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PLANTWIDE CONTROL STRUCTURE DESIGN OF STYRENE MONOMER PLANT

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บุญธรรม ศรีคำวงษ์ : การออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์สำหรับกระบวนการผลิตสไตรีน. (PLANTWIDE CONTROL STRUCTURE DESIGN OF STYRENE MONOMER PLANT) อ.ที่ปรึกษาวิทยานิพนธ์หลัก: ผศ. ดร.มนตรี วงศ์ศรี, 115 หน้า.

การออกแบบโครงสร้างการควบคุมแบบทั้งโรงงานคือการออกแบบระบบควบคุมของทั้งโรงงาน การออกแบบนี้มีความสำคัญมากเมื่อโรงงานที่ออกแบบมีหน่วยการผลิตเชื่อมโยงกันมาก โดยมีการนำสารตั้งต้นและพลังงานกลับมาใช้ใหม่ วิธีการออกแบบได้มีการพัฒนามามี 4 แบบคือ ใช้คณิตศาสตร์ ใช้ออพติไมเซชั่น ใช้ฮิวริสติก และวิธีผสม วิธีการออกแบบของวงศ์ศรีซึ่งเป็นวิธีฮิวริสติก ประกอบด้วย 5 ช่วงและ 8 ขั้นตอน โดยให้ความสำคัญกับการควบคุมเพื่อยึดให้กระบวนการผลิตอยู่ในสภาวะคงตัวที่ต้องการ โดยออกแบบลูปควบคุมจากตัวระบุปริมาณสารแต่ละชนิดในกระบวนการโดยใช้ตัวจัดการเพื่อควบคุมการไหลขององค์ประกอบ ลูปการควบคุมในระดับโรงงานอื่นๆ ถูกออกแบบโดยใช้การจัดการตัวรบกวนทางความร้อนและสสารเพื่อให้ได้ผลิตภัณฑ์ที่มีคุณภาพตามที่กำหนด ลักษณะสำคัญอีกอย่างหนึ่งคือการใช้มาตรสมรรถนะเชิงพลวัตในการเลือกโครงสร้างการควบคุมของหอกลั่น วิธีการออกแบบนี้ถูกประยุกต์ใช้กับโรงงานผลิตสไตรีนโมโนเมอร์ที่ออปติมัมของ Luyben (William L. Luyben, Design and Control of the Styrene Process, Ind. Eng. Chem., Vol. 50, No. 3, 2011) สมรรถนะของโครงสร้างการควบคุมใหม่ถูกเปรียบเทียบกับโครงสร้างกรณีอ้างอิง (S. Vasudevan and G. P. Rangaiah, Integrated Framework Incorporating Optimization for Plant-Wide Control of Industrial Processes, I&EC Research, Vol. 50, 2011 and S. Vasudevan, et. al., Application and evaluation of three methodologies for plantwide control of the styrene monomer plant, Ind. Eng. Chem. Vol. 48, No. 24, 2009) สมรรถนะของโครงสร้างการควบคุมใหม่เทียบเท่าหรือดีกว่าโครงสร้างกรณีศึกษา วิธีการออกแบบของวงศ์ศรีมีรายละเอียด เป็นขั้นตอนเข้าใจได้ง่าย สามารถนำไปใช้งานได้แม้สำหรับนักออกแบบหน้าใหม่

ภาควิชา วิศวกรรมเคมี ลายมือชื่อนิสิต .....

สาขาวิชา วิศวกรรมเคมี ลายมือชื่อ อ.ที่ปรึกษาวิทยานิพนธ์หลัก .....

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BOONTUM SIKUMWONG: PLANTWIDE CONTROL STRUCTURE DESIGN OF STYRENE MONOMER PLANT. ADVISOR: ASST. PROF. DR.MONTREE WONGSRI, D.Sc., 115 pp.

Plantwide control structure design is to design a control system for an entire plant. It becomes crucial when plant in considered consisting of many interconnected unit operations with material and energy recycles. The methodologies for plantwide control design have been developed in four styles. They are mathematical, optimization, heuristic and mixed approaches. Wongsri's Design Procedure, heuristic in nature, is illustrated in this paper. It comprises of five stages with eight steps, emphasizing the establishment of a fixture plant by designing control loops, using the material quantifiers and their handlers, to regulate material component flows. The other plant level loops are designed using material and heat disturbance management for quality control. Another important feature is the use of dynamic performance measures to select an appropriate structure for a control distillation column. This method is applied to the styrene monomer plant based on the optimum design process from Luyben (William L. Luyben, Design and Control of the Styrene Process, Ind. Eng. Chem., Vol. 50, No. 3, 2011). The performance of new design structure is compared with two base case control structures selected (S. Vasudevan and G. P. Rangaiah, Integrated Framework Incorporating Optimization for Plant-Wide Control of Industrial Processes, I&EC Research, Vol. 50, 2011 and S. Vasudevan, et. al., Application and evaluation of three methodologies for plantwide control of the styrene monomer plant, Ind. Eng. Chem. Vol. 48, No. 24, 2009). The new control structure performance is comparable with the base cases. The Wongsri design procedure is detailed, instructive, clearcut, and readily to be applied even for a novice designer.

Department: Chemical Engineering Student's Signature .....

Field of Study: Chemical Engineering Advisor's Signature .....

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CONTENTS

Page

Page

Page

LIST OF TABLES

PAGE

LIST OF FIGURES

PAGE

NOMENCLATURES

C_p	Specific heat capacity of the stream
CS	Control Structure
CVs	Controlled variables
DOF	Degree of Freedom
IAE	Integral of the Absolute value of the Error
K_u	Ultimate gain
MVs	Manipulate variables
P	Proportional controller
PI	Proportional -integral controller
PID	Proportional–integral–derivative controller
P_u	Ultimate period
PWC	Plantwide control
SOC	Self-optimize control
TPM	Throughput Manipulator

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

Previously, process control system is designed based on control of individual unit, but when all control loops are connected and work together, they cannot control the process because control loop has interacted with each other. The reason is that there are material and energy recycle in the complex process for improving economics but there are also difficulties in process control. The designed process control by using plantwide method can solve the problem because plantwide method is designed from overview of the entire plant.

The development of plantwide process control methodology can be classified based on heuristics, mathematical, optimization-based and mixed approaches. Mainly the purpose of all method is to design control structure to overcome the disturbances and maintain production objective, normally are process economics, plant reliability, environment effect and safety.

The heuristic-based methodologies are found to be attractive as they are easier to understand and implement. One of the most popular heuristic based methods to date is Luyben et al. 2 They proposed a comprehensive nine-step procedures that ranks control and operational objectives based on their importance. However, one major disadvantage of the heuristic-based methods is relying on experience. For self optimize control is mathematic base methodology. This method is systematic and rigorous which involves extensive computation as part of the local linear analysis and evaluation of the loss. Another problem by relying on mathematical tools is that the use of such controllability analysis tools to determine the controlled variables might result in those that are easier to control, rather than

those that are important to control. In this paper, apply Wongsri (2011), mixed approach to design control structures of styrene process and compare performance with the IFSHO method of Vasudevan et al. (2011).

1.2 Research Objectives

The objectives of this research are to design and evaluate plantwide control structures of styrene process using the new design procedure of Wongsri (2011).

1.3 Scope of research

The scopes of this research can be listed as follows.

1. Design plantwide control structures of styrene process using the new design procedure of Wongsri (2011).
2. Compare performance new design control structure with IFSHO method proposed by Vasudevan et al (2011) and SOC method which proposed by Skogestad.
3. Steady state and dynamics simulation by using HYSYS.

1.4 Contribution of Research

The contributions of this research are followed;

1. Study the styrene process and concerned information.
2. Simulate the steady state of the styrene process by using HYSYS
3. Design new plantwide control structures followed Wongsri (2011) design procedure.
4. Simulate the dynamic of the styrene process with control structures design.
5. Evaluate the dynamic performance of the designed control structures.
6. Analyze of the design.
7. Conclude the thesis.

1.5 Research Framework

The thesis matter is classified six chapters as follow;

Chapter I: It 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.

Chapter III: Covers some background information about plantwide and theory concerning with plantwide control fundamentals, control issues for distillation column and Wongsri plantwide control design procedure.

Chapter IV: Describes process description and steady state conditions.

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

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

This is followed by:

References

Appendix A: Styrene Process Stream and Equipment

Appendix B: Parameter Tuning of Control Structures

Appendix C: Detail of distillation control structure

Appendix D: Research review of Wongsri

CHAPTER II
LITERATURE REVIEW

Many methodologies for PWC system design have been proposed since the early 1990s. They can be systematically classified in two ways: based on the main approach in the method (approach-based classification) or based on the control structure employed (structure-based classification). There are four groups of classification based on approach, namely: heuristics (process oriented); mathematical (model oriented); optimization (algorithmic) and mixed methods. Structure-based classification divides into three groups, namely decentralized, centralized and mixed strategies. A detailed classification and review of the various PWC methodologies up to 2009 is provided in Vasudevan (2009).

The heuristics-based procedure proposed by Luyben et al. (1997, 1998) is the first complete procedure that generates an effective PWC structure for an entire process. The comprehensive nine-step heuristics procedure ranks control and operational objectives based on their importance. A unique solution is not produced as the design problem is open-ended. Luyben et al. (1997, 1998) applied their proposed procedure to the TE, HDA and vinyl acetate monomer (VAM) plants.

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 multivariables 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.

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 optimization and simulation. It was shown that the 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 the HDA process to select appropriate 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 in alkylation process. Then design eight plantwide control structures for the alkylation process using the 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 lower of total energy use.

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 heuristic 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.

Vasudevan et al. (2011) proposed modified of the Integrated Framework (IF) call IFSHO has an eight-level framework incorporated heuristics and optimization, together with the use of simulation throughout the procedure by apply with styrene process. The performance of the resulting control system is compared with the control system developed using the integrated framework of simulation and heuristics. The control structure found stable and robust in the face of disturbances.

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 cost, energy costs and raw material costs. The main design optimization variables in this process are the steam-to-EB ratio, reactor inlet temperature, and reactor size. New design of Luyben increase low pressure steam from 3,400 kmol/h to 4,000 kmol/h and decrease reactor temperature from 650°C to 560°C.

Kedsuda (2011) apply Wongsri (2009) method to design plantwide control structure of the styrene process by ethylbenzene dehydrogenation. The method used “Fixture Point Theorem” to select control variable and manipulate variable. She designed four differences control structure which is differentiated on distillation column section. From assessment of all control structures found that the designed control structure can handle disturbances and maintain better quality product than other control structures as compared by integral absolute error (IAE). In this paper Wongsri Design Procedure (2012) is applied to the styrene plant case study presented in the work of Luyben (2011) to demonstrate its effectiveness for developing a viable and stable control structure with good performance in the face of disturbances. The performance of this control structure is then compared with that developed by SOC and IFSHO using IAE and DDS.

CHAPTER III
THEORY

In this chapter, propose the fundamental to the design of plantwide control structures that consist of control degree of freedom of process, basic distillation control, and basic plug flow reactor control and relay feedback tuning method.

3.1 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 the entire plant consisting of many interconnected unit operations

3.2 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.

3.2.1 Material recycles

The 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 $\longrightarrow$ B $\longrightarrow$ 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.

3.2.2 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. Doe energy-intensive processes, the savings can be quite significant.

3.2.3 Chemical component inventories.

In chemical process can characterize a plant 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 the 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

3.3 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 flow

rates of the liquid and gas process streams. No controller tuning or inventory sizing is done in 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 establish 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 achieve able 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 flow rates 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 a 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 reaction types such as $A+B \longrightarrow$ product. In systems with consecutive reactions such as $A + B \longrightarrow M+C$ and $M+B \longrightarrow D + C$. 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 the 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 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 is 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:

$K_{ZN} = K_u/2.2$ $\quad$ $\mathbf{T}_{ZN}$ $P_u/1.2$

K_{TL} $K_u/3.2$ $\quad$ $\mathbf{T}_{TL}$ $=2.2P_u$

3.4 Degree of freedom

$$\mathbf{f = V - E}$$

Whenf = degree of freedom

V= total number of independent process variables

E=total number of independent equations

We consider in 3 cases that f=0 number of equation equal number of variable meaning is the process is exactly specified but if f<0 number of equation more than number of variable meaning is the process is over specified which we need to remove f number of equation. The most cases are f>0 number of variable more than the number of equation meaning is under specified which we need the f number of additional equations. There are two options of f number of equation first we are specifying more number of disturbance variable such as temperature and pressure of process or specific more number of control variables. However, Luyben suggest that degree of freedom is number of valves in the process.

