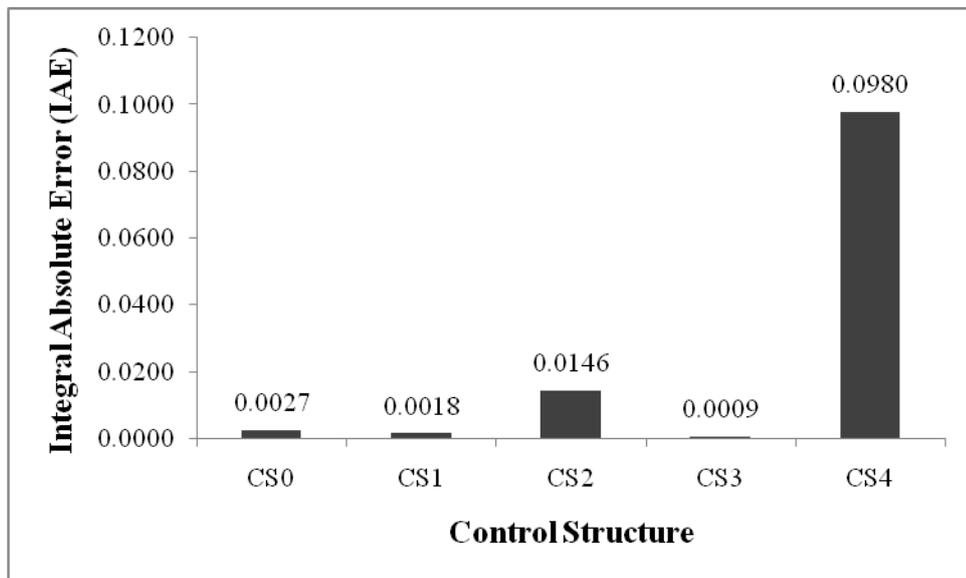


Figure 5.18 IAE for handle disturbances to changes in all disturbances testing of the composition controllers.

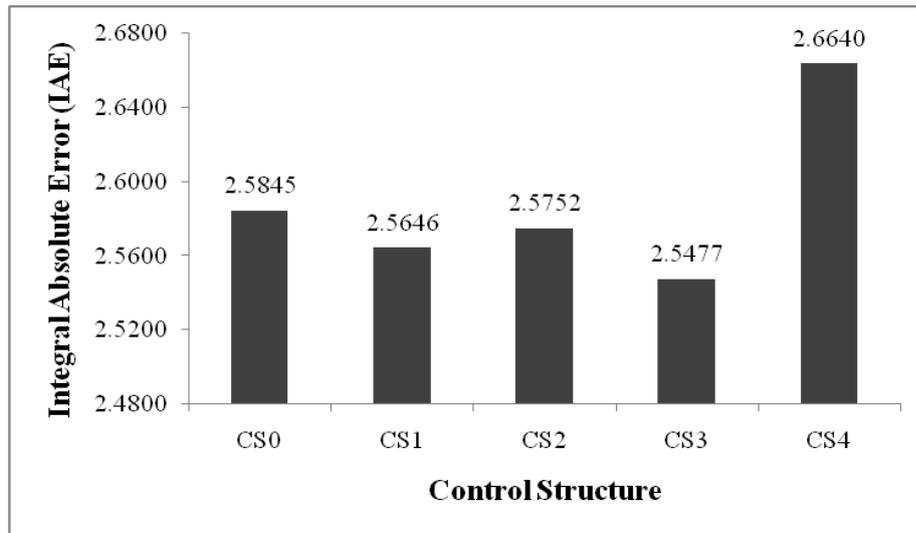


Figure 5.19 IAE for handle disturbances of all controllers to changes in all disturbances testing

For all disturbances testing, Figure 5.16, 5.17 and Figure 5.18 show the results of Integral Absolute Error (IAE) of temperature controller, pressure controller, and composition controller, respectively for handle disturbances. As for, Figure 5.18 shows the result IAE of complete controllers for handle disturbances. See that the control structure III (CS3) can handle disturbances and the best. Control structure I (CS1) is the second and control structure II and (CS2) is the third. From this result, if the designed control structure controls the product column with reflux flow fixed, it can handle all disturbances well.

Table 5.10 The IAE result of product composition variation to changes in total EB flow rate.

PRODUCT	Integral Absolute Error (IAE)				
	CS0	CS1	CS2	CS3	CS4
xSM@B1	0.0056	0.0065	0.0065	0.0033	0.0032
xSM@D1	0.0120	0.0120	0.0120	0.0051	0.0000

Table 5.11 The IAE result of product composition variation to changes in between LP steam feds to furnace and total EB ratio.

PRODUCT	Integral Absolute Error (IAE)				
	CS0	CS1	CS2	CS3	CS4
xSM@B1	0.0002	0.0003	0.0003	0.0005	0.0000
xSM@D1	0.0018	0.0017	0.0017	0.0028	0.0000

Table 5.12 The IAE result of product composition variation to changes in temperature inlet two reactors.

PRODUCT	Integral Absolute Error (IAE)				
	CS0	CS1	CS2	CS3	CS4
xSM@B1	0.0000	0.0000	0.0000	0.0001	0.0000
xSM@D1	0.0001	0.0001	0.0001	0.0001	0.0000

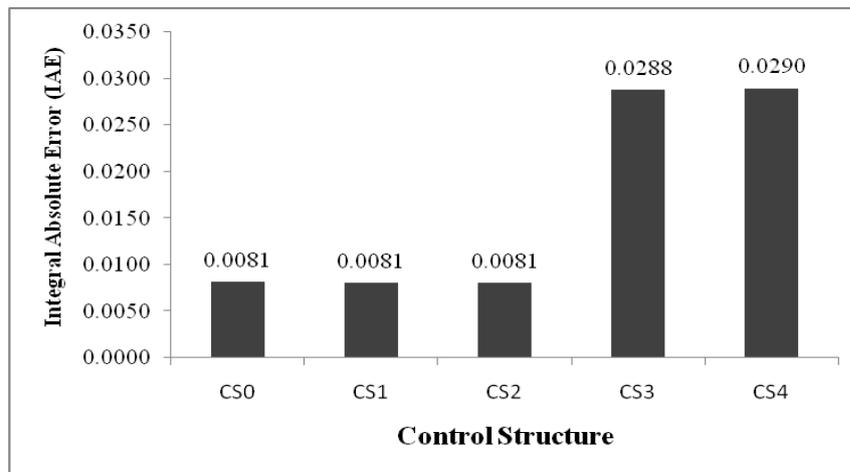


Figure 5.20 IAE of product composition variation to changes in all disturbances testing.

For all disturbances testing, Figure 5.20 shows the results of Integral Absolute Error (IAE) of all control structure product composition variation. The control structure (CS0 to CS3) can maintain product quality the best. From this result, if the designed control structure controls the Tray 54th ethylbenzene composition of product column by manipulating the DIB column reboiler duty (qr1) and the feed to reflux ratio structure, it can maintain product quality well

Table 5.13 The summation value of all energy used when changes total EB molar flow.

Energy	Summation Value (GJ/h)				
	CS0	CS1	CS2	CS3	CS4
q1	5.340E+07	1.499E+07	1.515E+07	1.50E+07	1.51E+07
q3	8.163E+06	6.331E+06	6.168E+06	6.16E+06	6.17E+06
q4	2.488E+08	2.141E+08	2.119E+08	2.07E+08	2.07E+08
qC1	1.051E+08	1.004E+08	1.003E+08	1.10E+08	1.11E+08
qC2	3.184E+06	2.783E+06	4.140E+06	3.04E+06	3.06E+06
qr1	3.853E+07	3.857E+07	3.853E+07	1.13E+07	1.14E+07
qr2	3.100E+06	5.827E+06	5.797E+06	7.56E+06	7.60E+06
kp1	6.288E+04	5.613E+04	5.659E+04	9.19E+03	2.30E+04
SUM	4.603E+08	3.830E+08	3.820E+08	3.591E+08	3.615E+08

Table 5.14 The summation value of all energy used when changes between steam feeds to furnace and total EB ratio.

Energy	Integral Absolute Error (IAE)				
	CS0	CS1	CS2	CS3	CS4
q1	8.021E+07	2.164E+07	2.247E+07	2.29E+07	2.28E+07
q3	9.078E+06	7.026E+06	7.043E+06	7.04E+06	7.04E+06
q4	3.886E+08	3.311E+08	3.340E+08	3.21E+08	3.21E+08
qC1	1.300E+07	1.429E+07	1.421E+07	1.42E+07	1.41E+07
qC2	2.269E+06	2.154E+06	1.716E+06	2.14E+06	2.14E+06
qr1	1.534E+06	1.574E+06	1.577E+06	1.36E+06	1.31E+06
qr2	1.006E+06	9.469E+05	9.385E+05	9.21E+05	9.26E+05
kp1	6.171E+04	6.265E+04	6.245E+04	6.30E+04	6.30E+04
SUM	4.958E+08	3.788E+08	3.821E+08	3.695E+08	3.694E+08

Table 5.15 The summation value of all energy used when changes temperature inlet of two reactors.

Energy	Integral Absolute Error (IAE)				
	CS0	CS1	CS2	CS3	CS4
q1	4.52E+07	4.54E+07	4.53E+07	4.51E+07	4.51E+07
q3	4.53E+07	4.17E+07	4.15E+07	4.14E+07	4.14E+07
q4	6.40E+04	2.12E+04	2.36E+04	1.98E+04	1.46E+04
qC1	1.02E+06	1.31E+06	1.50E+06	1.43E+06	1.15E+06
qC2	1.27E+05	1.37E+05	1.70E+05	1.29E+05	1.27E+05
qr1	1.76E+05	3.21E+05	3.65E+05	4.47E+05	1.21E+05
qr2	3.32E+04	9.42E+04	4.37E+04	1.08E+05	2.60E+04
kp1	9.32E+03	9.61E+03	8.01E+03	9.70E+03	9.66E+03
SUM	9.19E+07	8.90E+07	8.89E+07	8.87E+07	8.80E+07

For all disturbances testing, Figure 5.17 shows the summation value for all energy used of all control structure. Control structure II (CS3) is the most minimize energy used. Control structure IV (CS4) is the second. Control structure II (CS2) is the third.

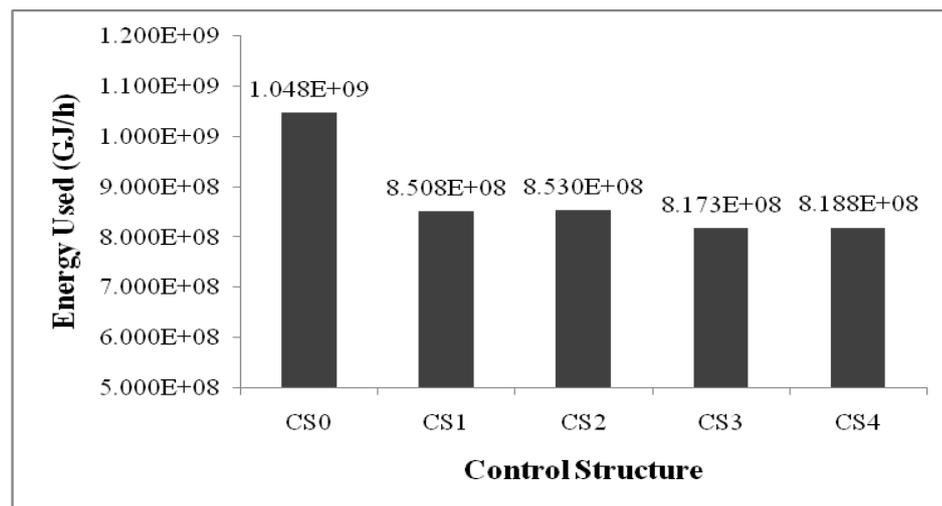


Figure 5.21 The summation value of all energy used when changes all disturbances testing.