3.5 Basic Distillation Control

Figure 3.1 shows a simple two-product distillation column and gives the notation we use for flow rates, compositions, and tray numbering. Feed is introduced in tray N_F, numbering from the bottom. There are N_T trays in the column. The molar flow rate is F, its composition is Z_j (mole fraction of component j), and its thermal condition is q (saturated liquid is q=1, saturated vapor is q=0). The heat transfer rates are Q_R in the reboiler and Q_C in the condenser. Distillate product is produced at a molar flow rate D with composition $X_{D,\,j}$. Bottoms product is produced at a molar flow rate B with composition $X_{B,\,j}$. Reflux and vapor boilup molar flow rates are R and V, and the reflux ratio is $RR = R/D$

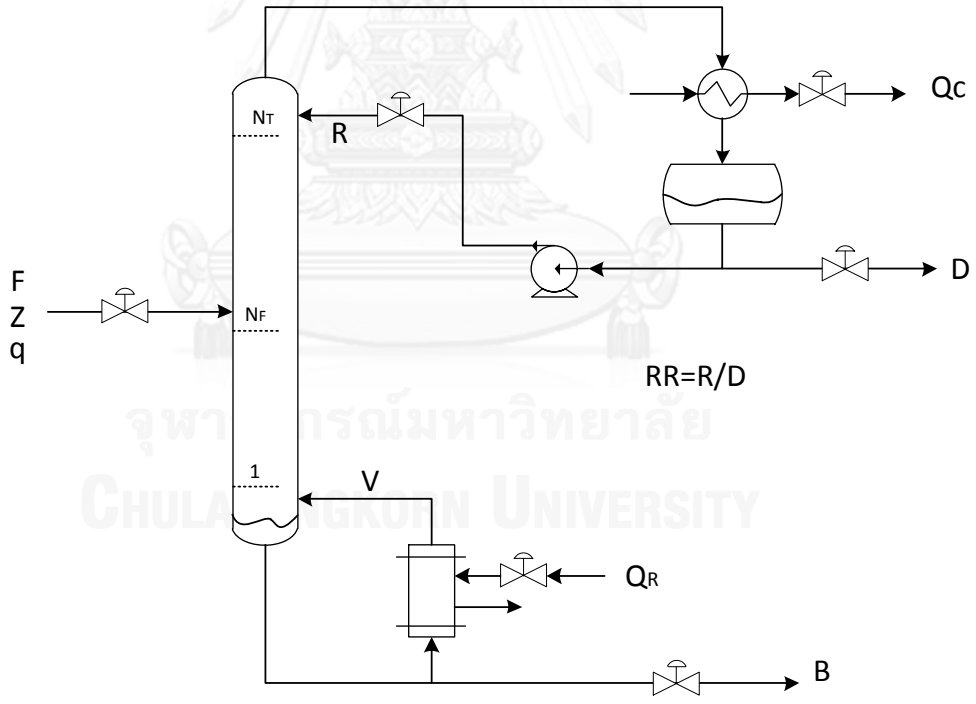


Figure 3.1 Six degree of freedom of basic distillation control

There are six control valves associated with the column, therefore there are sixes control degree of freedom. In case of supply constraint, set feed is throughput manipulator and name “on supply” but if product constraint set product flow is throughput manipulator and name “on demand”, however we can set any constraints of process to throughput manipulator such as re-boiler duty or cooling duty. One of the degree of freedom use to control condenser level by selecting one of these manipulate variables distillate flow, reflux flow, vapor boil-up, condenser cooling and feed if partial vapor. To use vapor boil-up to control condenser level, column should not too high due to there is a delay time, however distillate flow use to control condenser level and if high reflux ratio (RR > 4) use reflux flow to control condenser level base on “Richardson’s rule”. For reboiler level or bottom level, we select one from these manipulate variables bottom flow, vapor boil-up or reboiler heat duty, feed flow and reflux flow. To use the feed and reflux flow for control bottom level, it has some delay time depend on tall of the column. Next control variable is column pressure by selecting one from following manipulators condenser cooling duty, feed if partial vapor, reboiler duty. Remaining two degrees of freedom that will use to control composition or temperature of the column, but cannot fix flow rate of distillate or product due to fail to control when feed composition change.

The standard terminology of control structure by use two manipulates variables for control temperature or composition. The control structure controls both of temperature and composition of column call “Dual Composition Control” and if control only composition or temperature call “Single-end Composition Control”. Figure 3.2 shows *R-V* Control Scheme uses reflux to control distillate composition and vapor boil-up to control the bottom product composition.

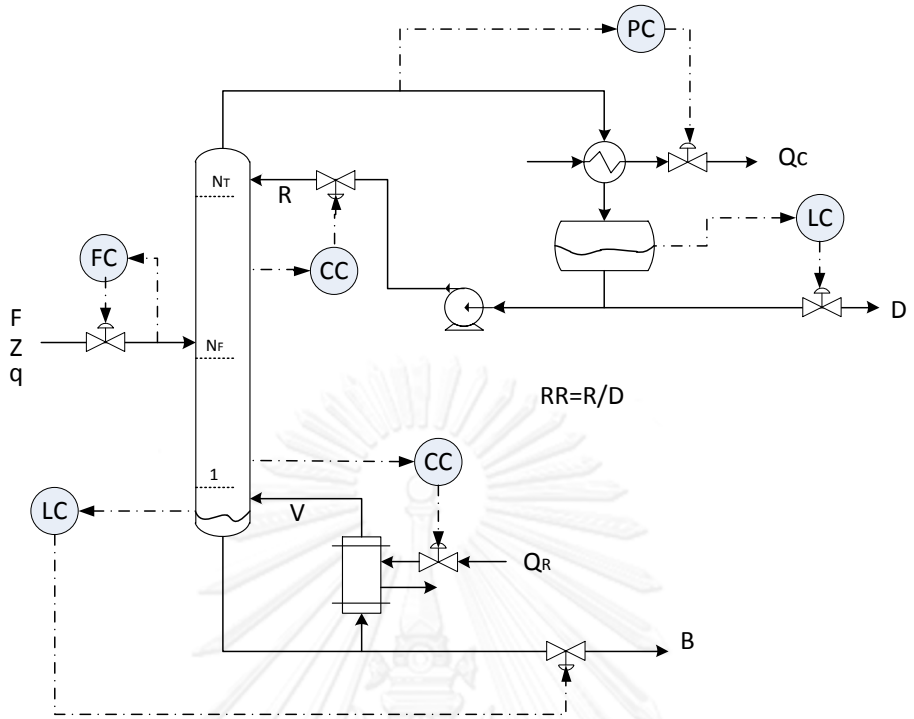


Figure 3.2 R-QR Distillation Control Scheme

When the reflux ratio more than four (RR>4) distillate flow too low to control condenser level so use reflux flow to control condenser level and distillate flow control composition.

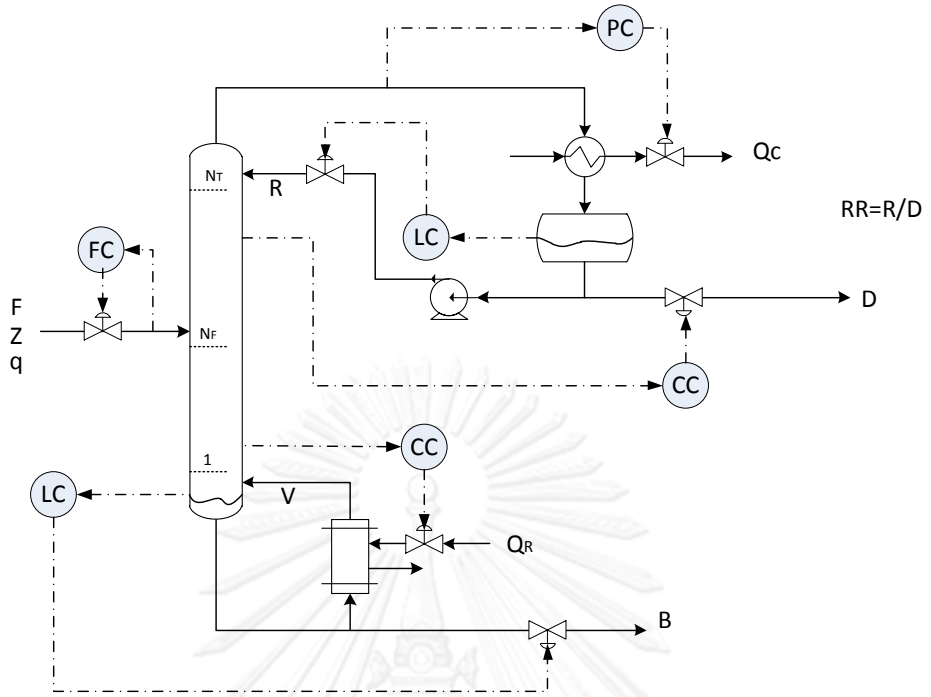


Figure 3.3 D-QR Distillation Control Scheme

However, when column has high boil-up rate uses bottom heat duty to control condenser or bottom level and us bottom flow to control composition call *R-B* Scheme.

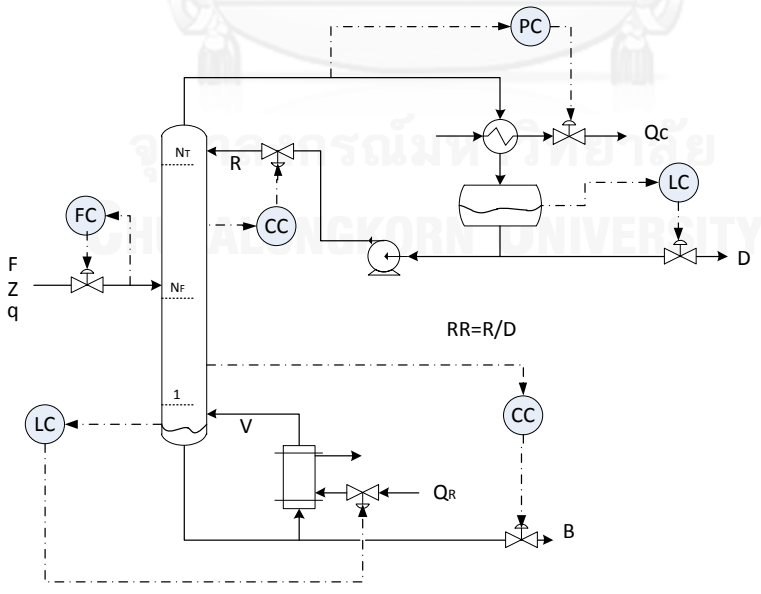


Figure 3.4 R-B Distillation Control Scheme

When the feed has low volatility distillation column need more trays and high reflux ratio, so use reflux flow to control condenser level and boil-up flow or re-boiler heat duty to control bottom level, use distillate flow and bottom flow control composition call *D-B* control scheme which is “Dual composition control”.

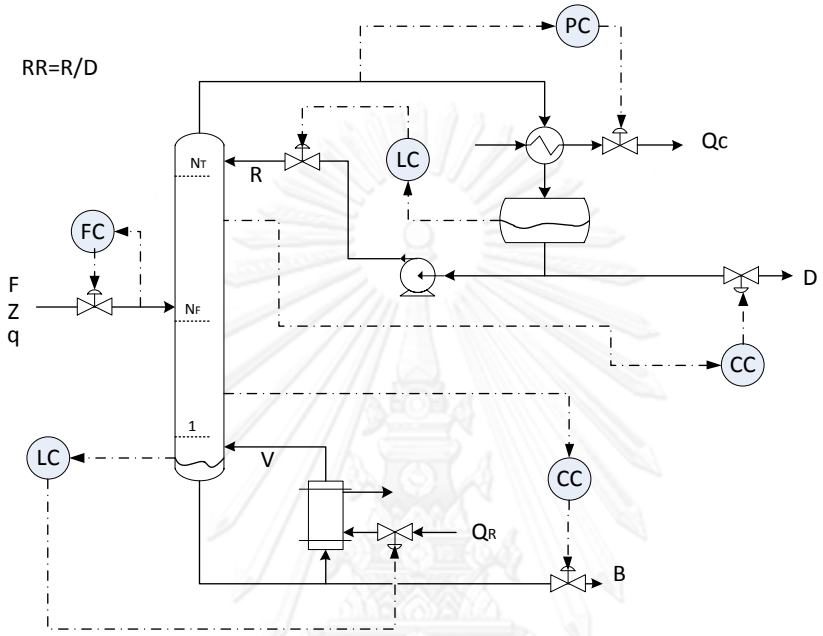


Figure 3.5 D-B Distillation Control Scheme

3.6 Dynamics Performance Criteria

Plantwide control structure design has received some attention from researchers and criteria to compare performance of control structure very important. Two aspects of control structure performance are stability and economic. The most popular of stability criteria are integral absolute error (IEA) and process-settling time and for economic aspect is profit per kilogram of product. *Konda and Rangaiah* have developed the new dynamic performance criteria that use in this research called dynamic disturbance sensitivity (DDS). DDS is equal to the sum of absolute accumulation of all the components in the process since the occurrence of the disturbance. DDS is more realistic as it is a dynamic measure and includes level and pressure effects as opposed to other steady-state measures, but the major limitation of DDS and many other measures is that they do not include the economic

quantification of dynamic performance. The DDS makes use of the strong correlation between the overall control system performance and the sum of the individual component accumulations.

$$DDS = \int_{t=0}^{ts} \left(\sum_{i=1}^{n} (absolute\ accumulation\ of\ component\ i) \right) dt$$

Where t_s is the time taken for the process to reach steady state and a smaller value of DDS indicates better control.

DDS gives an indication of the dynamic performance of the entire plant, but on the performance of the individual unit when apply the DDS will call unitwise DDS. The computation is the same as above, except that we now consider the accumulation of all the components in a single unit rather than the whole plant.

3.7 New Plantwide Control Structure Design (Wongsri's Design Procedure)

The plantwide control structure design is indeed a structural decision about placing control loops throughout the plant to achieve the design plantwide objectives. There are two levels: plant level and unit level designs. The proposed design procedure is carried out in five stages, the major design stage deal with plant level design.

The new plantwide control design procedure emphasizes on maintaining the nominal plant operating conditions, i.e. establishing a fixture plant. The amount of components is accounted by regulating the material quantifiers, which are places indicating the amount of the components using their handlers. The term material quantifier is used here to denote a more general term than the material inventory. Secondly, the procedure deals with the disturbances entered and occurred in the process using the proposed material and energy disturbance management.

Stage 1. Plant Information and Analysis.

Step 1: Gather of relevant plant information and control objectives including constraints for control. It is necessary to obtain all information relevant to process control, such as product quality, production rate, smooth operation, process and equipment constraints, plant safety, and environmental regulations.