The result of utility costs for each control structure to the change in all disturbances testing is shown in Figure 5.18.

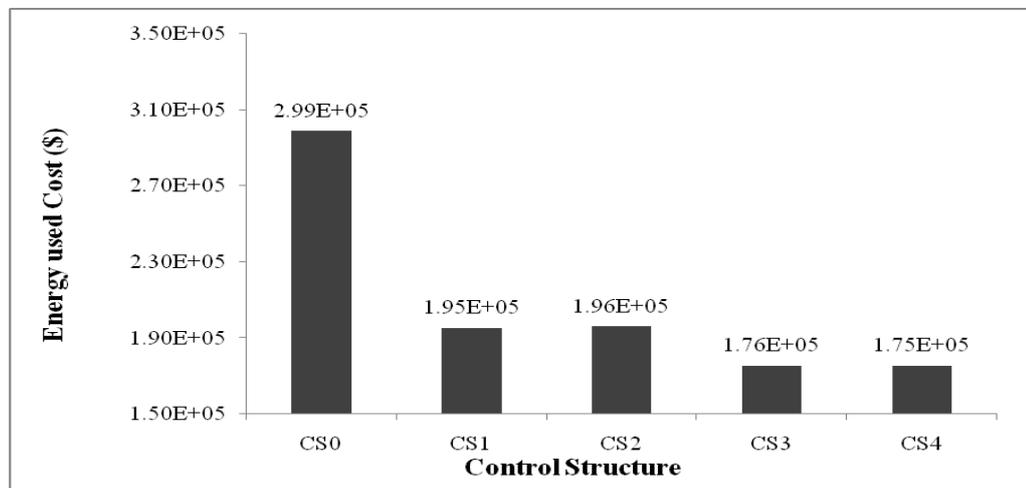


Figure 5.22 Utility costs for each control structure to the change in all disturbances testing.

CHAPTER VI

CONCLUTIONS AND RECOMMENDATIONS

6.1 Conclusion

The plantwide control problem is to develop a control structure for a complex and integrated process that satisfies the plant's design objectives. The cost of process is one of the important design plants. Therefore, this work developed the styrene process from research of Luyben (2011). That heat exchanger is designed to reduce energy in furnace. A 39.94 % reduction in utilities costs is achieved (5,190.52 K\$ per year) for a 1.29 % decrease in capital investment (from 8679.48 K\$ to 8,567.58 K\$) from the base case control structure based on evaluation of economic costs

At dynamic simulation, this work has discussed control structure design for styrene process, using new design procedure of Wongsri (2009). This procedure based on heuristics analysis. The precedence 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 and maintain product quality. 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 all disturbances, designed control structure III (CS3) is the best structure to handle disturbances. The base case and designed control structures (CS0, CS1, CS2, and CS4) with fixed reflux to feed ratios control structure for product column to maintain product quality shows that (CS1) can maintain product quality better than others structures. As the designed control structure IV (CS4) is the most minimize energy used.

From assessment of all control structures found that, single -end control of distillation column can be effective and for selection Reflux to Feed ratio scheme at product column. This ratio structure ability to handle feed flow rate disturbance and maintain styrene product quality as desired. In recycle column, tray 7th temperature is controlled by manipulating direct reflux flow. This control loop reduces to the interaction of control structures, and leads to faster response than manipulating with Reflux ratio. As a result, the designed control structure I (CS1) appropriate for the styrene process

New design procedure of Wongsri (2009) can 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, and dynamic simulation, a useful design procedure that leads to a good-performance plantwide control system.

6.2 Recommendations

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 Process equipment data

Unit operations	property	Size
Reactors (two PFRs)	Diameter (m)	3.3 m.
	Length (m)	8.0 m.
	Void fraction	0.44
3-phase separator	Volume (m ³)	24.06 m ³
FEHE	UA (kJ/h-°C)	3.06×10 ⁵
	Hot outlet temperature (°C)	500

Table A.2 Specifications for the distillation columns

Specification	1 st column	2 nd column
Number of tray	80	35
Feed to tray	35	15
Condenser P(kPa)	10	120
Reboiler P (kPa)	50	138
Reflux Ratio	5.05	10.60
% of bottom	0.0025 (EB)	0.01(ST)
% of distillate	0.01 (EB)	0.01(Tol)
Diameter (m)	4.57	1.38
Condenser volume (m ²)	19.43	2.38
Reboiler volume (m ²)	24.49	7.88

Table A.3 (Continuous) Data stream of styrene process

Name	C_heated	LPs2_heated	r2cooled	Feed1	FPR1out	feed2	PFR2_out	PFR2_rec	to_sep
Vapor Fraction	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0303
Temperature (°C)	500.00	605.76	392.40	560.00	523.50	560.00	537.09	537.07	40.00
Pressure (kPa)	270	270	150	270	240	210	180	180	120.000004
Molar Flow (kmol/h)	890.71	3387.99	4398.89	4278.70	4351.54	4351.54	4398.81	4398.89	4398.89
Mass Flow (kg/h)	40580.66	61034.95	101625.54	101615.61	101616.09	101616.09	101616.65	101625.54	101625.54
Comp Mole frac									
Ethylbenzene	0.3091	0.0000	0.0338	0.0643	0.0458	0.0458	0.0338	0.0338	0.0338
Styrene	0.0017	0.0000	0.0267	0.0004	0.0167	0.0167	0.0267	0.0267	0.0267
Hydrogen	0.0000	0.0000	0.0253	0.0000	0.0157	0.0157	0.0253	0.0253	0.0253
H2O	0.6871	1.0000	0.9089	0.9349	0.9191	0.9191	0.9089	0.9089	0.9089
Benzene	0.0004	0.0000	0.0006	0.0001	0.0004	0.0004	0.0006	0.0006	0.0006
Toluene	0.0016	0.0000	0.0022	0.0003	0.0012	0.0012	0.0022	0.0022	0.0022
Ethylene	0.0000	0.0000	0.0005	0.0000	0.0003	0.0003	0.0005	0.0005	0.0005
Methane	0.0000	0.0000	0.0017	0.0000	0.0008	0.0008	0.0017	0.0017	0.0017
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	0.0000	0.0000	0.0002	0.0000	0.0001	0.0001	0.0002	0.0002	0.0002

Table A.3 (Continuous) Data stream of styrene process

Name	C_heated	LPS2_heated	r2cooled	Feed1	FPR1out	feed2	PFR2_out	PFR2_rec	to_sep
Vapor Fraction	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0303
Temperature (°C)	500.00	605.76	392.40	560.00	523.50	560.00	537.09	537.07	40.00
Pressure (kPa)	270	270	150	270	240	210	180	180	120.00004
Molar Flow (kmol/h)	890.71	3387.99	4398.89	4278.70	4351.54	4351.54	4398.81	4398.89	4398.89
Mass Flow (kg/h)	40580.66	61034.95	101625.54	101615.61	101616.09	101616.09	101616.65	101625.54	101625.54
Comp Mole frac									
Ethylbenzene	0.3091	0.0000	0.0338	0.0643	0.0458	0.0458	0.0338	0.0338	0.0338
Styrene	0.0017	0.0000	0.0267	0.0004	0.0167	0.0167	0.0267	0.0267	0.0267
Hydrogen	0.0000	0.0000	0.0253	0.0000	0.0157	0.0157	0.0253	0.0253	0.0253
H2O	0.6871	1.0000	0.9089	0.9349	0.9191	0.9191	0.9089	0.9089	0.9089
Benzene	0.0004	0.0000	0.0006	0.0001	0.0004	0.0004	0.0006	0.0006	0.0006
Toluene	0.0016	0.0000	0.0022	0.0003	0.0012	0.0012	0.0022	0.0022	0.0022
Ethylene	0.0000	0.0000	0.0005	0.0000	0.0003	0.0003	0.0005	0.0005	0.0005
Methane	0.0000	0.0000	0.0017	0.0000	0.0008	0.0008	0.0017	0.0017	0.0017
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	0.0000	0.0000	0.0002	0.0000	0.0001	0.0001	0.0002	0.0002	0.0002

Table A.3 (Continuous) Data stream of styrene process

Name	organics	Light	water	Light out	F1	water out	L1	d1	b1	P-3out
Vapor Fraction	0.0000	1.0000	0.0000	1.0000	0.00	0.0000	1.0000	0.0000	0.0000	0.0000
Temperature (°C)	40.00	40.00	40.00	39.98	40.01	40.00	59.95	59.95	120.54	60.04
Pressure (kPa)	120	120	120	100	20.00	100	10	10	50	227
Molar Flow (kmol/h)	275.85	133.17	3989.87	133.17	275.85	3989.87	4.88	155.71	115.26	155.71
Mass Flow (kg/h)	28803.89	943.56	71878.09	943.56	28803.89	71878.09	457.48	16341.16	12005.25	16341.16
Comp Mole frac										
Ethylbenzene	0.5321	0.0139	0.0000	0.0139	0.5321	0.0000	0.7102	0.9186	0.0025	0.9186
Styrene	0.4225	0.0075	0.0000	0.0075	0.4225	0.0000	0.0055	0.0100	0.9975	0.0100
Hydrogen	0.0004	0.8333	0.0000	0.8333	0.0004	0.0000	0.0218	0.0000	0.0000	0.0000
H2O	0.0008	0.0612	1.0000	0.0612	0.0008	1.0000	0.0459	0.0000	0.0000	0.0000
Benzene	0.0091	0.0020	0.0000	0.0020	0.0091	0.0000	0.0754	0.0138	0.0000	0.0138
Toluene	0.0345	0.0024	0.0000	0.0024	0.0345	0.0000	0.1112	0.0575	0.0000	0.0575
Ethylene	0.0003	0.0174	0.0000	0.0174	0.0003	0.0000	0.0142	0.0000	0.0000	0.0000
Methane	0.0002	0.0558	0.0000	0.0558	0.0002	0.0000	0.0116	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	0.0001	0.0065	0.0000	0.0065	0.0001	0.0000	0.0042	0.0000	0.0000	0.0000