Step 2: Plant analysis. Several tasks to assist design decision in Step 2 are:

2.1 Control degree of freedom (CDOF). Each single independent stream, physical or virtual, material or energy, must have a handle or one control degree of freedom.

2.2 Heat pathways. The first pathway is heat generated by exothermic reactions and flows out to the environment. A second pathway carries heat from utilities into the process and to the environment. The third pathway is heat flow in process loops, internal to the process. The fourth pathway is accounted for the enthalpies entered and left the plant.

2.3 Material pathway is the flow path of a component or group of components from entry or generated points to exit or end points. Along the pathway there is the quantifier of each component that will use to control inventory of the component.

2.4 Material quantifier. A material quantifier is the place indicating the significant amounts of a chemical component (or a group of components) in the plant which can be handled quite readily by regulating at their handlers. In the case that the quantifier is a flow, it is, but not necessarily, the place that has the highest gain of component flow is the total flow.

2.5 Reaction section. It is necessary to obtain required information for control design of reactor section. In general, what kind of controlled variables used to regulate the reaction yield and where to measure such controlled variables? What is the best control strategy and all? If feeds and recycle streams are fixed, the only places that the material (total or component) flow rates altered are a reactor and also a separator.

2.6 Separation section. The appropriate directions of disturbances are analyzed and specified. A plus disturbance, D+ is the plus deviation of the mass load from the nominal load and the minus disturbance, D- is the minus deviation of the mass load. The paths of D+ and D- in the separation section are analyzed and then designed in order to shift plus or minus mass loads to the desired targets to achieve the plant operation objective, e.g. maintaining product quality and avoiding disturbance propagation and recycling. The paths of D+ and D- in the separation section must be shifted to the proper exits. In the case that there is no proper exit for D+ or shifting it through available exits will disturb the product quality, recycling it would be allowed.

Next, a good location of temperature control is the tray with the largest changes in the temperature from the initial steady state by changing of composition, total flow, temperature, and component flow during keeping the reboiler heat duty and reflux flow or reflux ratio or reflux fraction or boil up ratio constants.

2.7 Production rate control. Throughput changes must be achieved by altering reactor condition. However, reactor temperature, reactant concentration, reactor holdup would be somehow limited.

The production or throughput rate change by increasing/decreasing feed rate, should be accompanied by adjusting recycle flow accordingly.

Mode of operation: On-supply, On-demand, and on-internal. The mode of operation is dictated by a business objective and the mode of operation, such as on-supply (fixed feed rate), on-demand (fixed product rate), and on-internal (fixed internal flow rate) based on throughput manipulator (TPM) decision Price and Georgakis (1994). For on-internal control scheme, the throughput manipulator (where the production rate is set) is located inside the plant downstream of this location (normally at the bottleneck), the plant has to process whatever comes in, and upstream of this location the plant has to produce the desired quantity. The selection of on-supply, on-demand or on-internal should depend on the completeness of total control of the components.

In some processes, the separation section is placed before the reactor section, there are two locations to fix the material flows into the process: at the entrances of the reactor section or of the separation section. In the case that the reactor influent is fixed, the quantifiers (inventories) prior to this point must be controlled as 'on-demand production'.

Stage 2. Fixture plant and disturbance management, Plant level loop design.

This stage is a major design stage; plant control structure is created at plant level in two steps: Step 3 and Step 4. The plant control loop design procedure presented in this paper is explicit and systematic while the Luyben design procedure has some short comings, Konda (2005). There are two objectives: the plant nominal material balance is maintained; the heat and material disturbances must be rejected to the nearest exits or directed to less significant streams.

Step 3: Establish fixture plant.

Establish fixture plant. The principal idea of establishing a fixture plant is first to have a material-balanced in the plant by controlling each component at its quantifier, i.e. fixture point.

3.1 Keep the materials entered and/or reentered fixed. A fresh feed and/or a combined stream of make-up feed and recycle stream must be kept constant to maintain the plant inventory by flow/composition controls. A recycle flow should not be fixed. This leaves the recycle flow free to be adjusted; one degree of freedom is restored to the plantwide control design process. If the composition of the recycle stream differs from the fresh feed stream significantly, each recycle stream may be flow-controlled. However, in the case that the composition of the recycle reactant can be measured, the composition of the combined stream is controlled to keep the combined reactant flow in check.

In the case of changing throughput, the combined stream of make-up feed and recycle or the recycle stream is adjusted accordingly to maintain the material balance principle. Normally, the liquid recycles is adjusted automatically by its level

somewhere in the process. However, it might be not the case for the gaseous recycle flow, the additional ratio loop of the recycle and the feed is recommended.

3.2 Adjust the flow of exit material streams (products, byproducts, and inert) according to their accumulations. If the flows of the products are controlled (mode of operation is on-demand) the quantifiers of the products, e.g. levels of reflux drums indicating the plus/minus will be used to control the feeds.

3.3 Handle the material that not leaving the process. The reactor is the logical place to regulate a component fed or formed in the process and not leaving the process. If there is only one reactor and there is more than one component that not leaving the process, their kinetics must be similar, e.g. increasing the reactor temperature reduces or increases the amount of both components. Handlers of these components must be identified. If their kinetics are not compatible, we must provide exits for the incompatible components.

3.4 Control the amount of the rest of the component at their quantifiers. This step assures the rest of component inventory is regulated from a plantwide perspective. Setting the control at the specified quantifiers is like providing coordination over different sections of the plant to ensure that the rate of accumulation of each component in the overall process is zero.

3.5 Maintain the production rate.

3.5.1 Consume the limiting reactant. Determine the most appropriate manipulate variable to control the limiting reactant for the economic reason, i.e. the reactor temperature, the reactor pressure, or the reactor holdup.

3.5.2 Maintain the production rate. The product rate can be maintained through 3.2.1. If this is done and the production rate does not reach the objective or the production demand, the limiting reactant feed rate must be increased. However, the design constraints may limit this strategy concerning increasing the reactant feed rate.

Step 4: Disturbance Management for Quality Control.

Disturbance management for quality control. The nominal conditions of process streams are maintained by specifying the disturbance shifting directions. The principles of disturbance management are following:

4.1 Heat disturbance management. The heat disturbance is divided into two categories. Heat disturbance of category 1 (HDC1) is the heat disturbance that does not instantly affect on the qualities of process streams, such as heat disturbance in a process stream toward a heater, a cooler, or a process-to-process heat exchanger. Heat disturbance of category 2 (HDC2) is the heat disturbance that will affect the process stream qualities where an additional phase is created or introduced, and the equilibrium is altered; or where chemical reactions are undergoing, such as separators and reactors.

4.1.1 Direct the HDC1 to the environment via the next and nearest exit points, usually heaters or coolers, to keep the thermal conditions of the process stream fixed.

4.1.2 Direct the HDC2 to less significant output streams of separators. This rule is generally apt to a separator using heat as a separating agent.

4.2 Material disturbances management (MDM). The configurations of the control loops are decided base on the desired material pathways. As in the case of heat disturbance management, we should direct the material disturbances to the environment via the next and nearest exit points to avoid disturbance recycling and propagation.

Many industrial distillation columns use some type of single-end temperature control because of its simplicity and low maintenance cost. This step presents a procedure to determine the control structure of a distillation column with desired material disturbances (D+ and D-) following step 2.6 by using a dynamics process simulator for various single-end control structures, namely constant reflux flow (R), constant reflux ratio (RR), constant reflux-to-feed ratio (R/F), constant reflux fraction (R/(R+D)), constant boil-up ratio (V/B). Several kinds of material disturbances in feed,

such as temperature, flow rate, composition, and component flow rate are generated to test the disturbance shifting ability of these control structures. In addition, the principals of the material disturbance management are as follows:

4.2.1 Direct the material disturbances of byproducts, inerts, and unconverted raw materials to the environment via the next and nearest exit points.

4.2.2 For the main products, the minus disturbances should follow Rule 4.2.1. However, the main product plus disturbances should be allowed to propagate to their exits.

4.2.3 MDM rule for the recycle streams: their plus disturbances of unreacted raw materials are permitted, however, their minus disturbances must not be allowed to economize the make-ups.

The selection of the distillation control structures is carried out in two steps: preliminary screening using steady-state simulation and the selected candidates are further tested by rigorous dynamic simulation.

Stage 3. Unit Level Loop and Enhanced Loop Designs

Control loop design at this stage is solely based on individual unit operations.

Step 5: Design the rest of the control loops. Normally, the rest of the control loops is inventory loops which are self-regulating and less crucial. They can be designed using unit-based approach.

5.1 Design the control loops for the remaining control variables, i.e. the rest of the inventory.

5.2 Adding simple enhanced controls, e.g. cascade, feed forward controls.

Stage 4. Energy Management and Optimization.

The supplementary design activities involve heat exchanger network design and control, and plant operation and design optimization.

Step 6: Energy management via heat exchanger networks.

In the case that the exothermic heat of reaction is large enough to heat some process cold streams, i.e. potential heat exchanger networks or alternative heat integrated processes (HIPs) exist, a heat exchanger network must be designed and a HEN must be resilient, i.e. delivering the exchange streams to their target temperature. The resilient heat exchanger network with specified load disturbances can be designed using Wongsri's method [33]. The design of a control system that prevents the propagation of the heat disturbance of Wongsri and Hermawan is recommended.

Step 7: Optimize economics or improve control performance.

Since the design of a chemical process evidently affects its control performance dynamically, another part of the problem's open-ended nature is the opportunity to alter the process design. The design and control issue remains as yet an open research area in terms of the plantwide control design.

Stage 5. Design Validation.

The validation of the design control structures using rigorous nonlinear simulation is inevitable, whatever may be the design procedure.

Step 8: Validate the designed control structures by rigorous dynamic simulation. The measures would be costs, raw material and energy consumptions, control performances of the total plant or some selected loops, etc. Expected disturbances must be listed to perform the disturbance test on the plant with designed control structures.

CHAPTER IV
STYRENE PROCESS

4.1 Introduction

Styrene is an important chemical that use to produce polystyrene and ABS. Two processes to produce Styrene are Ethylbenzene dehydrogenation and PO/SM process and more than 60% of the Styrene produce by Ethylbenzene dehydrogenation process.

4.2 Reaction Kinetics

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

Styrene production

$$C_6H_5CH_2CH_3 \rightleftharpoons C_6H_6CHCH_2 + H_2 \quad (1)$$

(Ethylbenzene) (Styrene) + (Hydrogen)

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

$$C_6H_5CH_2CH_3 \rightarrow C_6H_6 + C_2H_4 \quad (2)$$

(Ethylbenzene) (Benzene) + (Ethylene)

$$C_6H_5CH_2CH_3 + H_2 \rightarrow C_6H_5CH_3 + CH_4 \quad (3)$$

(Ethylbenzene) + (Hydrogen) (Toluene) + (Methane)

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

$$2H_2O + C_2H_4 \rightarrow 2CO + 4H_2 \quad (4)$$

(Water) + (Ethylene) (Cabonmonoxide) + (Hydrogen)

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

$$H_2O + CH_4 \rightarrow CO + 3H_2 \quad (5)$$

(Water) + (Ethylene) (Cabonmonoxide) + (Hydrogen)

$$H_2O + CO \rightarrow CO_2 + H_2 \quad (6)$$

(Water) + (Cabonmonoxide) (Cabondioxide) + (Hydrogen)

Table 4.1 Reaction Kinetics. (Luyben, 2011)

Reactions	k	E (kJ/kmol)	Concentration
(1) Forward	0.044	90,981	P_{EB}
(1) Reverse	6×10^{-8}	61,127	$P_S P_H$
(2)	27,100	207,989	P_{EB}
(3)	6.484×10^{-7}	91,515	$P_{EB}P_H$
(4)	4.487×10^{-7}	103,997	$(P_W)^2P_E$
(5)	2.564×10^{-6}	6,723	$P_W P_M$
(6)	1,779	73,638	$P_W P_{CO}$

Overall reaction rates have units of kmol s^{-1} 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 (LPS1) are initially mixed to convert to gas phase before preheat 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 heated in a furnace and mixed with the preheated mixture. Then, mixture fed to adiabatic PFRs in series with a heater in between for control reactor feed temperature.

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 sending to the three-phase separator, where lighter 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 operated 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 recycling back to the reaction section.