Table A.3 (Continuous) Data stream of styrene process

Name	vent	F2	P-4out	ST	d2	b2	P-5out	Tol/Ben	p-6out	EB_rec
Vapor Fraction	0.9841	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0117	0.0000	0.0000
Temperature (°C)	141.66	60.09	120.65	120.69	100.75	147.67	100.85	98.78	147.82	147.85
Pressure (kPa)	150	127	300	100	120	138	300	100	400	300
Molar Flow (kmol/h)	4.88	155.71	115.26	115.26	9.76	145.95	9.76	9.76	145.95	145.95
Mass Flow (kg/h)	457.48	16341.16	12005.25	12005.25	869.50	15471.67	869.50	869.50	15471.67	15471.67
Comp Mole frac										
Ethylbenzene	0.7102	0.9212	0.0025	0.0025	0.0100	0.9793	0.0100	0.0100	0.9793	0.9793
Styrene	0.0055	0.0100	0.9975	0.9975	0.0000	0.0107	0.0000	0.0000	0.0107	0.0107
Hydrogen	0.0218	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0459	0.0000	0.0000	0.0000	0.0007	0.0000	0.0007	0.0007	0.0000	0.0000
Benzene	0.0754	0.0131	0.0000	0.0000	0.2201	0.0000	0.2201	0.2201	0.0000	0.0000
Toluene	0.1112	0.0557	0.0000	0.0000	0.7688	0.0100	0.7688	0.7688	0.0100	0.0100
Ethylene	0.0142	0.0000	0.0000	0.0000	0.0002	0.0000	0.0002	0.0002	0.0000	0.0000
Methane	0.0116	0.0000	0.0000	0.0000	0.0001	0.0000	0.0001	0.0001	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	0.0042	0.0000	0.0000	0.0000	0.0001	0.0000	0.0001	0.0001	0.0000	0.0000

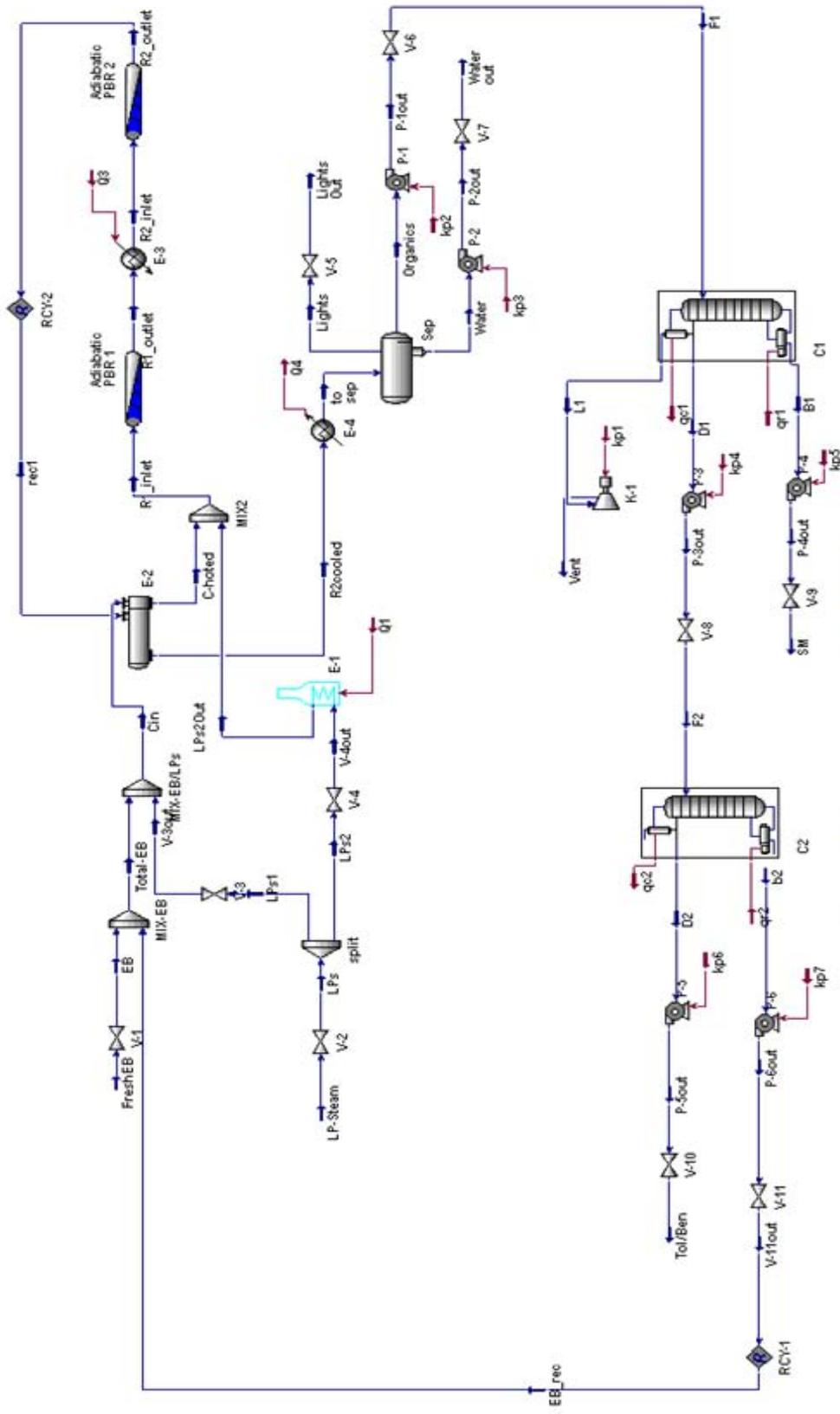


Figure A.1 Styrene process via HYSYS

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 and Pressure Loops

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 of $\tau_I = 0.3$ 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.

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 setpoint 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 $K_C = 2$. 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 flowrate 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 $\tau_I = 10$ minutes.

B.3 Relay- Feedback Testing

The relay-feedback test is a tool that serves 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 an 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 setpoint with the controller output switching every time the process variable (PV) signal crosses the setpoint. Figure B.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.

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

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 is 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, K_U and the ultimate period, P_U permits us to calculate controller settings. There are several methods that require only these two parameters. The Ziegler-Nichols tuning equations for a PI controller are:

$$K_C = K_U / 2.2 \quad (2)$$

$$\tau_I = P_U / 1.2 \quad (3)$$

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:

$$K_C = K_U / 3.2 \quad (4)$$

$$\tau_I = 2.2P_U \quad (5)$$

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.

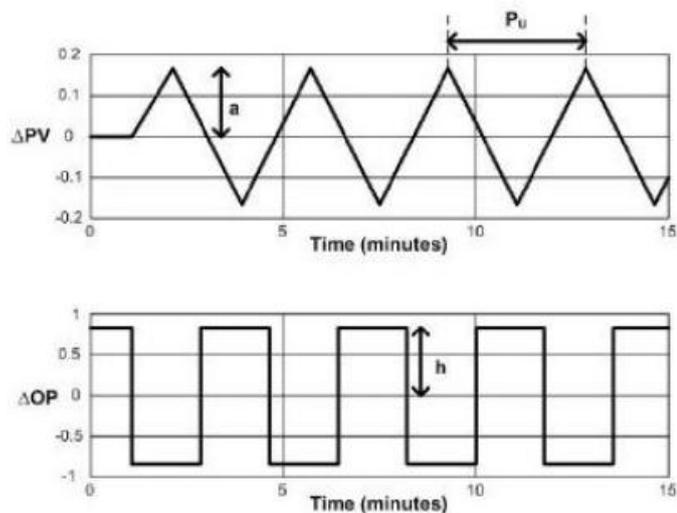


Figure B.1 Input and Output from Relay-Feedback Test (Luyben, W., Plantwide Dynamic Simulations in Chemical Processing and Control, p. 30, 1998.)

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

Table B.2 Same parameters tuning for styrene process in all control structures

Controller	Controlled Variables	Manipulated Variables	Action	Tunning Parameter		
				K _c	τ_I	τ_D
FC-TotalEB	Total EB (Fresh+Recycle) flow rate	V-1	Reverse	0.5	0.3	-
FC-LPs1	LPs1/Total EB flow rate ratio	V-3	Reverse	0.36	0.03	-
FC-LPs2	LPs2/Total EB flow rate ratio	V-4	Reverse	0.20	0.02	-
TC-R1	PFR1 inlet temperature	Q1	Reverse	1.15	1.41	0.3
TC-R2	PFR2 inlet temperature	Q3	Reverse	6.13	1.17	0.26
TC-Sep	3-phase separator temperature	Q4	Direct	0.20	0.12	-
PC-Sep	3-phase separator pressure	V-5	Direct	2.49	10	-
LC-Sep	3-phase separator liquid %level	V-6	Direct	19.80	0.51	-
LP-Aq	3-phase separator aqueous level	V-7	Direct	4.21	0.18	-
PC-Cond1	Condenser pressure of product column	qc1	Direct	2	10	-
LC-Cond1	Condenser level of product column	V-8	Direct	2	-	-
TC-L1	Gas top temperature of product column	wk1	Reverse	6.78	2.52	.56
PC-Cond2	Condenser pressure of recycle column	qc2	Direct	2	10	-
LC-Reb2	Reboiler level of recycle column	V-11	Direct	2	-	-
CC-Tol	Tray toluene composition of recycle column	qr2	Direct	1.70	10.6	2.36

Table B.3 Different parameter tuning parameters for styrene in all control structures.

Controller	Controlled Variables	Manipulated Variables	Action Controller	Tunning Parameter		
				Kc	τI	τD
CS1-CS3						
LC-Reb1	Reboiler level	V-9	Direct	2.00	-	-
CS4						
LC-Reb1	Reboiler level	qr1	Direct	5.16	-	-
CS1-CS3						
CC-EB	Tray EB composition in product column	qr1	Direct	2.88	10.50	2.34
CS4						
CC-EB	Tray EB composition in product column	V-9	Reverse	2.88	10.50	2.34
CS1-CS2,CS4						
R1/F1	Reflux/feed flow ratio	Reflux flow SP	Direct		spreadsheet	
FC-Reflux1	Reflux flow (cascade)	Reflux flow	Reverse	0.5	0.3	-
CS3						
FC-Reflux1	Reflux flow	Reflux flow	Reverse	0.5	0.3	-
CS1,CS3,CS4						
LC-Cond2	Condenser level of recycle column	V-10	Direct	2	-	-
CS2						
TC-S7	Tray 7 th Temperature of recycle column	distillate/reflux SP	Direct	3.02	11.70	2.60
FC-Reflux2	distillate/reflux flow	distillate flow	Reverse	0.21	0.008	-
CS1,CS3,CS4						
TC-S7	Tray 7 th Temperature of recycle column	reflux flow	Reverse	3.88	13.20	2.93