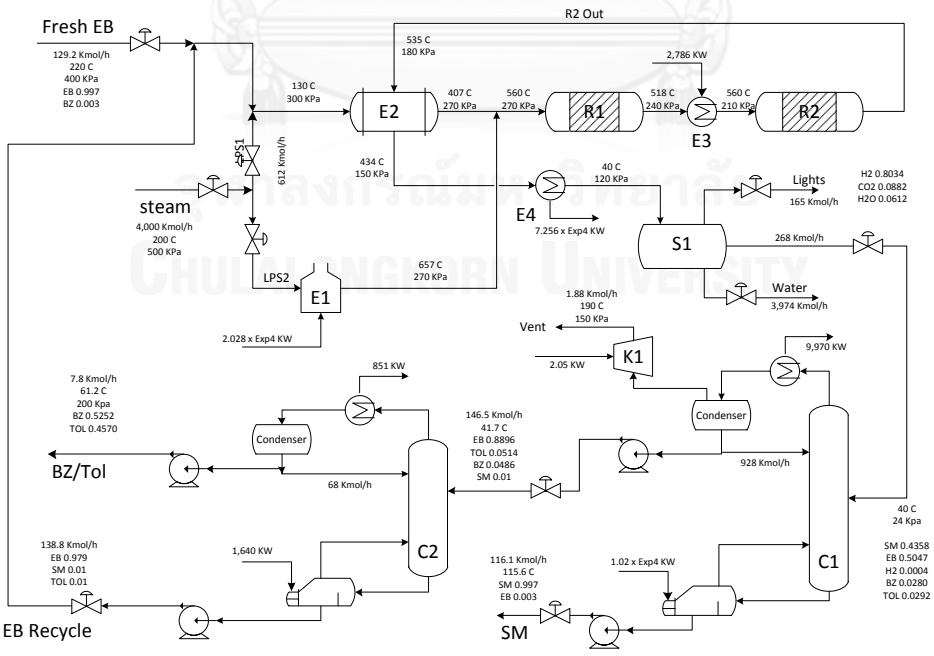


Figure 4.1 Flowsheet for styrene process

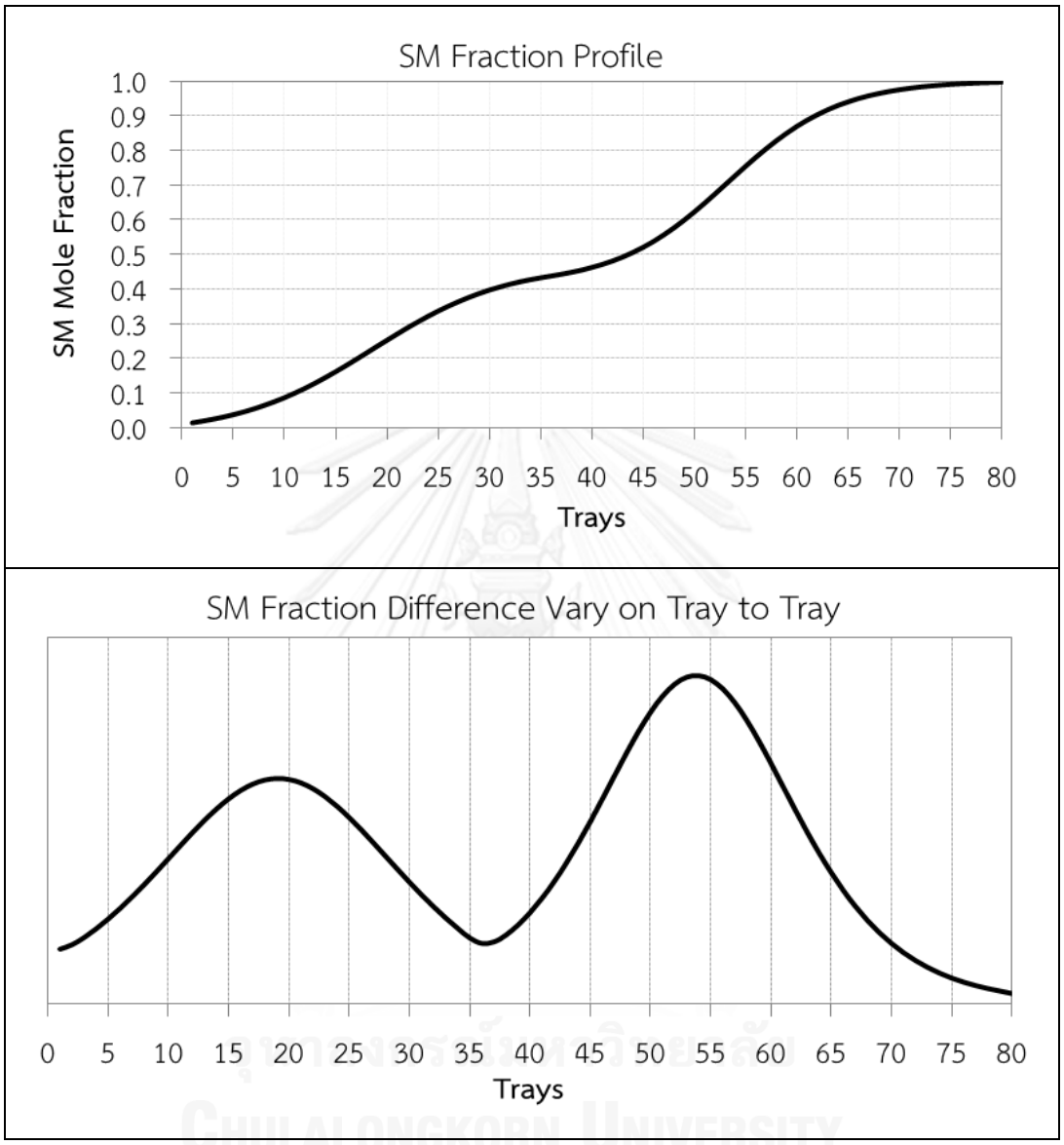


Figure 4.2 Column C1 SM fraction profile and difference.

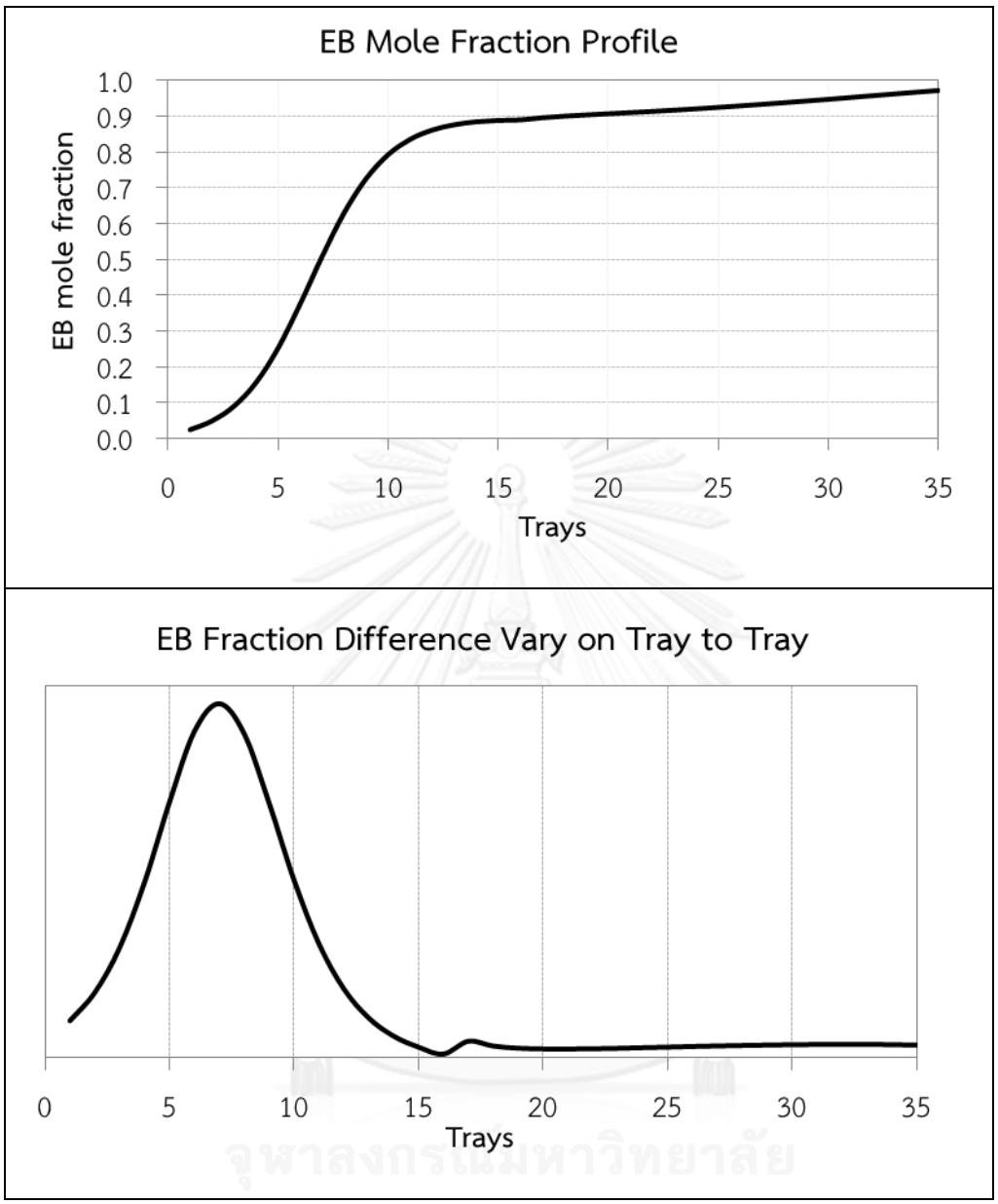


Figure 4.3 Recycle column C2 EB fraction profile and difference

4.4 Evaluation of cost

Styrene is produced by Dehydrogenation of Ethyl-benzene. The reaction is endothermic, non-equimolar and reversible that, production yield can be improved by high temperature and low pressure. Therefore, increase of steam ratio will suppress partial pressure of the reactants and the conversion will be improved.

The important conditions of this process are steam to EB ratio and reactor inlet temperature. We can increase yield by an increase steam ratio, but trade off with high utility cost to increase the steam temperature to superheat however If we suppress side reactions by decrease reactor temperature, EB recycle will increase meaning is process consume more energy to separate of EB.

Luyben (2011) purposed optimum condition for the process buy keep steam to an EB ratio of 14.34 while fresh EB feed 132.8 Kmol/h and keep the reactor inlet temperature of 560°C.

Table 4.2 Base case economic calculation

Product	**Price, $/Kg**	**Mass Flow, Kg/h**
Styrene	0.8	12,122
Ben/Tol	0.325	1,193
Light Out	0.05	2,113
Feed	**Cost, $/Kg**	**Mass Flow, Kg/h**
Ethylbenzene	0.4	14,692
LPS	0.013	72,060
Utility	**Cost, $/KJ**	**Energy, KJ/h**
Electricity	0.0000168	62,403
Cooling water	0.000000354	50,051,617
Fuel	0.000006	72,995,541.2
Steam Utility	0.00000778	50,051,617

4.5 Steady State Evaluation

The steady state model is built in HYSYS. The flowsheet information is obtained from Luyben (2011) and Kedsuda (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.

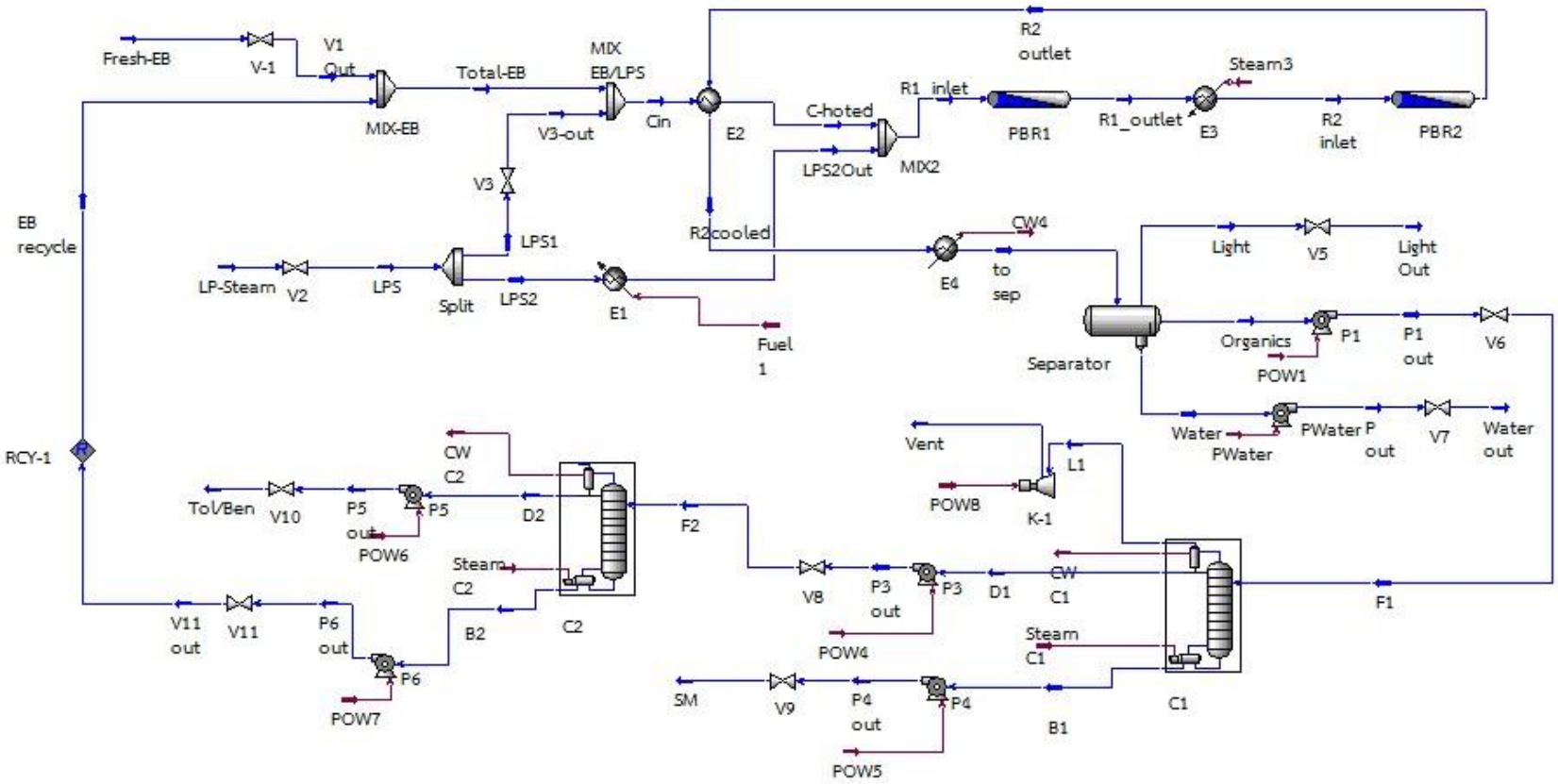


Figure 4.4 Presents the steady state flowsheet built in HYSYS.