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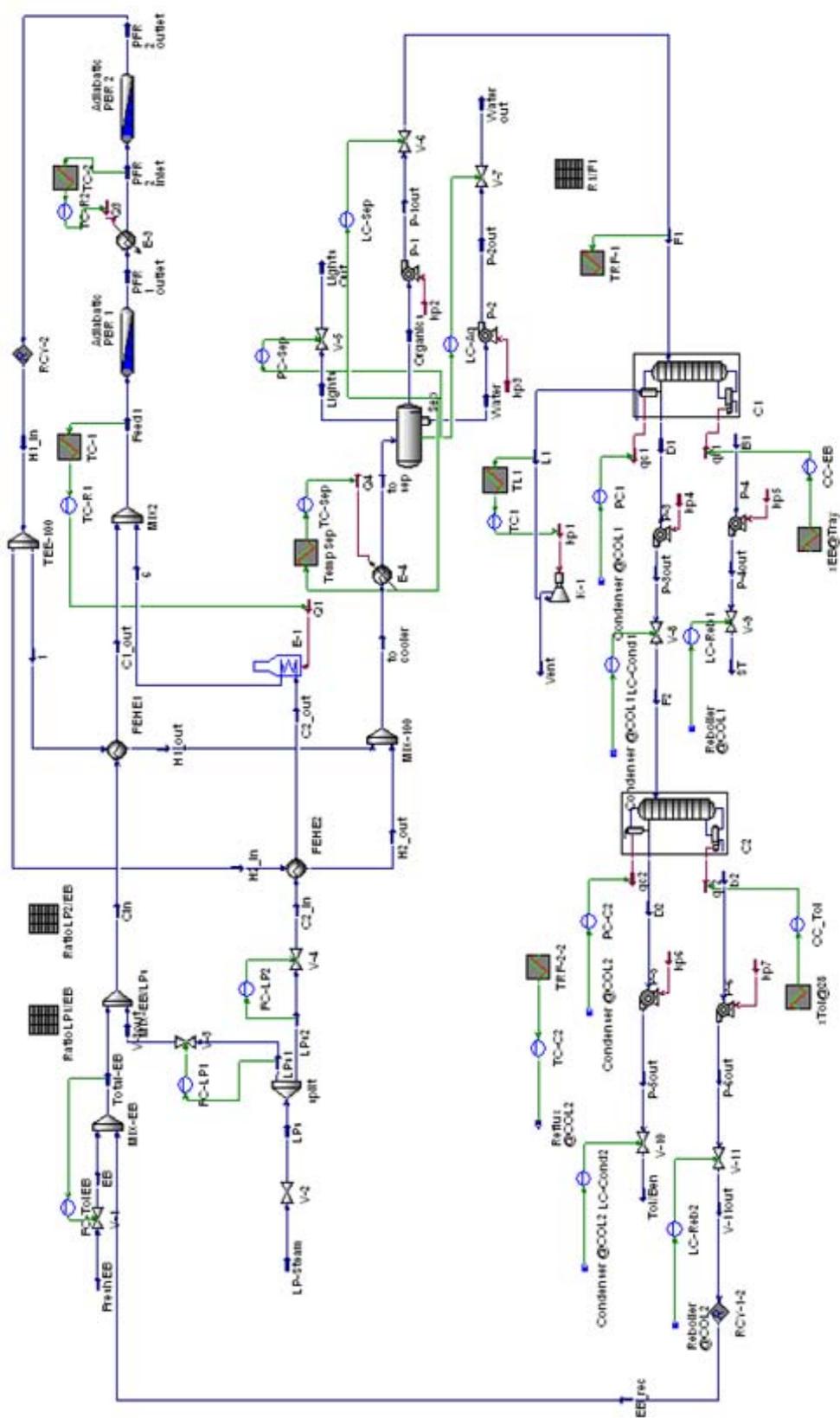


Figure B.2 Designed control structure I (CS1) for styrene process

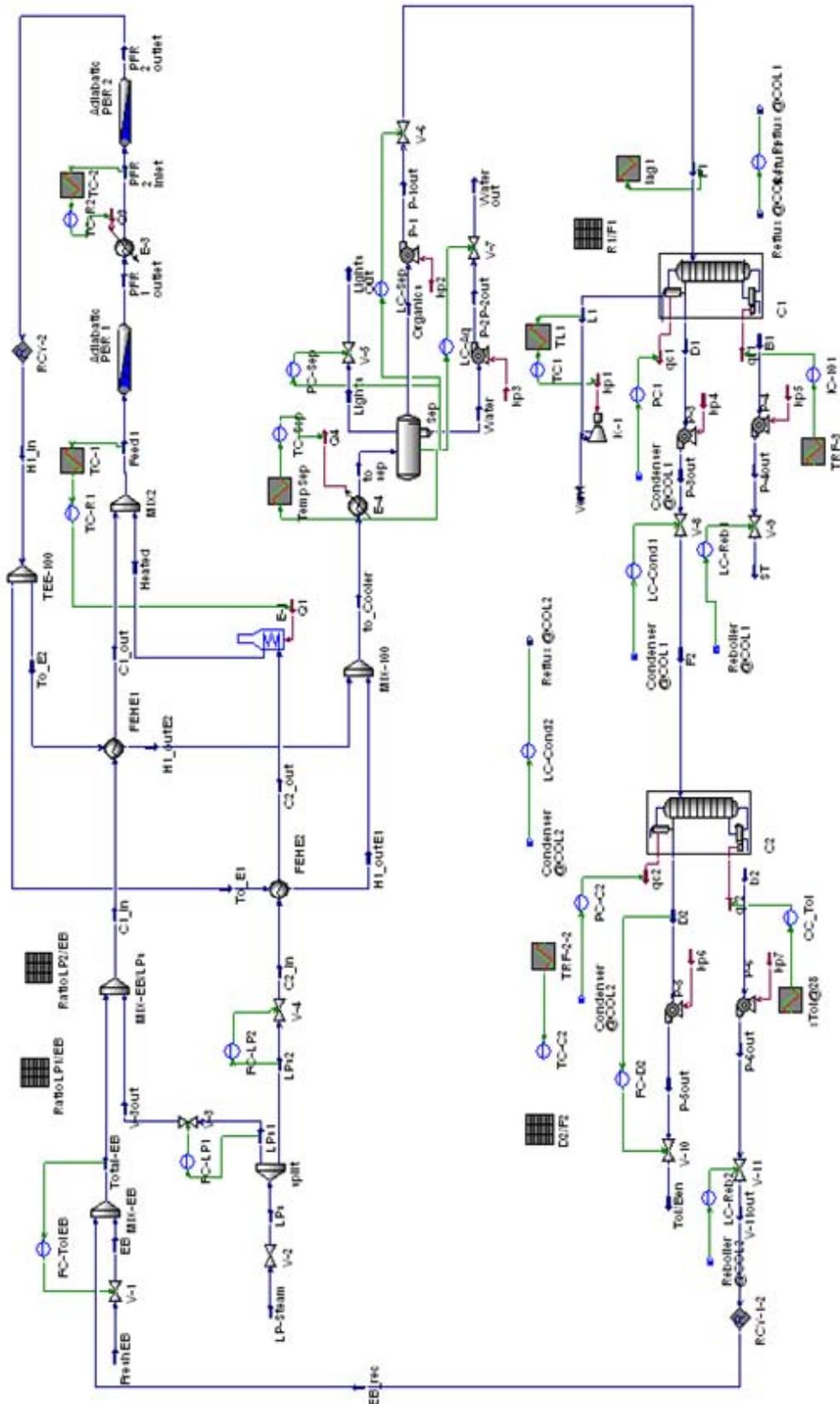


Figure B.3 Designed control structure II (CS2) for styrene process

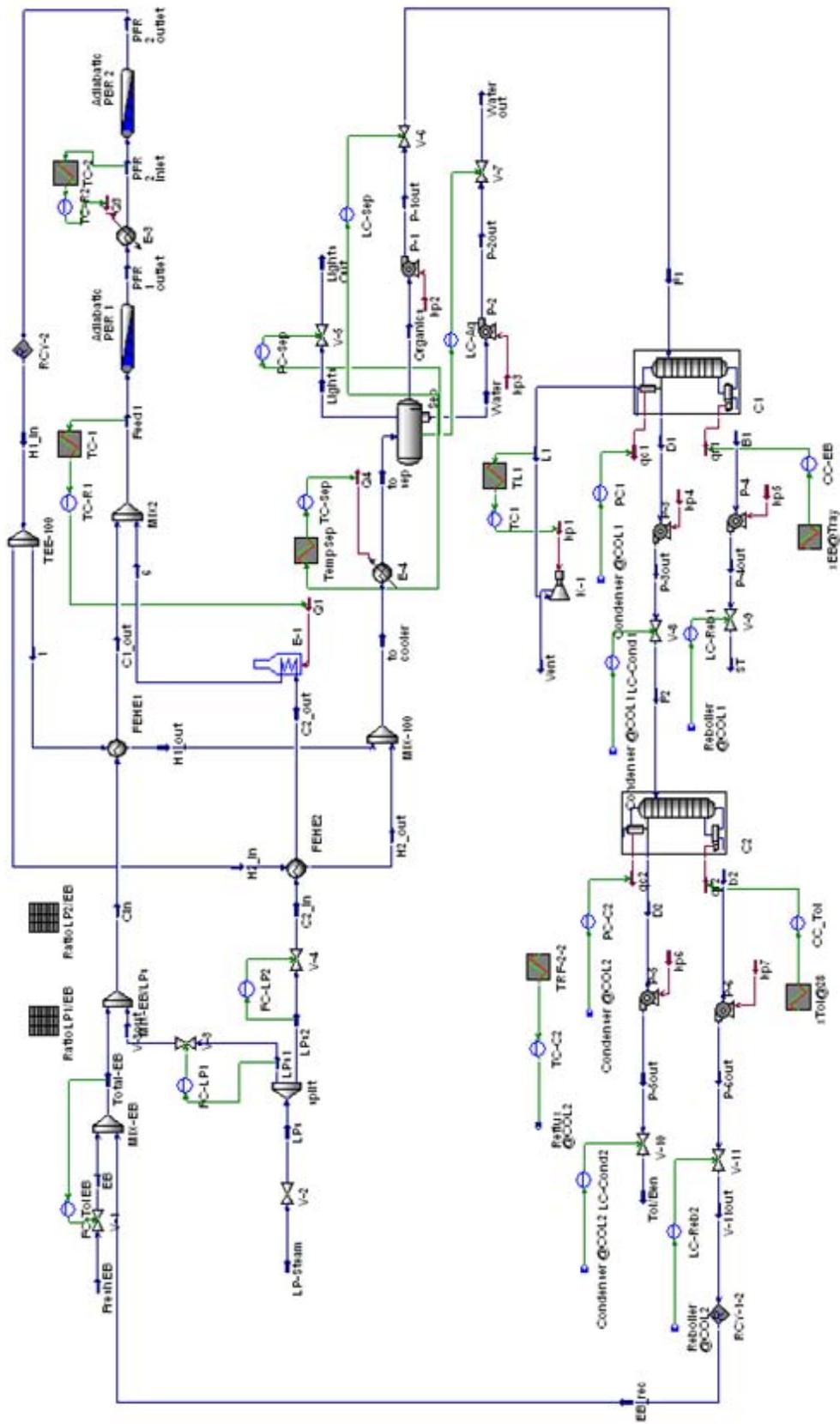


Figure B.4 Designed control structure III (CS3) for styrene process

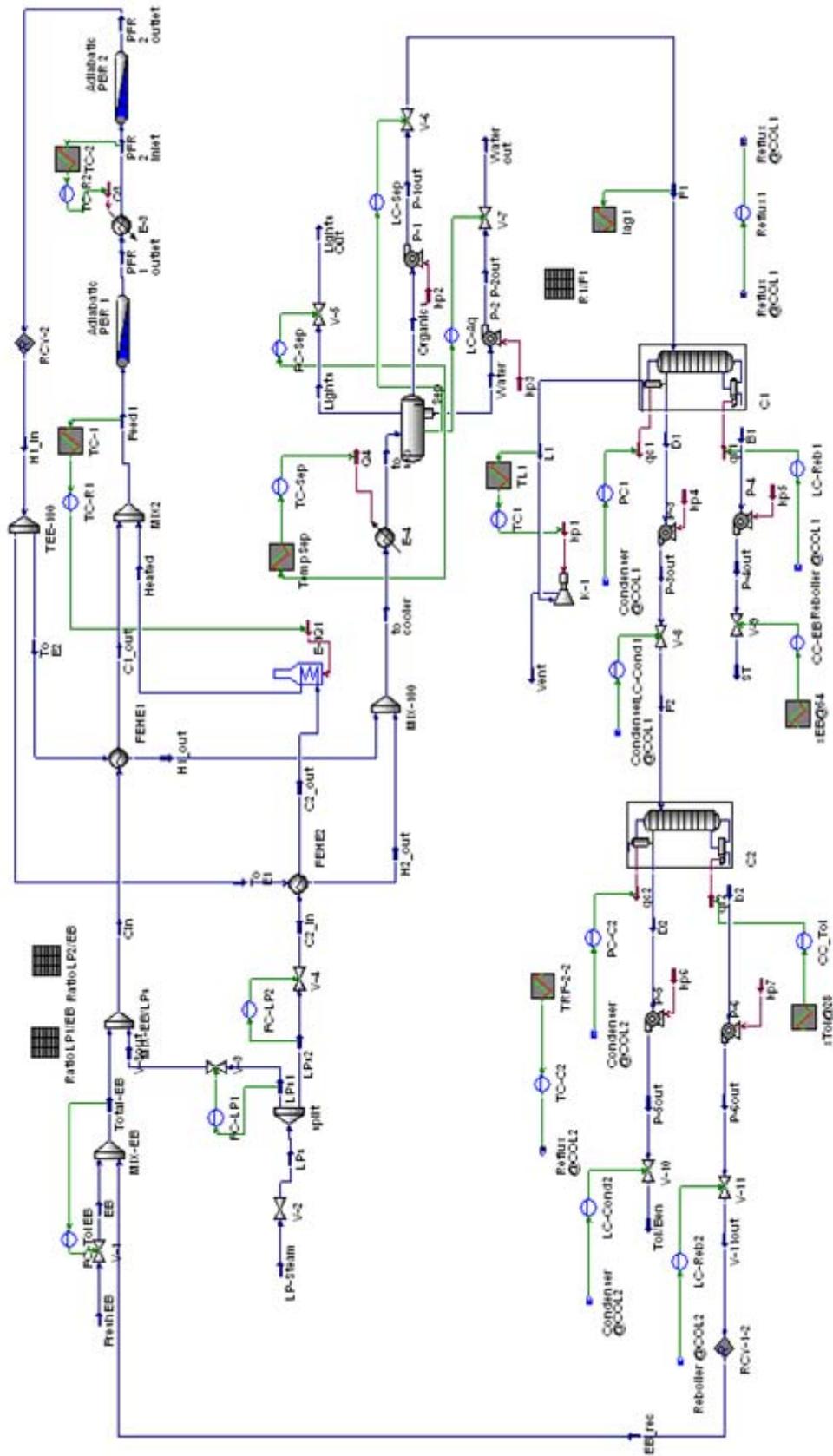


Figure B.5 Designed control structure IV (CS4) for styrene process

APPENDIX C

FIXTURE POINT THEOREM DATA

Table C.1 List of Manipulated Variables of the process

Manipulated Variables	Description
V-1	Ethylbenzene fresh feed valve
V-2	Low-pressure steam feed valve
V-3	Low-pressure steam split to mix valve
V-4	Low-pressure steam to furnace valve
V-5	3-phase separator vapor valve
V-6	3-phase separator organics valve
V-7	3-phase separator aqueous valve
V-8	Product column distillate valve
V-9	Product column bottom valve
V-10	Recycle column distillate valve
V-11	Recycle column bottom valve
Q1	Furnace heat duty
Q3	Intermediate heater duty
Q4	Cooler cooling water flow
Wk1	Product column overhead compressor duty
Qc1	Product column condenser duty
Qr1	Product column reboiler duty
Qc2	Recycle column condenser duty
Qr2	Recycle column reboiler duty
Reflux@C1	Product column reflux flow
Reflux@C2	Recycle column reflux flow

Table C.2 IAE Result of the process vessel

name	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
3-phase Separator	0.0821	2.7247	0.1647	1.3694	0.0079	0.0009	2.9022	1.1766	0.6215	1.1475	0.0020
	0.0023	0.0026	0.0022	0.0025	0.0006	0.0001	0.0213	0.0159	0.0007	0.0025	0.0003
	0.0288	0.5409	0.0446	0.3300	0.0005	0.0000	0.2054	0.0205	0.0014	0.0005	0.0005
	0.0184	0.2702	0.0172	0.1456	0.0009	0.0001	0.0438	0.0223	0.0016	0.0011	0.0006
Condenser @ C1	0.0005	0.0227	0.0005	0.0073	0.0000	0.0000	0.1781	0.3380	0.0036	0.0002	0.0020
	0.0003	0.0133	0.0007	0.0057	0.0000	0.0000	0.0074	0.0089	0.0059	0.0001	0.0003
Reboiler @ C1	0.0234	0.4530	0.0196	0.1400	0.0008	0.0003	2.4318	2.0569	0.4158	0.0018	0.0032
	0.0008	0.0208	0.0011	0.0095	0.0000	0.0000	0.5126	0.4833	0.0022	0.0000	0.0023
Condenser @ C2	0.0018	0.0594	0.0017	0.0199	0.0001	0.0000	0.0956	0.0482	0.0152	0.0002	0.0087
	0.0407	0.4198	0.0728	0.1643	0.0005	0.0000	2.4112	2.3098	0.0167	0.0007	0.0174

Table C.2 (Continued) IAE Result of the process vessel

name	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	Sum IAE
3-phase Separator											
	%L	0.0902	0.0486	0.0000	0.0657	0.0075	0.0656	0.0031	0.1123	0.1146	10.7134
	%Aq	0.0002	0.0015	0.0000	0.0009	0.0009	0.0019	0.0005	0.0023	0.0014	0.0831
	T	0.0003	0.0088	0.0000	0.0053	0.0022	0.0645	0.0010	0.0107	0.0211	7.9867
	P	0.0004	0.0063	0.0000	0.0031	0.0021	0.0173	0.0010	0.0177	0.0052	1.8737
Condenser @ C1											
	%L	0.1253	0.0146	0.0000	0.1183	0.0050	0.0211	0.0005	0.0005	0.0114	3.1570
	P	0.0021	0.0039	0.0000	0.0434	0.0008	0.0042	0.0001	0.0001	0.0021	0.7733
Reboiler @ C1											
	%L	0.0791	0.1159	0.1274	0.0760	0.0115	0.1371	0.0039	0.0250	0.0458	9.9573
Condenser @ C2											
	%L	0.0005	0.0048	0.0000	0.0007	0.1217	0.0247	0.0078	0.0006	0.1371	3.0037
	P	0.0062	0.0960	0.0000	0.0524	0.0258	0.5534	0.0142	0.0029	0.2449	9.2424
Reboiler @ C2											
	%L	0.0102	0.6566	0.0000	0.3120	0.0320	1.1204	0.0085	0.3718	0.5122	6.7595

Table C.3 IAE Result of Composition for the Process Stream

name	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
B1 (EB)	0.0117	0.0447	0.0035	0.0123	0.0004	0.0001	0.3201	0.0764	0.0101	0.0003	0.0049
D2 (EB)	0.0007	0.0198	0.0013	0.0089	0.0002	0.0002	0.5157	0.0029	0.0305	0.0003	0.0028
B2 (Toluene)	0.0008	0.0237	0.0014	0.0091	0.0004	0.0003	0.5208	0.0043	0.0359	0.0005	0.0071

Table C.3 (Continued) IAE Result of Composition for the Process Stream

name	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	Sum IAE
B1 (EB)	0.0035	0.0559	0.0684	0.0001	0.0325	0.053	0.0201	0.0082	0.1740	0.1429	0.0951
D2 (EB)	0.0064	0.0489	0.0240	0.0002	0.0506	0.0219	0.0157	0.0208	0.0028	0.0261	0.2452
B2 (Toluene)	0.0235	0.0267	0.0265	0.0003	0.0075	0.0258	0.1208	0.0253	0.0037	0.0181	0.3394

Table C.4 IAE Result of Tray EB Composition Deviation for Product column

Tray	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
1	0.0041	0.2240	0.0022	0.0154	0.0025	0.0002	0.1533	0.1418	0.1626	0.0015	0.0107
2	0.0018	0.1284	0.0014	0.0070	0.0012	0.0001	0.0465	0.4375	0.0689	0.0007	0.0037
3	0.0013	0.1147	0.0007	0.0042	0.0009	0.0001	0.1362	0.6704	0.0407	0.0005	0.0089
4	0.0014	0.1222	0.0005	0.0076	0.0008	0.0001	0.2018	0.8367	0.0310	0.0005	0.0119
5	0.0015	0.1337	0.0006	0.0123	0.0008	0.0001	0.2501	0.9766	0.0270	0.0005	0.0141
6	0.0017	0.1458	0.0008	0.0168	0.0008	0.0001	0.2906	1.1055	0.0251	0.0005	0.0161
7	0.0019	0.1577	0.0011	0.0213	0.0008	0.0002	0.3272	1.2284	0.0240	0.0005	0.0179
8	0.0023	0.1696	0.0015	0.0261	0.0009	0.0002	0.3618	1.3461	0.0235	0.0005	0.0197
9	0.0026	0.1813	0.0019	0.0311	0.0009	0.0002	0.3949	1.4577	0.0233	0.0005	0.0213
10	0.0031	0.1927	0.0023	0.0364	0.0009	0.0002	0.4262	1.5616	0.0235	0.0005	0.0229
11	0.0037	0.2039	0.0027	0.0418	0.0010	0.0003	0.4555	1.6555	0.0239	0.0005	0.0243
12	0.0046	0.2146	0.0031	0.0474	0.0011	0.0003	0.4822	1.7372	0.0246	0.0005	0.0256
13	0.0058	0.2248	0.0034	0.0532	0.0012	0.0005	0.5057	1.8052	0.0257	0.0005	0.0267
14	0.0074	0.2344	0.0036	0.0592	0.0014	0.0006	0.5252	1.8570	0.0273	0.0006	0.0275
15	0.0095	0.2436	0.0036	0.0652	0.0015	0.0007	0.5401	1.8909	0.0292	0.0007	0.0281
16	0.0124	0.2524	0.0035	0.0713	0.0017	0.0008	0.5499	1.9054	0.0314	0.0008	0.0284