CHAPTER V
CONTROL STRUCTURES DESIGN

Design of plantwide control structure depends on knowledge and experience of the designer. Complex processes have many control degrees of freedom so difference control structures design can be reached. Recently, research to apply new design methodology to any processes, however important is performance of control structure. Wongsri's design procedure has 5 stages of guideline which, the designer clearly and easy to follow. Overviews of the method start with process information are collected such as plant constraints and objectives of the process. The second step is analysis the process by identifying a number of control degree of freedom, heat pathway, material pathway and material quantifier. For matching of control variables and manipulate variables are in stage 2 and stage 3 which outcome is plantwide control structure and in the last two stages is improve the structure and test dynamic performance.

5.1 Design of Plantwide control structures.

Stage 1. Plant Information and Analysis

Step 1: Gather of relevant plant information and control objectives, including constraints for control.

The plant information is presented in Section III. The objective of styrene process is producing styrene monomer more than 100,000 metric tons per year or 110 kmole/h with purity 99.7%. Feed temperature of reactor R1 and reactor R2 is 560°C. The temperature of the three-phase separator should be at or 40°C and pressure 120 kPa in order to maximize organics product recovery. Steam to EB ratio around 14.93. Fresh EB feed 130 kmole/h.

Step 2: Plant analysis

2.1 Control degree of freedom (CDOF)

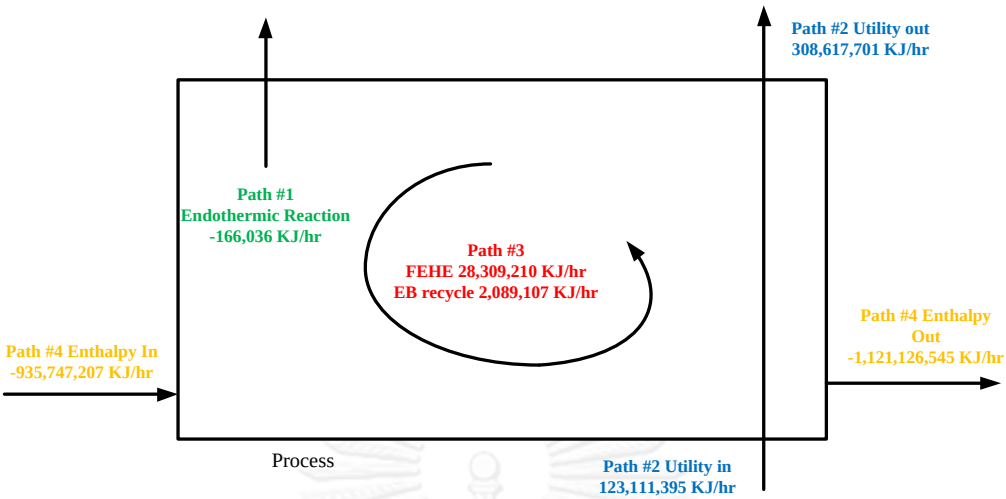
There are total of 20 independent streams, according to Wongsri simple rule, there are 20 CDOFs. They are given in Table 5.1.

Table 5.1 Detail of control degree of freedom (CDOF)

Unit	**Independence Stream**	**Quantity**	**CDOF**
Streams	Fresh EB, Fresh Steam, LPS1	3	3
Compressor	Work	1	1
Plug flow reactor	Inter-heater heat duty, furnace heat duty	2	2
Cooler	Coolant flow rate	1	1
Distillation column	Condenser heat duty, Bottoms flow rate, Distillate flow rate, Reboiler heat duty, Reflux fow rate	2	10
3-Phase separator	Light flow rate, Water flow rate, Organic flow rate	3	3
Total			20

2.2 Heat pathways

The first pathway is an endothermic heat of reactions that are consumed in the two reactors provided by furnace and inter-heater E3. The second pathway carries utility heat such as pumps and reboilers via the process streams and pass them to the environment via condensers and cooler E4. The third pathway is heat carried by the process stream loop, EB recycle. The fourth pathway is heat flow in the process via fresh feeds of EB and steam and leave the process via 5 exit streams. All four pathways are shown in Figure 5.1.



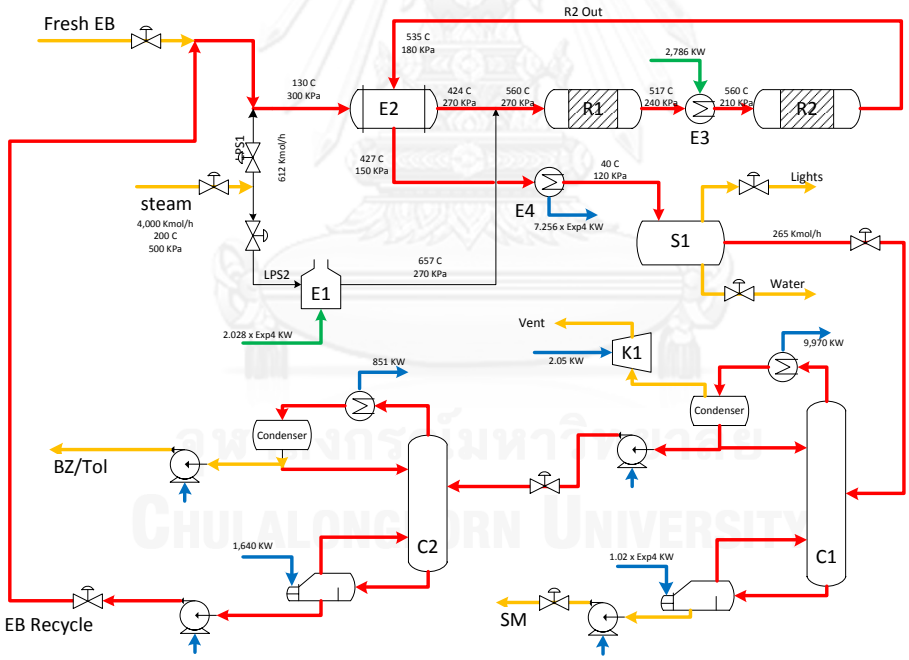


Figure 5.1 Heat pathway of styrene process

2.3 Material pathways

There are 5 material pathways of EB, steam/water, Styrene, Benzene/Toluene and Hydrogen. Benzene and Toluene are considered as a single pathway because they leave the process in a single stream. Ethyl-Benzene pathway start from fresh feed through reactors, 3-phase separator, product column, recycle column and recycle to mix with fresh EB feed as show as dash line in Figure 5.2. Low-pressure steam pathway from fresh feed to reactors and 3-phase separator finally removes from the process as water that show as dash line in Figure 5.3. Product styrene generates in reactors and move along the process to product column distillation and remove from the process at bottoms product as shown in Figure 5.4. By products benzene and toluene are generate from side reaction and remove as distillate of recycle column as show in Figure 5.5. Light pathway show in Figure 5.6 is pathway of light gas which create by side reaction and remove from process as 3-phase separator mix gas.