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
17	0.0160	0.2609	0.0033	0.0778	0.0018	0.0008	0.5542	1.8998	0.0338	0.0008	0.0285
18	0.0207	0.2696	0.0030	0.0847	0.0020	0.0009	0.5529	1.8740	0.0363	0.0008	0.0283
19	0.0267	0.2784	0.0027	0.0923	0.0023	0.0009	0.5459	1.8286	0.0389	0.0010	0.0278
20	0.0342	0.2879	0.0025	0.1007	0.0026	0.0011	0.5334	1.7651	0.0413	0.0010	0.0270
21	0.0435	0.2983	0.0027	0.1101	0.0030	0.0011	0.5157	1.6851	0.0433	0.0011	0.0259
22	0.0549	0.3097	0.0038	0.1210	0.0035	0.0012	0.4936	1.5908	0.0449	0.0012	0.0247
23	0.0684	0.3224	0.0062	0.1335	0.0042	0.0013	0.4679	1.4849	0.0461	0.0013	0.0233
24	0.0846	0.3365	0.0094	0.1482	0.0052	0.0014	0.4393	1.3705	0.0466	0.0014	0.0218
25	0.1036	0.3521	0.0132	0.1653	0.0064	0.0016	0.4087	1.2505	0.0462	0.0015	0.0202
26	0.1257	0.3691	0.0177	0.1854	0.0079	0.0018	0.3764	1.1276	0.0452	0.0015	0.0187
27	0.1515	0.3875	0.0227	0.2092	0.0098	0.0019	0.3434	1.0035	0.0434	0.0019	0.0170
28	0.1810	0.4068	0.0286	0.2371	0.0120	0.0021	0.3101	0.8803	0.0407	0.0025	0.0152
29	0.2143	0.4263	0.0353	0.2696	0.0147	0.0023	0.2769	0.7592	0.0374	0.0032	0.0136
30	0.2518	0.4452	0.0431	0.3069	0.0180	0.0024	0.2434	0.6414	0.0338	0.0042	0.0121
31	0.2934	0.4626	0.0522	0.3495	0.0220	0.0027	0.2094	0.5270	0.0334	0.0054	0.0107
32	0.3392	0.4771	0.0628	0.3977	0.0268	0.0031	0.1740	0.4160	0.0361	0.0069	0.0095

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
33	0.3893	0.4876	0.0754	0.4517	0.0327	0.0037	0.1357	0.3082	0.0422	0.0085	0.0085
34	0.4436	0.4933	0.0900	0.5119	0.0397	0.0044	0.0928	0.2031	0.0527	0.0105	0.0079
35	0.4091	0.3676	0.0766	0.4497	0.0338	0.0039	0.0965	0.1987	0.0935	0.0056	0.0078
36	0.3768	0.3550	0.0638	0.4085	0.0288	0.0037	0.1064	0.2095	0.1926	0.0037	0.0080
37	0.3465	0.3974	0.0516	0.3807	0.0245	0.0034	0.1189	0.2290	0.2911	0.0038	0.0084
38	0.3182	0.4864	0.0402	0.3641	0.0209	0.0032	0.1331	0.2556	0.3925	0.0038	0.0093
39	0.2926	0.6105	0.0290	0.3571	0.0177	0.0031	0.1492	0.2891	0.5007	0.0038	0.0103
40	0.2696	0.7585	0.0181	0.3576	0.0146	0.0029	0.1674	0.3296	0.6186	0.0039	0.0111
41	0.2495	0.9267	0.0072	0.3651	0.0120	0.0028	0.1879	0.3776	0.7480	0.0043	0.0119
42	0.2325	1.1137	0.0050	0.3809	0.0097	0.0026	0.2109	0.4333	0.8907	0.0048	0.0128
43	0.2183	1.3179	0.0166	0.4076	0.0076	0.0024	0.2366	0.4966	1.0472	0.0058	0.0136
44	0.2070	1.5372	0.0292	0.4438	0.0055	0.0020	0.2649	0.5673	1.2180	0.0063	0.0143
45	0.1983	1.7695	0.0419	0.4875	0.0039	0.0017	0.2952	0.6448	1.4016	0.0071	0.0153
46	0.1923	2.0105	0.0548	0.5373	0.0026	0.0017	0.3271	0.7275	1.5962	0.0080	0.0164
47	0.1887	2.2543	0.0678	0.5916	0.0015	0.0015	0.3598	0.8137	1.7984	0.0088	0.0174
48	0.1867	2.4939	0.0805	0.6482	0.0014	0.0015	0.3919	0.9009	2.0031	0.0098	0.0185

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
49	0.1859	2.7205	0.0927	0.7050	0.0022	0.0017	0.4224	0.9860	2.2036	0.0107	0.0198
50	0.1861	2.9243	0.1040	0.7595	0.0033	0.0016	0.4499	1.0659	2.3925	0.0113	0.0209
51	0.1866	3.0954	0.1140	0.8085	0.0043	0.0014	0.4729	1.1362	2.5612	0.0119	0.0220
52	0.1867	3.2241	0.1222	0.8495	0.0050	0.0013	0.4901	1.1932	2.7007	0.0122	0.0228
53	0.1858	3.3029	0.1284	0.8798	0.0058	0.0014	0.5001	1.2335	2.8031	0.0124	0.0234
54	0.1833	3.3268	0.1319	0.8969	0.0064	0.0015	0.5022	1.2543	2.8619	0.0125	0.0233
55	0.1790	3.2943	0.1328	0.8996	0.0068	0.0017	0.4956	1.2543	2.8727	0.0123	0.0231
56	0.1725	3.2075	0.1313	0.8876	0.0071	0.0018	0.4812	1.2331	2.8351	0.0121	0.0224
57	0.1645	3.0715	0.1278	0.8616	0.0072	0.0018	0.4596	1.1923	2.7517	0.0116	0.0212
58	0.1550	2.8946	0.1225	0.8232	0.0071	0.0017	0.4319	1.1344	2.6277	0.0108	0.0195
59	0.1442	2.6869	0.1156	0.7745	0.0069	0.0015	0.3996	1.0629	2.4706	0.0098	0.0175
60	0.1327	2.4590	0.1076	0.7186	0.0065	0.0012	0.3642	0.9818	2.2892	0.0084	0.0155
61	0.1208	2.2213	0.0990	0.6588	0.0063	0.0014	0.3275	0.8949	2.0932	0.0069	0.0134
62	0.1088	1.9835	0.0897	0.5968	0.0058	0.0017	0.2907	0.8062	1.8905	0.0055	0.0117
63	0.0971	1.7524	0.0801	0.5346	0.0052	0.0019	0.2551	0.7188	1.6883	0.0044	0.0102
64	0.0859	1.5339	0.0709	0.4745	0.0047	0.0020	0.2216	0.6348	1.4926	0.0034	0.0086

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
65	0.0755	1.3317	0.0621	0.4175	0.0043	0.0021	0.1906	0.5560	1.3076	0.0024	0.0071
66	0.0660	1.1485	0.0541	0.3648	0.0041	0.0023	0.1625	0.4834	1.1366	0.0015	0.0056
67	0.0575	0.9843	0.0468	0.3166	0.0039	0.0025	0.1376	0.4175	0.9814	0.0009	0.0044
68	0.0498	0.8385	0.0403	0.2733	0.0036	0.0027	0.1161	0.3586	0.8422	0.0011	0.0036
69	0.0428	0.7110	0.0344	0.2344	0.0031	0.0024	0.0972	0.3066	0.7184	0.0013	0.0032
70	0.0363	0.6009	0.0292	0.2005	0.0026	0.0022	0.0812	0.2615	0.6105	0.0013	0.0028
71	0.0306	0.5054	0.0246	0.1699	0.0023	0.0020	0.0674	0.2218	0.5160	0.0013	0.0026
72	0.0258	0.4235	0.0207	0.1431	0.0019	0.0018	0.0552	0.1868	0.4337	0.0012	0.0026
73	0.0215	0.3537	0.0171	0.1197	0.0017	0.0018	0.0450	0.1560	0.3625	0.0011	0.0025
74	0.0181	0.2950	0.0143	0.1002	0.0016	0.0018	0.0368	0.1296	0.3024	0.0011	0.0024
75	0.0151	0.2450	0.0119	0.0833	0.0015	0.0019	0.0303	0.1071	0.2515	0.0010	0.0022
76	0.0126	0.2025	0.0098	0.0693	0.0014	0.0020	0.0247	0.0878	0.2083	0.0011	0.0020
77	0.0105	0.1664	0.0081	0.0574	0.0014	0.0021	0.0199	0.0710	0.1715	0.0011	0.0017
78	0.0089	0.1360	0.0066	0.0471	0.0015	0.0024	0.0159	0.0564	0.1400	0.0012	0.0017
79	0.0074	0.1103	0.0051	0.0380	0.0013	0.0024	0.0123	0.0438	0.1128	0.0011	0.0016
80	0.0062	0.0884	0.0039	0.0303	0.0013	0.0022	0.0091	0.0332	0.0894	0.0011	0.0015

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	Sum IAE
1	0.0242	0.0476	0.0744	0.0002	0.2139	0.0117	0.0621	0.0050	0.0337	0.0272	1.2181
2	0.0381	0.0822	0.1003	0.0001	0.3683	0.0045	0.0673	0.0005	0.0184	0.0465	1.4235
3	0.0651	0.1412	0.0786	0.0000	0.4857	0.0057	0.1426	0.0021	0.0133	0.0719	1.9849
4	0.0875	0.1856	0.0524	0.0001	0.5656	0.0072	0.1950	0.0036	0.0116	0.0921	2.4150
5	0.1078	0.2240	0.0273	0.0001	0.6303	0.0081	0.2380	0.0048	0.0109	0.1099	2.7786
6	0.1276	0.2604	0.0141	0.0002	0.6873	0.0088	0.2773	0.0058	0.0106	0.1268	3.1226
7	0.1474	0.2962	0.0303	0.0002	0.7385	0.0094	0.3155	0.0068	0.0105	0.1435	3.4796
8	0.1675	0.3319	0.0620	0.0003	0.7839	0.0098	0.3532	0.0079	0.0104	0.1600	3.8389
9	0.1875	0.3672	0.0973	0.0003	0.8230	0.0109	0.3905	0.0089	0.0103	0.1764	4.1879
10	0.2072	0.4016	0.1365	0.0003	0.8548	0.0124	0.4269	0.0099	0.0102	0.1924	4.5228
11	0.2264	0.4348	0.1790	0.0003	0.8786	0.0142	0.4620	0.0108	0.0101	0.2080	4.8375
12	0.2444	0.4659	0.2236	0.0004	0.8934	0.0159	0.4949	0.0117	0.0102	0.2229	5.1246
13	0.2609	0.4943	0.2692	0.0004	0.8986	0.0176	0.5250	0.0124	0.0102	0.2368	5.3782
14	0.2754	0.5192	0.3146	0.0006	0.8941	0.0192	0.5516	0.0130	0.0103	0.2496	5.5919
15	0.2876	0.5401	0.3586	0.0007	0.8802	0.0207	0.5737	0.0136	0.0106	0.2613	5.7604
16	0.2971	0.5566	0.4002	0.0009	0.8570	0.0222	0.5910	0.0141	0.0109	0.2717	5.8795