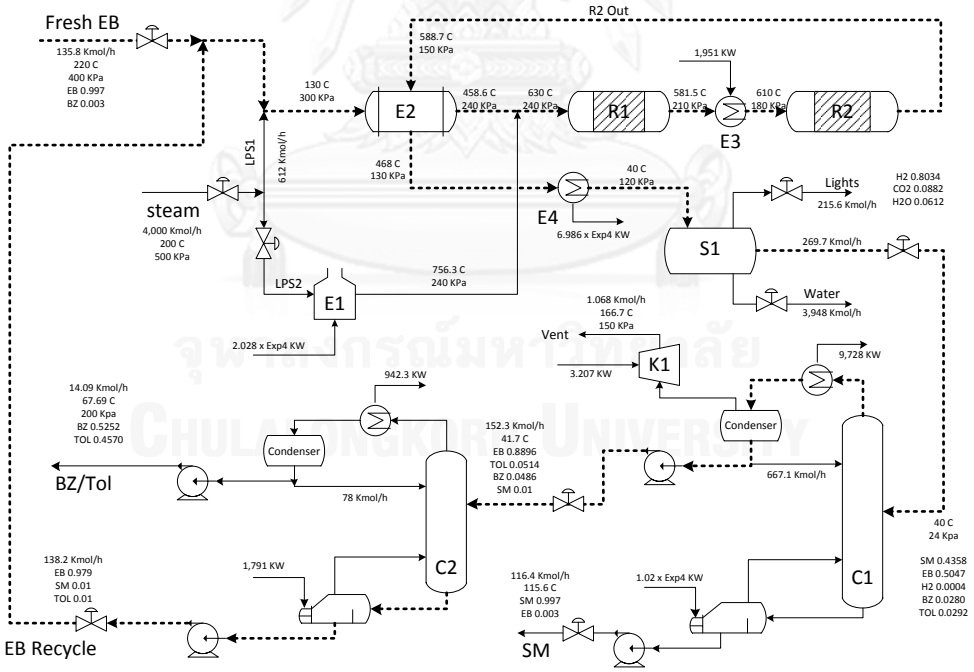


Figure 5.2 Ethyl-Benzene pathway show as dash line

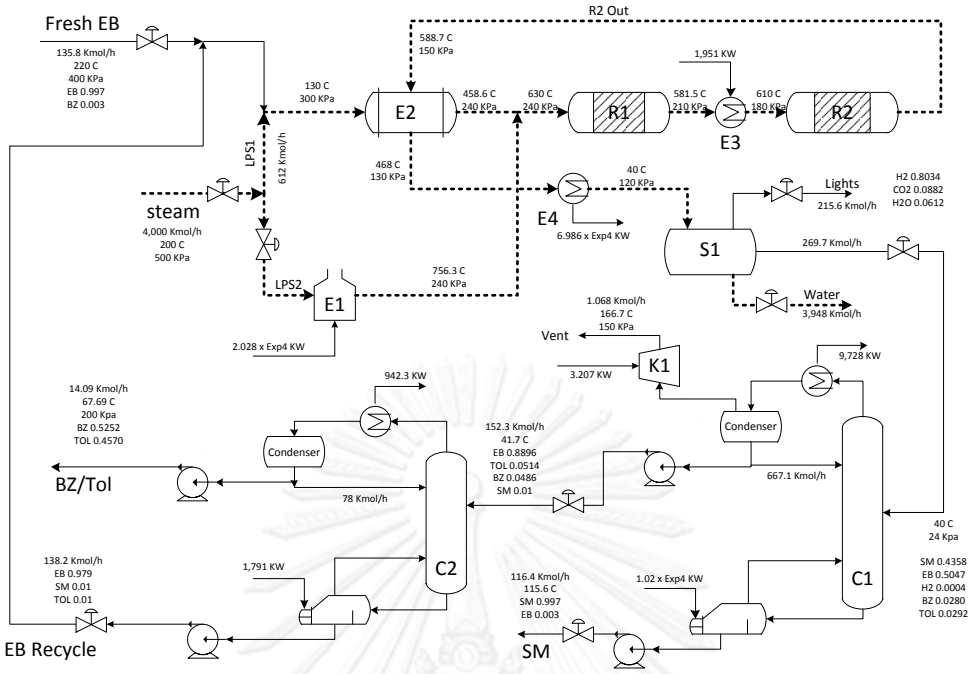


Figure 5.3 Steam pathway show as dash line

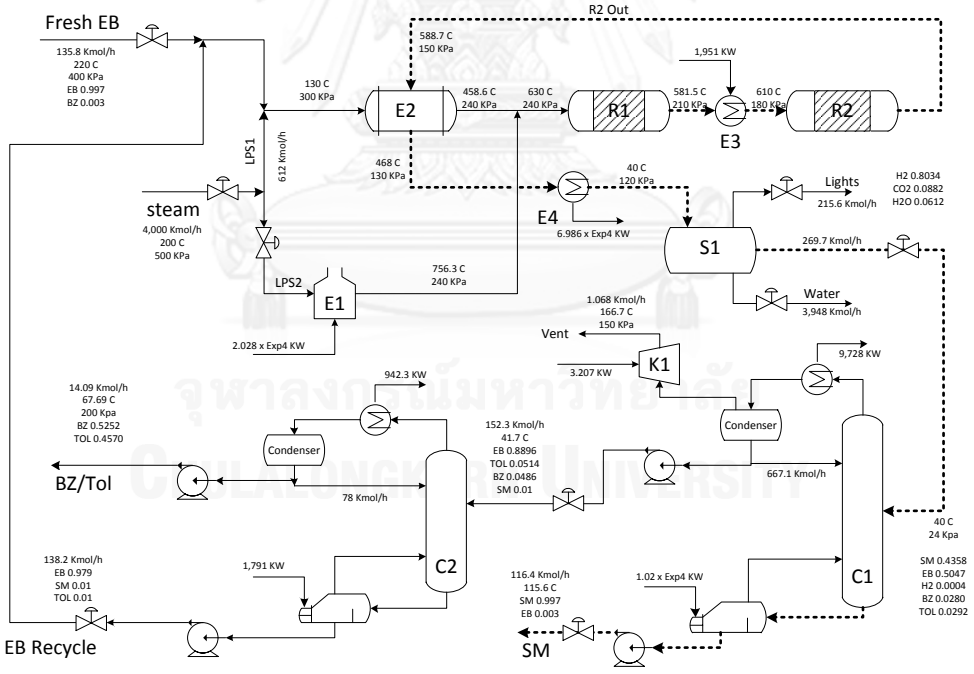


Figure 5.4 SM pathway as show in dash line

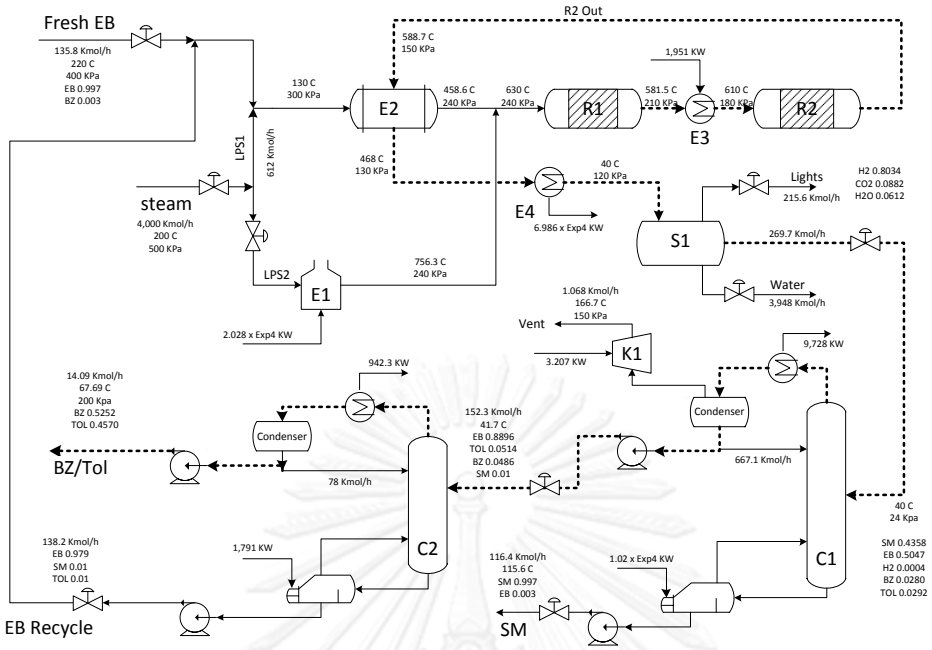


Figure 5.5 Benzene/Toluene pathway as show in dash line

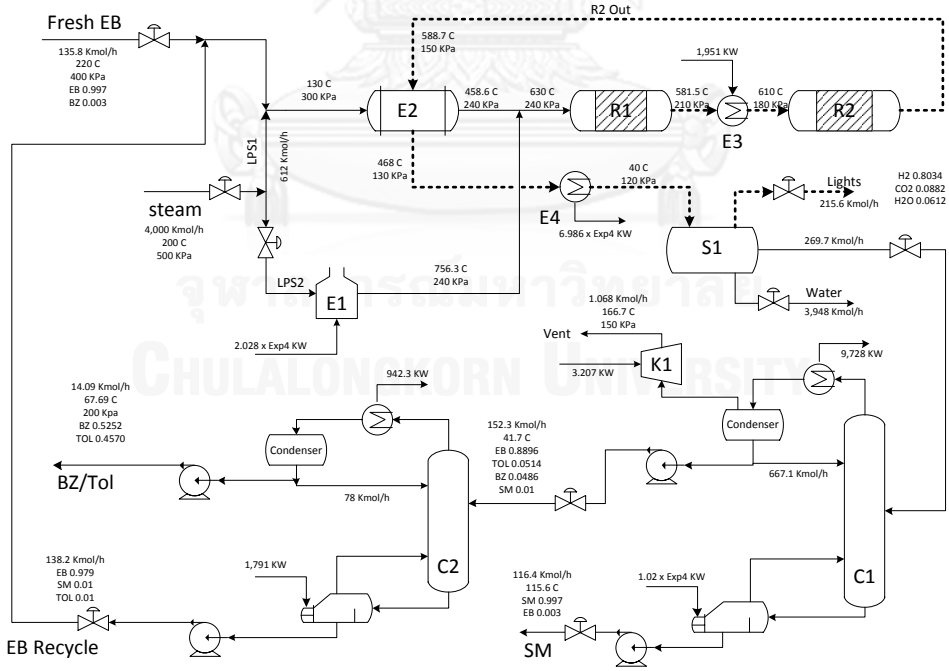


Figure 5.6 Light pathway as show in dash line.

2.4 Material quantifier

The material quantifiers are useful to design control loops for component balance as discussed in Section 3.7. The place indicating significant amounts of EB, steam, Styrene, Benzene/Toluene and Hydrogen are total flow of EB, fresh feed of steam, C1 bottom level, C2 condenser level and 3-phase separator pressure respectively. A quantifier can be flow rate, liquid level, or pressure which is more presentable for our purpose than the word inventory. Their handlers are 5 independent streams having strong cause-and-effect relationship with them, listed in Table 5.1. The 5 CDOFs are used. The quantifiers and their handlers are presented in Table 5.2.

Table 5.2 Quantifiers and handlers of components

Component	Quantifier	Handler
EB	Total flow	Fresh EB feed flow
Steam	Total flow	Fresh Steam flow
Styrene	C1 Reboiler level	Bottom flow rate
Benzene/Toluene	C2 Condensor level	Distillate flow rate
Hydrogen	3-Phase Separator pressure	Light gas outflow rate

2.5 Reaction Section

The objective of this step is to define dominant control variables to control reaction yield. In the styrene process two parameters are reactor inlet temperature and partial pressure of ethylbenzene are dominant control variables. Luyben (2011) suggested that when increase temperature conversion of ethylbenzene increase but decrease selectivity due to conversion of side reactions also increase.

In order to suppress side reactions, partial pressure of ethylbenzene has to decrease by increase low pressure steam feed rate. However Luyben (2011) purposed optimum conditions with 4,000 kmol/h of low pressure steam or ratio 14.9 of total EB and both reactor feed temperature are 560°C due to optimize margin. To

study the behavior of reactions by varying reactor feed temperatures 560°C, 600°C and 650°C, for study effect of ethylbenzene partial pressure by varying low pressure steam flow from 2,500 kmol/h, 3,200 kmol/h and 4,000 kmol/h. From the Figure 5.7 show, that high partial pressure of ethylbenzene gives high conversion to styrene when reaction starts after that when the partial pressure of styrene increase reverse reaction will take place. As a result, the dominant controlled variable should be the reactor temperature adjusted by furnace duty for reactor R1 and adjusted by inter-heater for reactor R2 by keeping the steam to EB ratio 14.93.

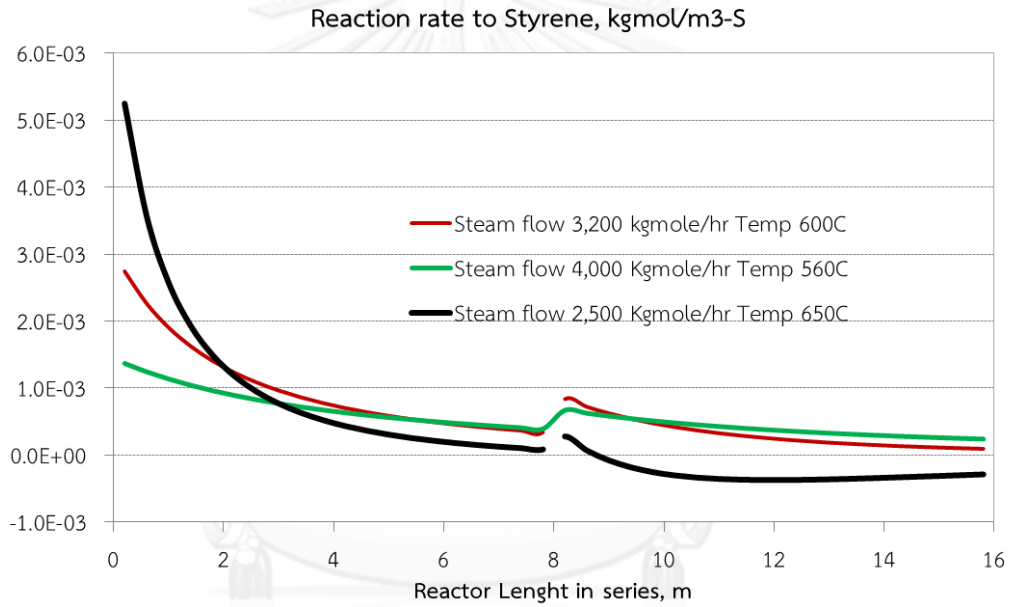


Figure 5.7 Reaction rate to styrene that vary on steam flow and reactor temperature

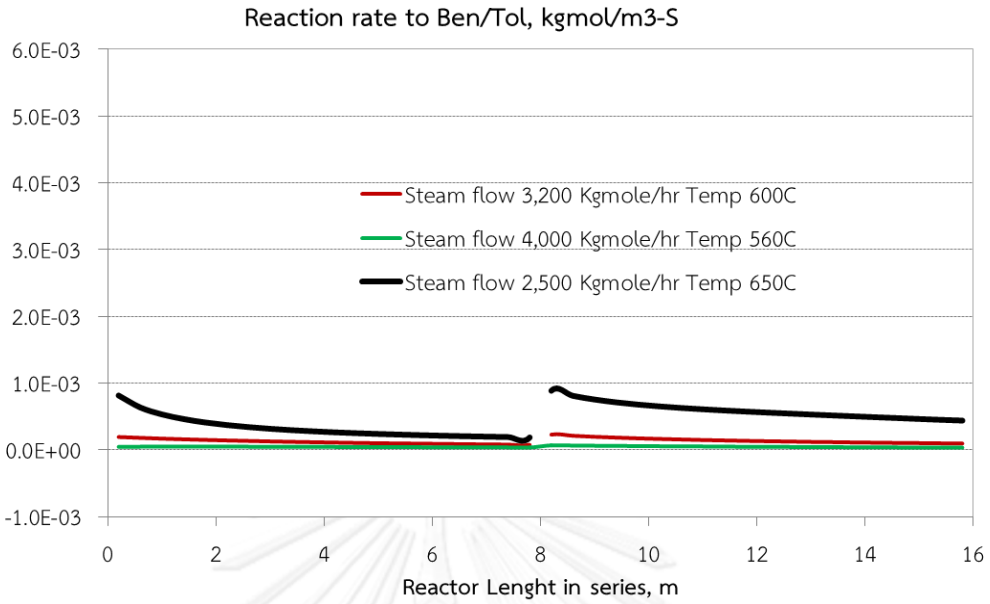


Figure 5.8 Reaction rate to Ben/TOL that vary on steam flow and temperature

2.6 Separation Section

The proper directions of material disturbances are analyzed and specified in this section. To begin with, the product column C1 separates the organic liquid into the styrene product in the bottoms and ethylbenzene, toluene, benzene in the distillate. Light key is ethylbenzene and heavy key is styrene so the bottom product is styrene rich with 99.7% minimum. The distillate is rich in ethylbenzene. The minus disturbance of styrene (SM-) should be shifted to the top of C1 and the plus (SM+) should be kept in the bottom to maximize styrene production rate. While the plus disturbance of ethylbenzene (EB+) and minus (EB-) should be directed to the top of C1 purpose to keep SM purity as specified, not give away or off specification. However, the minus disturbances of styrene cannot be shifted to the top of the column because only small amount of styrene leak to the top.

For column C2, since the top product is benzene/toluene (LK) and bottom product is ethylbenzene (HK), TOL+ and TOL- goes to the top of the column to keep EB recycle stable as possible. To maintain the EB recycle purity, EB- should go to the top and EB+ should go to bottom, but the EB availability on top is a very small portion; hence, this is not possible. The specified directions of material disturbances are shown in Figure 5.9.

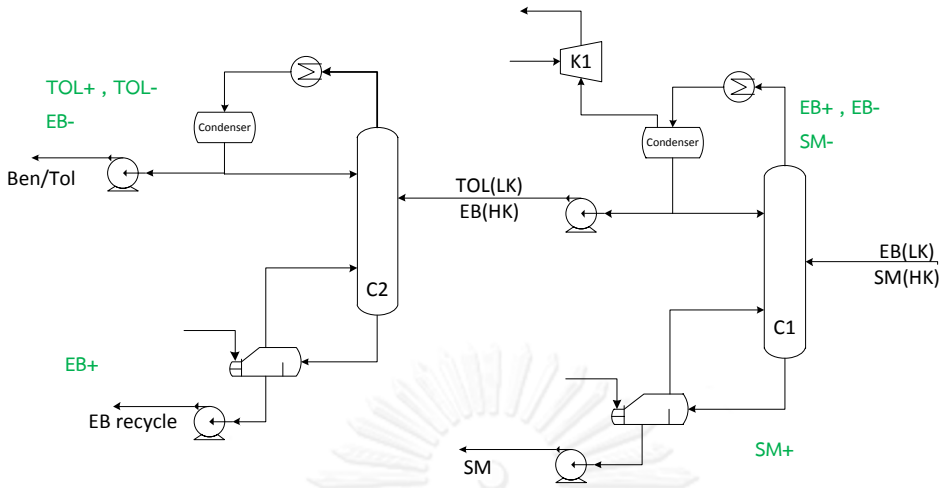


Figure 5.9 The directions of material disturbances

To select the control tray location of distillation column by sensitivity test is also determined in this step. Table 5.3 show four differences scenario for the steady state test and temperature or composition profile are collected. The tray that most sensitive on disturbances will be selected for control tray. However, sometime difference disturbances lead to difference sensitive trays so, dynamic test in Step 4.2 will be conducted for select appropriate control structure of distillation columns. Figure 5.10 shows that the most sensitive tray is tray 53 for stripping section but for rectifying section only Figure 5.10(f) reveal small sensitivity on tray 18. Figure 5.11 shows that the most sensitive tray is tray 53 for stripping section but for rectifying section only Figure 5.11(f) reveal small sensitivity on tray 18. Figure 5.12 also selected sensitive trays on tray 18 and tray 53. Normally we run sensitivity of temperature profile, but product column of styrene process operate under vacuum, so that sensitivity on composition is more appropriate.

Figure 5.14 shows a sensitivity test of column C2, found that sensitive tray locate on the top section and for stripping section only found sensitive on tray 28 in Figure 5.14(d) and Figure 5.15(d).

Table 5.3 Four-sensitivity test of distillation column

Sensitivity	*Fixed*	*Vary*
1	Qr and Reflux Flow	Total feed flow, Feed temperature, Component flow and Feed composition
2	Qr and Reflux Ratio	Total feed flow, Feed temperature, Component flow and Feed composition
3	Qr	Reflux Ratio
4	Reflux Ratio	Qr

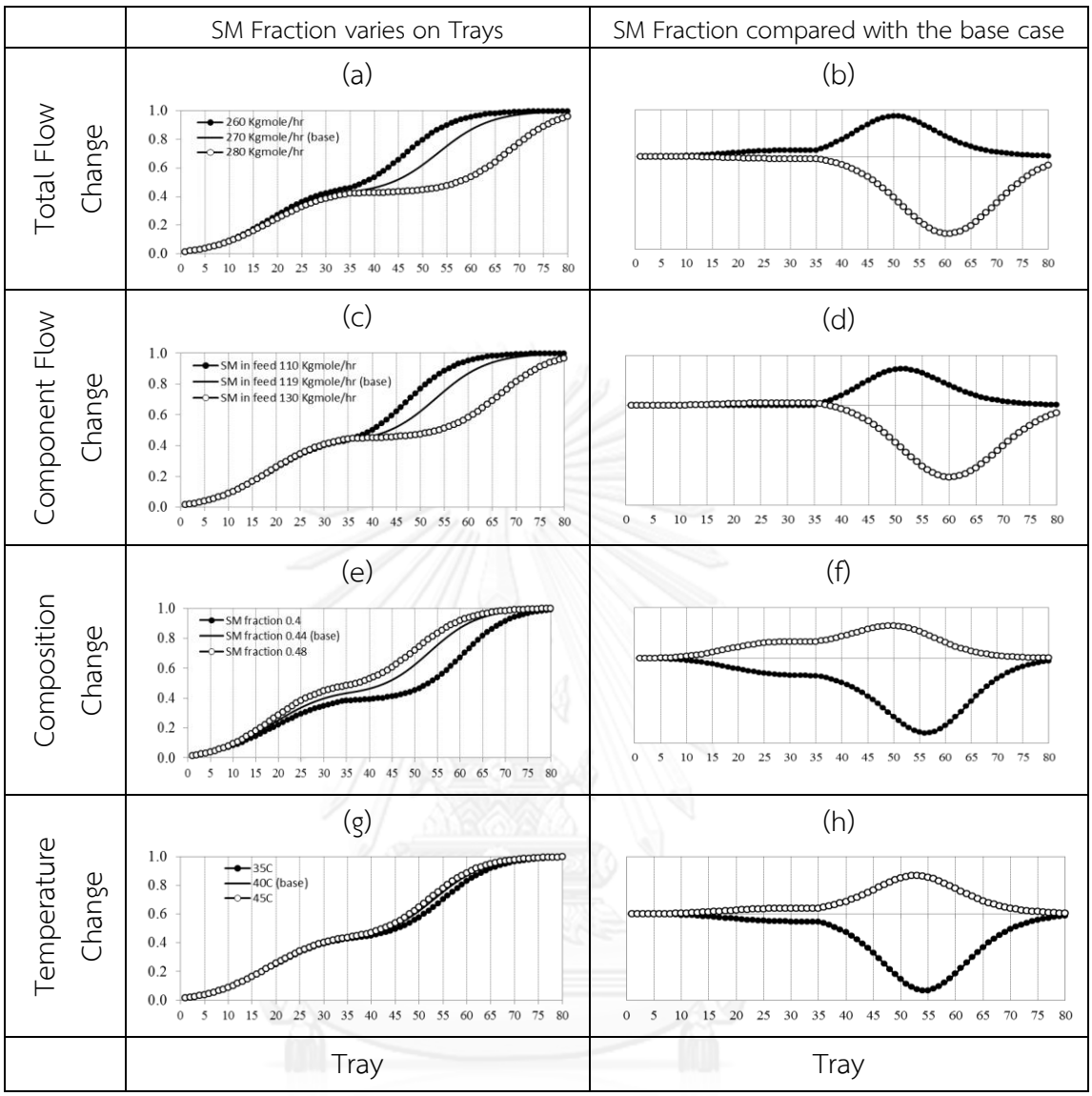


Figure 5.10 Column C1 disturbance test for sensitivity 1: fixed Qr and reflux flow

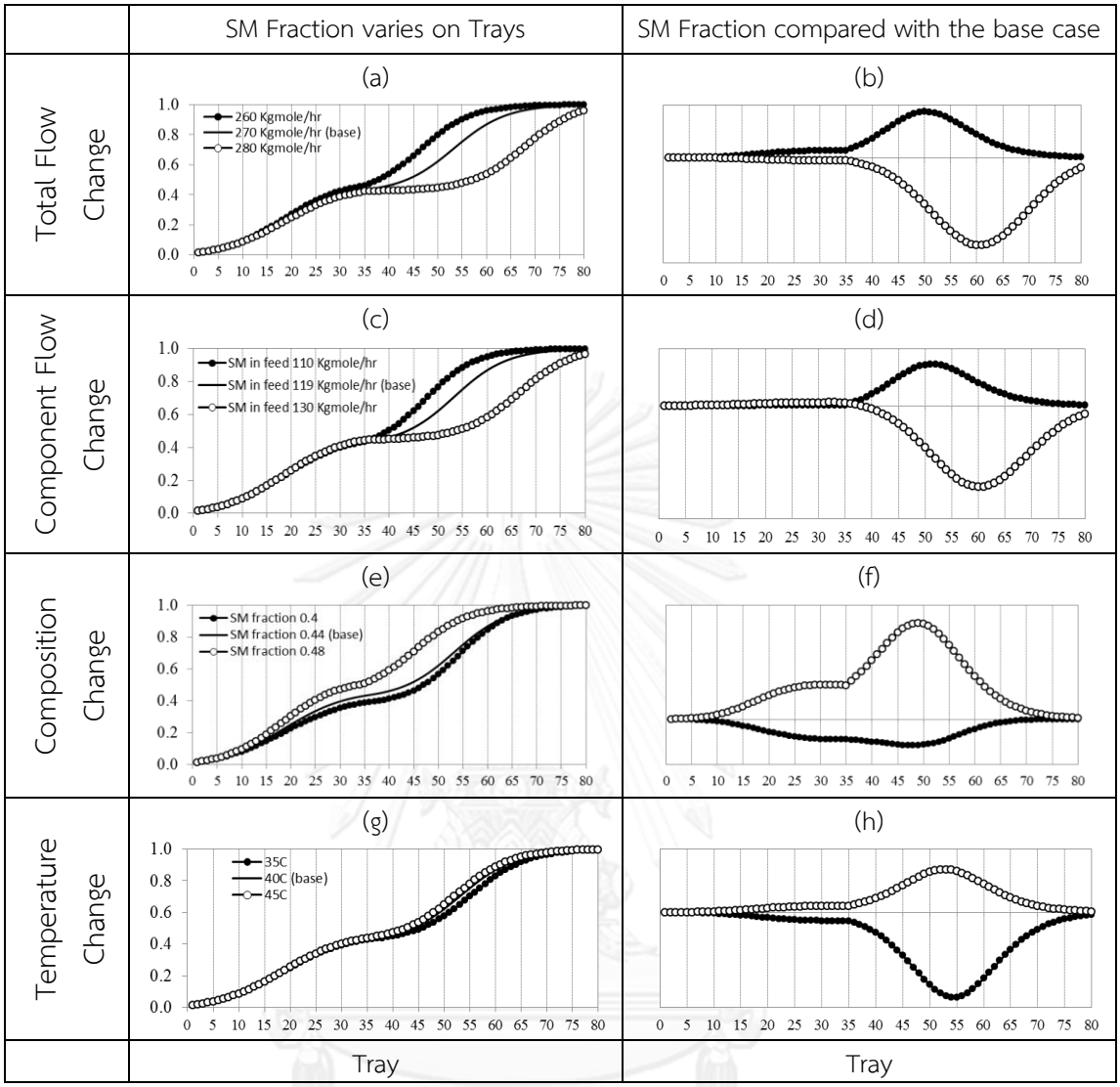


Figure 5.11 Column C1 disturbance test for sensitivity 2: fixed Qr and reflux ratio

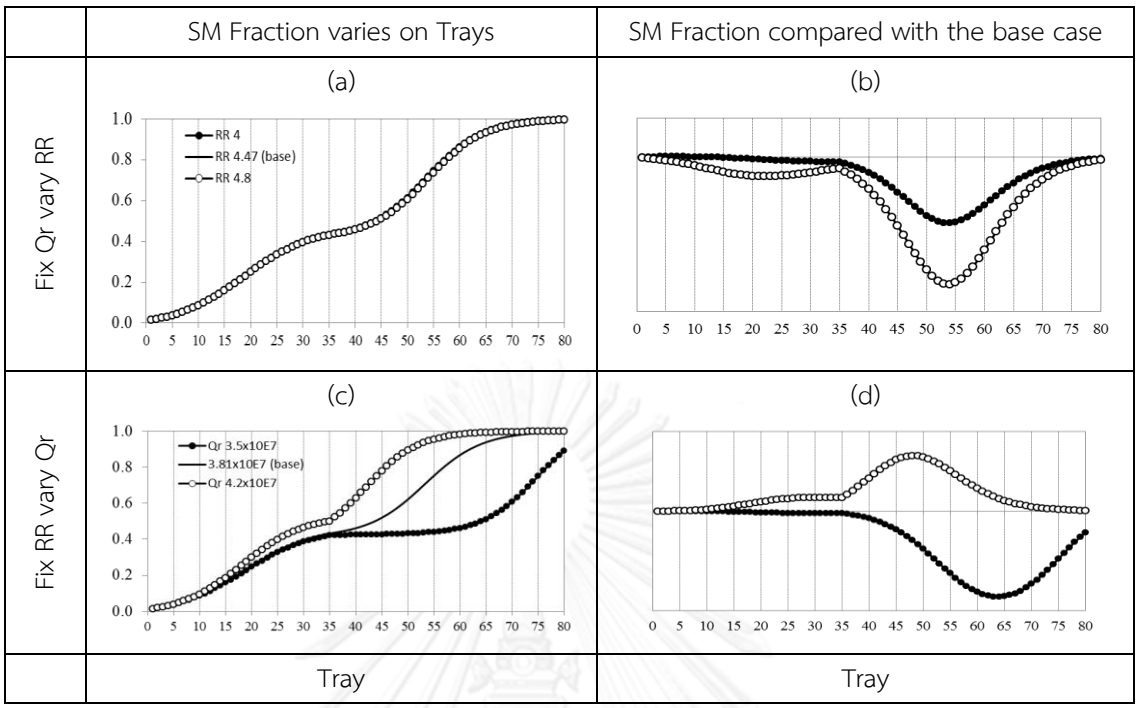


Figure 5.12 Column C1 disturbance test for sensitivity 3 and sensitivity 4

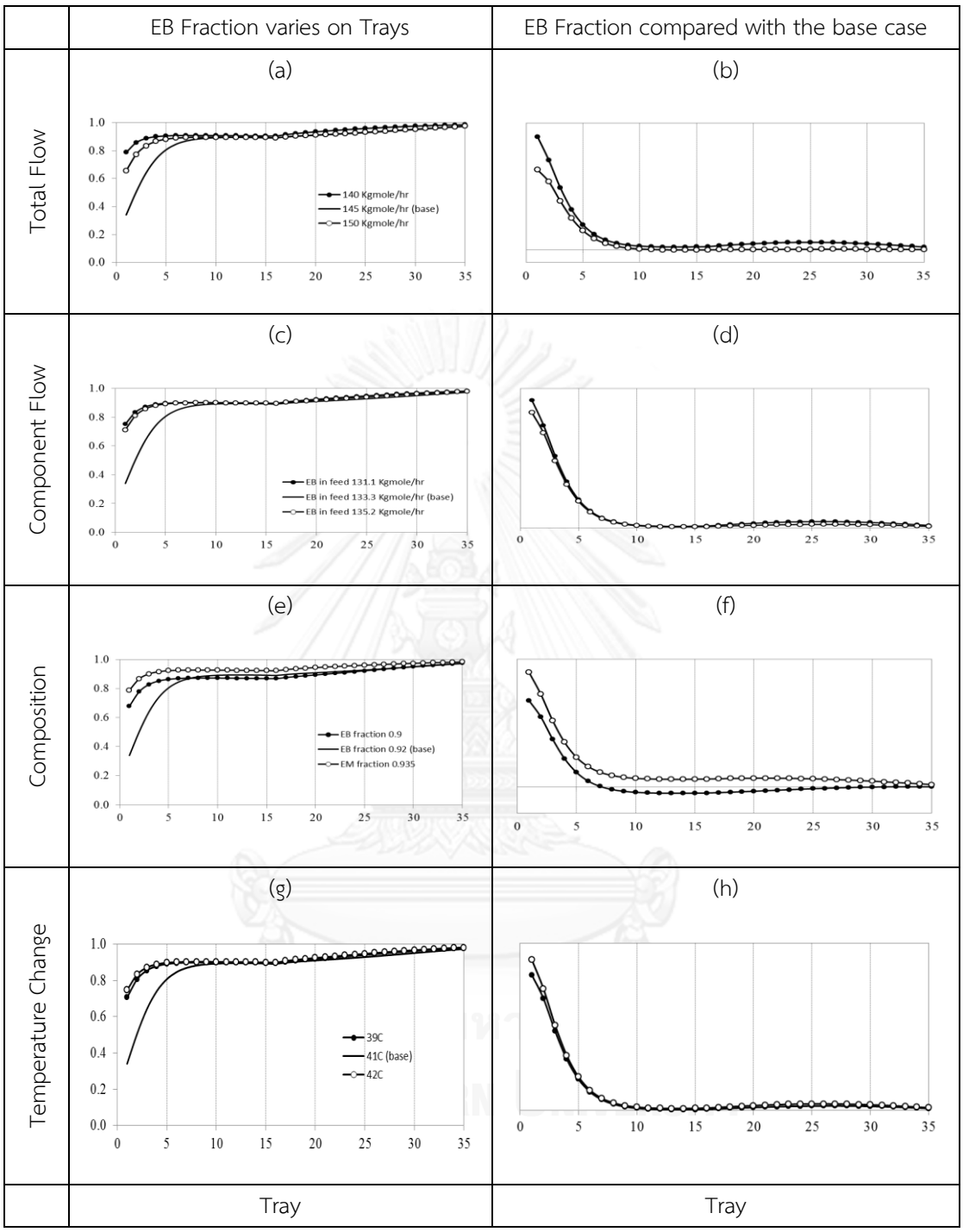


Figure 5.13: Column C2 disturbance test for sensitivity 1: fixed Qr and reflux flow