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	Sum IAE
17	0.0957	0.6962	0.2373	0.0004	0.1142	0.0307	0.4855	0.0056	0.0982	0.8354	4.5424
18	0.0790	0.7669	0.1785	0.0004	0.1770	0.0433	0.4911	0.0086	0.1116	0.9430	4.7491
19	0.0876	0.6823	0.1237	0.0005	0.1710	0.0318	0.4679	0.0056	0.0899	0.8457	4.2489
20	0.0972	0.6167	0.0864	0.0006	0.1670	0.0245	0.4549	0.0048	0.0722	0.7606	4.0418
21	0.1082	0.5648	0.0648	0.0006	0.1667	0.0201	0.4461	0.0045	0.0567	0.6849	3.9726
22	0.1210	0.5245	0.0794	0.0008	0.1702	0.0186	0.4413	0.0047	0.0422	0.6174	4.0474
23	0.1357	0.4952	0.1511	0.0011	0.1776	0.0189	0.4405	0.0055	0.0286	0.5580	4.2754
24	0.1529	0.4764	0.2380	0.0013	0.1887	0.0204	0.4457	0.0066	0.0163	0.5066	4.6047
25	0.1728	0.4681	0.3365	0.0012	0.2036	0.0231	0.4582	0.0079	0.0081	0.4631	5.0354
26	0.1957	0.4697	0.4474	0.0012	0.2220	0.0263	0.4791	0.0092	0.0157	0.4273	5.5905
27	0.2219	0.4810	0.5716	0.0012	0.2443	0.0304	0.5086	0.0110	0.0307	0.3992	6.2700
28	0.2507	0.5010	0.7085	0.0012	0.2697	0.0346	0.5472	0.0126	0.0462	0.3782	7.0454
29	0.2824	0.5294	0.8581	0.0013	0.2985	0.0397	0.5935	0.0145	0.0619	0.3639	7.9099
30	0.3162	0.5646	1.0191	0.0010	0.3301	0.0453	0.6467	0.0167	0.0778	0.3556	8.8473
31	0.3517	0.6049	1.1884	0.0010	0.3637	0.0510	0.7053	0.0187	0.0936	0.3523	9.8341
32	0.3877	0.6486	1.3614	0.0009	0.3987	0.0571	0.7668	0.0210	0.1091	0.3531	10.8409

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	Sum IAE
33	0.0957	0.6962	0.2373	0.0004	0.1142	0.0307	0.4855	0.0056	0.0982	0.8354	4.5424
34	0.0790	0.7669	0.1785	0.0004	0.1770	0.0433	0.4911	0.0086	0.1116	0.9430	4.7491
35	0.0876	0.6823	0.1237	0.0005	0.1710	0.0318	0.4679	0.0056	0.0899	0.8457	4.2489
36	0.0972	0.6167	0.0864	0.0006	0.1670	0.0245	0.4549	0.0048	0.0722	0.7606	4.0418
37	0.1082	0.5648	0.0648	0.0006	0.1667	0.0201	0.4461	0.0045	0.0567	0.6849	3.9726
38	0.1210	0.5245	0.0794	0.0008	0.1702	0.0186	0.4413	0.0047	0.0422	0.6174	4.0474
39	0.1357	0.4952	0.1511	0.0011	0.1776	0.0189	0.4405	0.0055	0.0286	0.5580	4.2754
40	0.1529	0.4764	0.2380	0.0013	0.1887	0.0204	0.4457	0.0066	0.0163	0.5066	4.6047
41	0.1728	0.4681	0.3365	0.0012	0.2036	0.0231	0.4582	0.0079	0.0081	0.4631	5.0354
42	0.1957	0.4697	0.4474	0.0012	0.2220	0.0263	0.4791	0.0092	0.0157	0.4273	5.5905
43	0.2219	0.4810	0.5716	0.0012	0.2443	0.0304	0.5086	0.0110	0.0307	0.3992	6.2700
44	0.2507	0.5010	0.7085	0.0012	0.2697	0.0346	0.5472	0.0126	0.0462	0.3782	7.0454
45	0.2824	0.5294	0.8581	0.0013	0.2985	0.0397	0.5935	0.0145	0.0619	0.3639	7.9099
46	0.3162	0.5646	1.0191	0.0010	0.3301	0.0453	0.6467	0.0167	0.0778	0.3556	8.8473
47	0.3517	0.6049	1.1884	0.0010	0.3637	0.0510	0.7053	0.0187	0.0936	0.3523	9.8341
48	0.3877	0.6486	1.3614	0.0009	0.3987	0.0571	0.7668	0.0210	0.1091	0.3531	10.8409

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	Sum IAE
49	0.4230	0.6931	1.5326	0.0010	0.4339	0.0630	0.8283	0.0232	0.1238	0.3566	11.8291
50	0.4557	0.7356	1.6962	0.0009	0.4671	0.0684	0.8873	0.0252	0.1376	0.3615	12.7552
51	0.4844	0.7733	1.8459	0.0010	0.4974	0.0731	0.9402	0.0270	0.1495	0.3663	13.5724
52	0.5075	0.8031	1.9749	0.0012	0.5231	0.0768	0.9832	0.0283	0.1594	0.3695	14.2350
53	0.5237	0.8231	2.0762	0.0014	0.5425	0.0794	1.0136	0.0292	0.1669	0.3700	14.7025
54	0.5319	0.8315	2.1442	0.0015	0.5543	0.0809	1.0289	0.0296	0.1713	0.3671	14.9424
55	0.5314	0.8271	2.1750	0.0016	0.5578	0.0812	1.0277	0.0295	0.1724	0.3603	14.9363
56	0.5223	0.8098	2.1675	0.0014	0.5526	0.0800	1.0098	0.0291	0.1702	0.3491	14.6836
57	0.5052	0.7801	2.1230	0.0013	0.5389	0.0775	0.9764	0.0281	0.1653	0.3335	14.2000
58	0.4807	0.7398	2.0451	0.0014	0.5174	0.0737	0.9292	0.0264	0.1579	0.3141	13.5144
59	0.4505	0.6915	1.9398	0.0017	0.4896	0.0692	0.8704	0.0242	0.1484	0.2921	12.6674
60	0.4164	0.6373	1.8133	0.0020	0.4562	0.0638	0.8041	0.0221	0.1375	0.2679	11.7050
61	0.3794	0.5793	1.6718	0.0021	0.4186	0.0578	0.7334	0.0195	0.1259	0.2423	10.6738
62	0.3421	0.5205	1.5236	0.0021	0.3795	0.0516	0.6609	0.0171	0.1137	0.2169	9.6188
63	0.3051	0.4628	1.3754	0.0020	0.3399	0.0457	0.5889	0.0150	0.1015	0.1921	8.5762
64	0.2695	0.4074	1.2314	0.0018	0.3009	0.0400	0.5196	0.0130	0.0898	0.1684	7.5744

Table C.4 (Continued) IAE Result of Tray EB Composition Deviation for Product column

Tray	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	SumIAE
65	0.2360	0.3554	1.0933	0.0018	0.2636	0.0346	0.4546	0.0110	0.0789	0.1464	6.6327
66	0.2051	0.3077	0.9617	0.0018	0.2287	0.0296	0.3948	0.0093	0.0689	0.1261	5.7628
67	0.1770	0.2643	0.8422	0.0019	0.1965	0.0250	0.3407	0.0077	0.0599	0.1076	4.9762
68	0.1517	0.2257	0.7346	0.0021	0.1674	0.0210	0.2923	0.0064	0.0518	0.0910	4.2736
69	0.1297	0.1920	0.6383	0.0018	0.1420	0.0176	0.2494	0.0056	0.0444	0.0768	3.6523
70	0.1102	0.1628	0.5521	0.0014	0.1197	0.0147	0.2121	0.0050	0.0377	0.0647	3.1096
71	0.0933	0.1379	0.4753	0.0012	0.1008	0.0124	0.1787	0.0043	0.0317	0.0547	2.6340
72	0.0788	0.1168	0.4075	0.0009	0.0850	0.0104	0.1495	0.0038	0.0264	0.0463	2.2216
73	0.0663	0.0984	0.3470	0.0008	0.0719	0.0087	0.1243	0.0034	0.0219	0.0390	1.8643
74	0.0552	0.0822	0.2952	0.0009	0.0607	0.0071	0.1035	0.0028	0.0183	0.0324	1.5616
75	0.0457	0.0683	0.2491	0.0008	0.0511	0.0058	0.0858	0.0024	0.0154	0.0269	1.3022
76	0.0377	0.0562	0.2075	0.0009	0.0430	0.0047	0.0709	0.0021	0.0128	0.0221	1.0794
77	0.0310	0.0458	0.1694	0.0010	0.0358	0.0037	0.0584	0.0019	0.0109	0.0180	0.8872
78	0.0255	0.0371	0.1376	0.0011	0.0299	0.0029	0.0477	0.0016	0.0091	0.0145	0.7247
79	0.0210	0.0302	0.1105	0.0012	0.0253	0.0023	0.0383	0.0014	0.0075	0.0119	0.5858
80	0.0170	0.0243	0.0868	0.0014	0.0210	0.0020	0.0302	0.0012	0.0059	0.0097	0.4661

Table C.5 IAE Result of Tray Temperature Deviation for Recycle column

Tray	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
1	0.0228	0.5745	0.0078	0.0922	0.0115	0.0007	0.5277	0.7628	0.4678	0.0034	0.0981
2	0.0205	0.5764	0.0107	0.1467	0.0058	0.0005	0.2471	0.8915	0.1139	0.0007	0.0232
3	0.0240	0.7622	0.0265	0.2634	0.0015	0.0005	0.9542	1.2170	0.2228	0.0019	0.1066
4	0.0301	1.0136	0.0452	0.4059	0.0031	0.0005	1.8072	1.6274	0.5523	0.0047	0.2147
5	0.0360	1.2648	0.0644	0.5530	0.0073	0.0006	2.6698	2.0127	0.9449	0.0078	0.3291
6	0.0380	1.3926	0.0778	0.6498	0.0112	0.0007	3.2459	2.1629	1.2965	0.0105	0.4171
7	0.0336	1.2958	0.0782	0.6421	0.0132	0.0008	3.2470	1.9200	1.4726	0.0118	0.4395
8	0.0244	1.0087	0.0659	0.5340	0.0128	0.0007	2.6981	1.4668	1.4199	0.0112	0.3911
9	0.0168	0.7231	0.0476	0.3843	0.0108	0.0007	1.9135	1.1285	1.2172	0.0095	0.3058
10	0.0127	0.5328	0.0307	0.2491	0.0085	0.0006	1.1938	0.9542	0.9835	0.0077	0.2217
11	0.0111	0.4347	0.0180	0.1510	0.0066	0.0005	0.6642	0.8979	0.7914	0.0061	0.1574
12	0.0108	0.3984	0.0099	0.0882	0.0054	0.0005	0.3194	0.9020	0.6596	0.0051	0.1146
13	0.0110	0.3969	0.0072	0.0511	0.0046	0.0005	0.1832	0.9331	0.5798	0.0044	0.0886
14	0.0114	0.4145	0.0070	0.0364	0.0043	0.0005	0.1684	0.9776	0.5373	0.0042	0.0736
15	0.0119	0.4431	0.0074	0.0357	0.0042	0.0005	0.1757	1.0308	0.5192	0.0043	0.0654
16	0.0109	0.3881	0.0066	0.0331	0.0040	0.0005	0.1887	0.9085	0.4880	0.0039	0.0648
17	0.0098	0.3339	0.0058	0.0299	0.0037	0.0005	0.4078	0.7856	0.4507	0.0036	0.0629
18	0.0087	0.2835	0.0051	0.0269	0.0033	0.0005	0.6123	0.6696	0.4113	0.0034	0.0605

Table C.5 (Continued) IAE Result of Tray Temperature Deviation for Recycle column

Tray	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
1	0.0522	1.8188	0.0616	0.6683	0.0074	0.0007	1.8188	3.3300	0.7538	0.0071	0.3467
2	0.0535	1.8248	0.0587	0.6541	0.0061	0.0007	1.8248	3.2530	0.6847	0.0064	0.3277
3	0.0556	1.8684	0.0597	0.6666	0.0058	0.0010	1.8684	3.2499	0.6756	0.0064	0.3281
4	0.0575	1.9284	0.0624	0.6926	0.0058	0.0007	1.9284	3.2955	0.7161	0.0066	0.3430
5	0.0596	1.9983	0.0668	0.7278	0.0067	0.0009	1.9983	3.3639	0.7947	0.0072	0.3673
6	0.0610	2.0551	0.0710	0.7602	0.0077	0.0007	2.0551	3.4186	0.8881	0.0078	0.3929
7	0.0609	2.0673	0.0729	0.7735	0.0083	0.0006	2.0673	3.4234	0.9592	0.0083	0.4080
8	0.0594	2.0229	0.0713	0.7595	0.0085	0.0007	2.0229	3.3699	0.9778	0.0084	0.4050
9	0.0570	1.9424	0.0672	0.7263	0.0083	0.0007	1.9424	3.2858	0.9461	0.0082	0.3877
10	0.0547	1.8581	0.0628	0.6892	0.0078	0.0007	1.8581	3.2063	0.8913	0.0078	0.3658
11	0.0528	1.7886	0.0587	0.6583	0.0072	0.0007	1.7886	3.1480	0.8385	0.0075	0.3468
12	0.0513	1.7366	0.0558	0.6360	0.0067	0.0010	1.7366	3.1101	0.7989	0.0076	0.3332
13	0.0505	1.6982	0.0541	0.6211	0.0064	0.0006	1.6982	3.0865	0.7740	0.0072	0.3241
14	0.0501	1.6678	0.0531	0.6111	0.0066	0.0008	1.6678	3.0707	0.7606	0.0071	0.3183
15	0.0495	1.6412	0.0522	0.6037	0.0066	0.0007	1.6412	3.0582	0.7545	0.0070	0.3149
16	0.0515	1.7298	0.0518	0.6145	0.0048	0.0006	1.7298	3.0535	0.6946	0.0068	0.3048
17	0.0528	1.7808	0.0519	0.6205	0.0039	0.0007	1.7808	3.0527	0.6561	0.0062	0.2987
18	0.0536	1.8092	0.0519	0.6226	0.0035	0.0008	1.8092	3.0558	0.6292	0.0063	0.2946

Table C.5 (Continued) IAE Result of Tray Temperature Deviation for Recycle column

Tray	V-1	V-2	V-3	V-4	Q1	Q3	Q4	V-5	V-6	V-7	Q5
19	0.0539	1.8241	0.0518	0.6225	0.0030	0.0007	1.8241	3.0619	0.6080	0.0059	0.2916
20	0.0541	1.8320	0.0518	0.6215	0.0028	0.0006	1.8320	3.0702	0.5902	0.0057	0.2890
21	0.0542	1.8356	0.0516	0.6200	0.0027	0.0007	1.8356	3.0797	0.5743	0.0058	0.2867
22	0.0542	1.8365	0.0515	0.6181	0.0025	0.0008	1.8365	3.0893	0.5598	0.0055	0.2845
23	0.0543	1.8356	0.0514	0.6161	0.0025	0.0007	1.8356	3.0986	0.5462	0.0055	0.2825
24	0.0543	1.8334	0.0512	0.6140	0.0023	0.0007	1.8334	3.1065	0.5335	0.0053	0.2804
25	0.0542	1.8300	0.0511	0.6118	0.0022	0.0007	1.8300	3.1126	0.5216	0.0052	0.2784
26	0.0542	1.8257	0.0509	0.6095	0.0020	0.0007	1.8257	3.1160	0.5105	0.0053	0.2766
27	0.0541	1.8206	0.0507	0.6072	0.0020	0.0007	1.8206	3.1162	0.5003	0.0051	0.2745
28	0.0541	1.8148	0.0505	0.6048	0.0020	0.0007	1.8148	3.1123	0.4909	0.0049	0.2725
29	0.0540	1.8082	0.0504	0.6022	0.0021	0.0007	1.8082	3.1037	0.4824	0.0049	0.2704
30	0.0538	1.8008	0.0501	0.5993	0.0020	0.0009	1.8008	3.0899	0.4746	0.0049	0.2683
31	0.0537	1.7928	0.0498	0.5965	0.0019	0.0006	1.7928	3.0707	0.4678	0.0051	0.2663
32	0.0535	1.7842	0.0495	0.5933	0.0018	0.0006	1.7842	3.0457	0.4619	0.0047	0.2642
33	0.0533	1.7748	0.0492	0.5901	0.0018	0.0006	1.7748	3.0146	0.4566	0.0047	0.2620
34	0.0531	1.7647	0.0489	0.5867	0.0018	0.0007	1.7647	2.9780	0.4521	0.0048	0.2598
35	0.0528	1.7543	0.0486	0.5830	0.0018	0.0005	1.7543	2.9358	0.4483	0.0045	0.2575

Table C.5 (Continued) IAE Result of Tray Temperature Deviation for Recycle column

Tray	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	Sum IAE
1	0.1365	3.3300	4.0329	0.0010	1.9057	0.5807	16.9180	0.1304	0.1118	8.8848	44.8972
2	0.1467	3.2530	3.9715	0.0012	1.8784	0.5882	16.7508	0.1759	0.1075	8.7230	44.2909
3	0.1446	3.2499	4.0039	0.0011	1.9280	0.5334	16.8725	0.1395	0.1093	8.8496	44.6173
4	0.1277	3.2955	4.1116	0.0011	2.0543	0.4040	17.1817	0.0186	0.1156	9.2343	45.5813
5	0.1011	3.3639	4.2675	0.0013	2.2487	0.3098	17.5583	0.1645	0.1265	9.8027	47.3357
6	0.0779	3.4186	4.4222	0.0012	2.4699	0.4232	17.8320	0.3494	0.1389	10.3719	49.2236
7	0.0713	3.4234	4.5094	0.0013	2.6350	0.4868	17.8442	0.4449	0.1484	10.6626	50.0769
8	0.0772	3.3699	4.4940	0.0013	2.6713	0.4582	17.5821	0.3956	0.1513	10.4937	49.4006
9	0.0907	3.2858	4.4035	0.0012	2.5815	0.3600	17.1910	0.2338	0.1479	9.9604	47.6279
10	0.1103	3.2063	4.2934	0.0012	2.4323	0.3547	16.8285	0.0396	0.1417	9.3250	45.7355
11	0.1311	3.1480	4.2008	0.0011	2.2869	0.5150	16.5639	0.1275	0.1356	8.7841	44.5895
12	0.1480	3.1101	4.1365	0.0015	2.1735	0.6358	16.3924	0.2477	0.1309	8.3967	43.8468
13	0.1610	3.0865	4.0959	0.0013	2.0960	0.7147	16.2835	0.3262	0.1284	8.1423	43.3568
14	0.1702	3.0707	4.0719	0.0009	2.0466	0.7629	16.2099	0.3745	0.1272	7.9789	43.0275
15	0.1764	3.0582	4.0585	0.0008	2.0161	0.7907	16.1540	0.4027	0.1268	7.8709	42.7850
16	0.1679	3.0535	3.9815	0.0010	1.9561	0.7817	16.1439	0.4023	0.1218	7.8534	42.7057
17	0.1620	3.0527	3.9275	0.0009	1.9149	0.7765	16.1421	0.4024	0.1180	7.8333	42.6355
18	0.1577	3.0558	3.8852	0.0009	1.8815	0.7720	16.1460	0.4020	0.1150	7.8089	42.5617

Table C.5 (Continued) IAE Result of Tray Temperature Deviation for Recycle column

Tray	qc1	V-8	qr1	V-9	R1	qc2	qr2	V-10	V-11	R2	Sum IAE
19	0.1542	3.0619	3.8490	0.0010	1.8507	0.7675	16.1541	0.4011	0.1122	7.7809	42.4800
20	0.1514	3.0702	3.8163	0.0010	1.8213	0.7632	16.1637	0.3999	0.1099	7.7517	42.3983
21	0.1493	3.0797	3.7856	0.0008	1.7927	0.7590	16.1730	0.3987	0.1077	7.7219	42.3151
22	0.1477	3.0893	3.7566	0.0010	1.7647	0.7550	16.1804	0.3973	0.1056	7.6922	42.2290
23	0.1469	3.0986	3.7290	0.0008	1.7379	0.7514	16.1842	0.3960	0.1037	7.6631	42.1404
24	0.1465	3.1065	3.7026	0.0008	1.7118	0.7481	16.1829	0.3949	0.1019	7.6347	42.0459
25	0.1469	3.1126	3.6775	0.0009	1.6873	0.7453	16.1749	0.3939	0.1002	7.6070	41.9443
26	0.1477	3.1160	3.6536	0.0009	1.6644	0.7429	16.1587	0.3932	0.0985	7.5800	41.8328
27	0.1491	3.1162	3.6308	0.0009	1.6431	0.7410	16.1329	0.3929	0.0970	7.5537	41.7094
28	0.1509	3.1123	3.6088	0.0008	1.6237	0.7397	16.0962	0.3928	0.0955	7.5277	41.5707
29	0.1529	3.1037	3.5880	0.0009	1.6063	0.7389	16.0479	0.3931	0.0942	7.5021	41.4150
30	0.1551	3.0899	3.5680	0.0010	1.5908	0.7383	15.9874	0.3936	0.0928	7.4762	41.2384
31	0.1576	3.0707	3.5487	0.0008	1.5775	0.7383	15.9141	0.3947	0.0915	7.4502	41.0420
32	0.1602	3.0457	3.5299	0.0007	1.5663	0.7387	15.8280	0.3961	0.0903	7.4240	40.8235
33	0.1629	3.0146	3.5117	0.0007	1.5572	0.7393	15.7296	0.3977	0.0891	7.3969	40.5822
34	0.1656	2.9780	3.4939	0.0009	1.5501	0.7402	15.6197	0.3997	0.0879	7.3690	40.3203
35	0.1683	2.9358	3.4761	0.0006	1.5451	0.7414	15.4985	0.4021	0.0869	7.3403	40.0366

VITA

Miss Keadsuda Machuay was born in Suphanburi, Thailand on March 13, 1986. She received the Bachelor Degree in Chemical Engineering from Rajamangala University of Technology Thanyaburi in 2005. After that she entered the Graduate School of Chulalongkorn University to pursue the Master of Engineering in Chemical Engineering and completed in 2009.