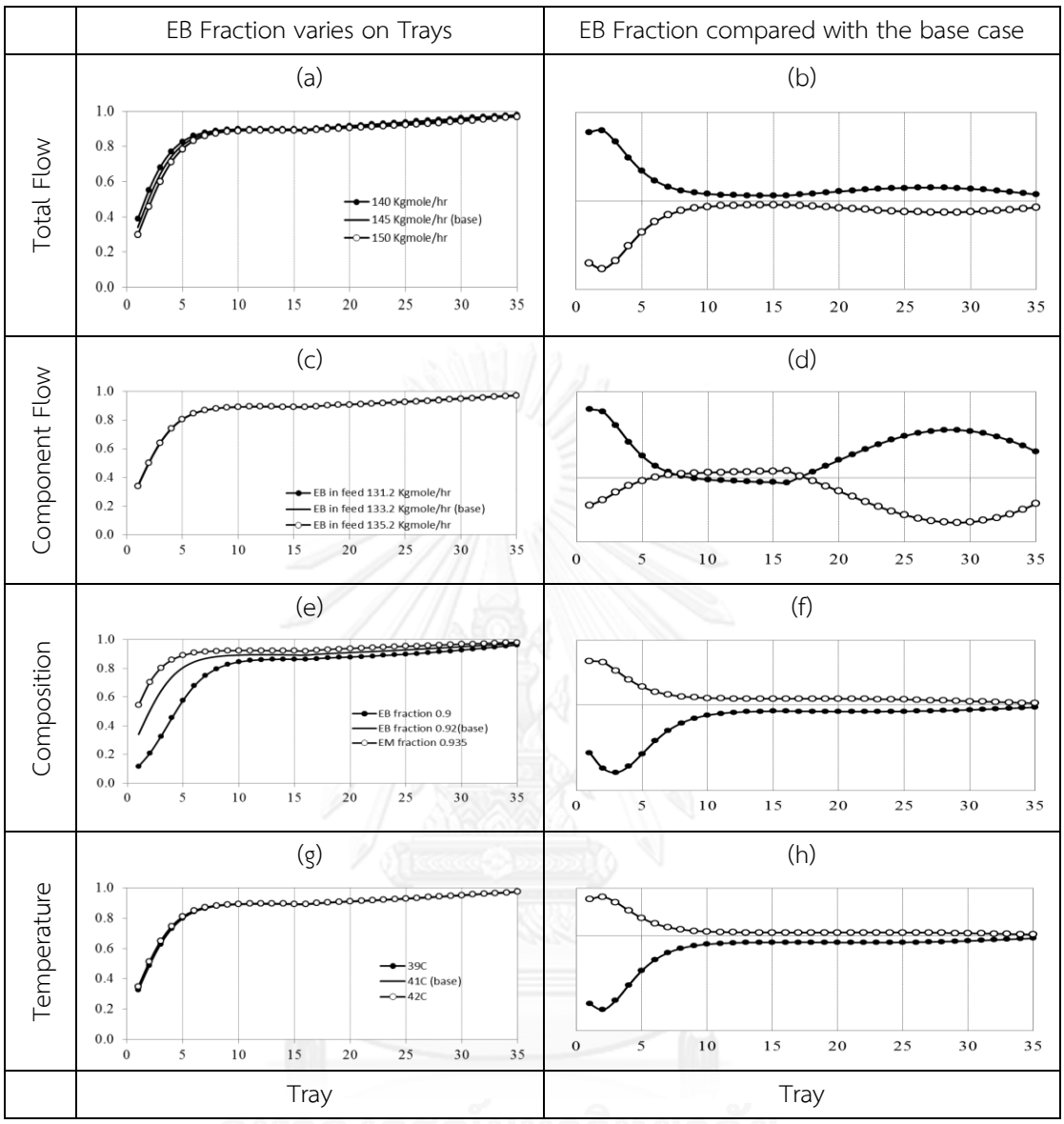


Figure 5.14 Column C2 disturbance test for sensitivity 2: fixed Qr and reflux ratio

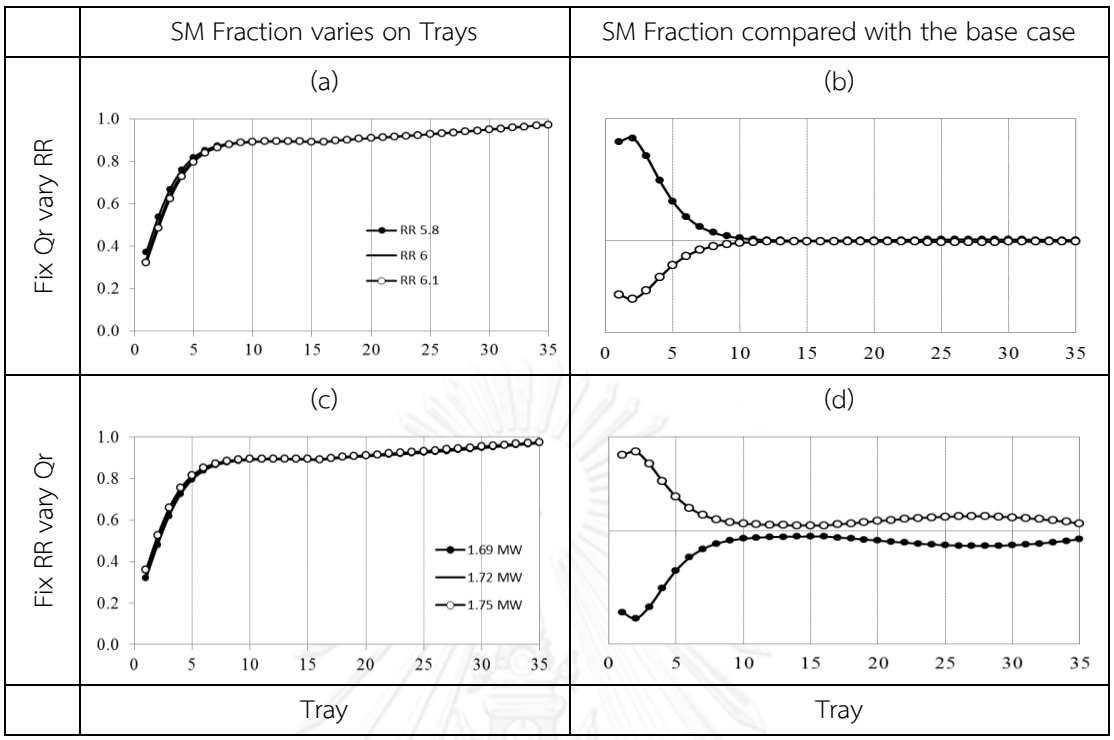


Figure 5.15 Column C2 disturbance test for sensitivity 3 and sensitivity 4

2.7 Production rate control

The production rate will vary on a total EB feed to the process and keep the ratio of EB/Steam constant for control partial pressure of EB in reactors.

Mode of operation. Mode of operation of the styrene process is on-supply while only one reactant is ethylbenzene.

Stage 2. Fixture plant and disturbance management, Plant level loop design

Step 3: Establish fixture plant. Establishing a material balanced plant by regulating all components at their quantifiers.

3.1 Keep the materials entered and/or reentered fixed.

There are two material streams fed into the process: fresh EB and low pressure steam. The recycle stream is EB. The composition of recycled EB is not much difference from fresh EB composition so we can regulated the total EB entered and reentered the process by adjusting fresh EB stream. Low pressure steam feed is flow control and its flow is ratioed to total EB flow. Low pressure steam is 14.9 times higher than total EB flow. Figure 5.16 show flowsheet with controllers for step 3.1

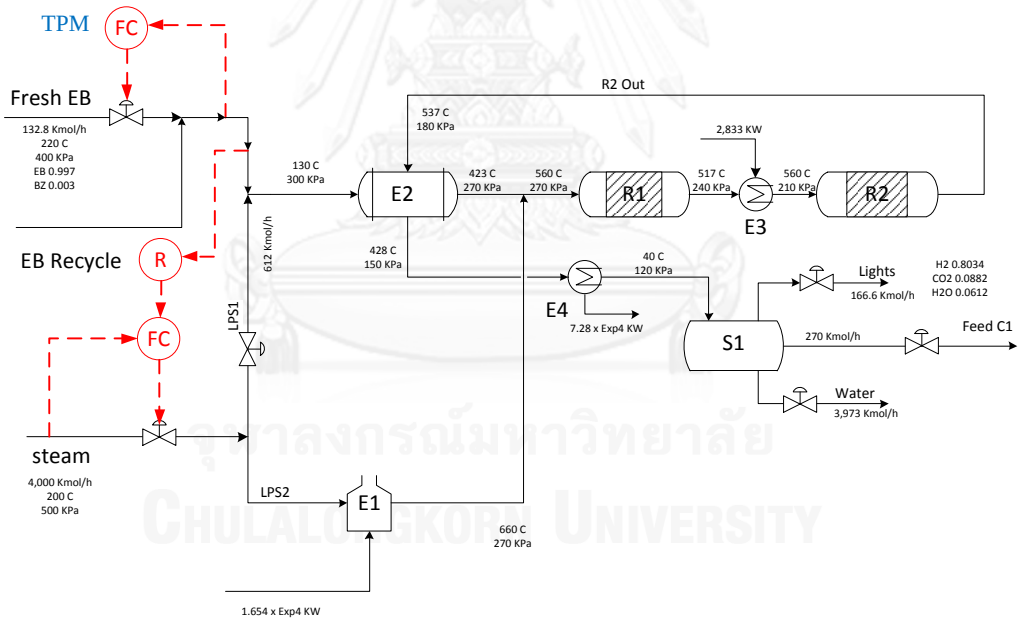


Figure 5.16 Flowsheet with controllers for step 3.1

3.2 Adjust the flow of exit material streams.

Four exit material streams are balanced by adjusting their accumulation. Hydrogen out is adjusted according to 3-phase separator pressure, its quantifier. Water out is regulated by maintaining its accumulation, i.e. 3-phase separator water level. SM product accumulated at C1 bottom, its quantifier, is regulated by adjusting C1 bottom flow. Benzene and toluene (byproduct) represented by C2 reflux drum level is regulated by adjusting C2 distillate flow.

3.3 Handling the material that not leaving the process.

There is no component not leaving the process.

3.4 Control the amount of the rest of the component at their quantifiers. All components are controlled at their quantifiers designed in Steps 3.1 and 3.2.

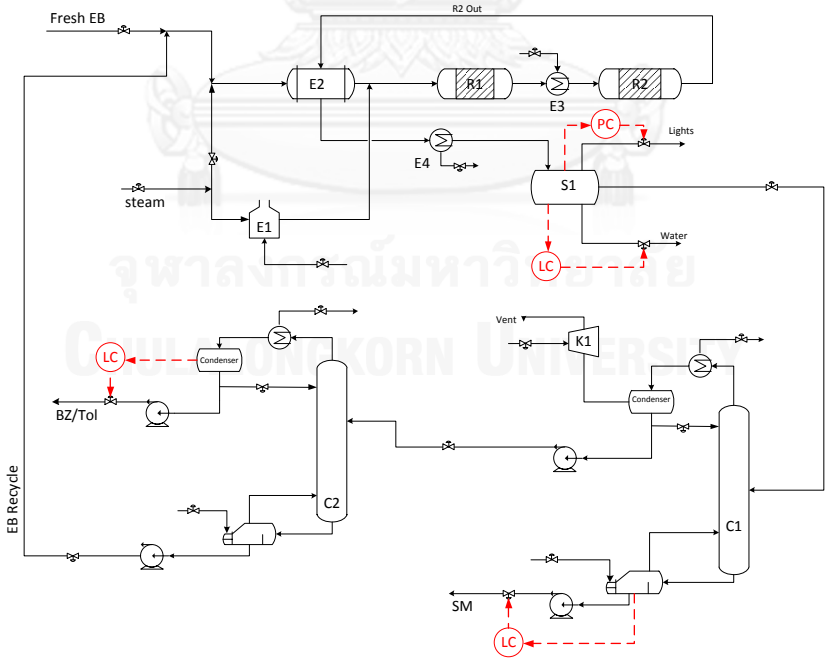


Figure 5.17 Flowsheet with controllers for step 3.2

3.5 Regulate the production rate.

The SM production rate is controlled by adjusting the temperature of reactor R2 feed according to styrene fraction in column C1 feed. The reactor R2 feed temperature control loop is needed.

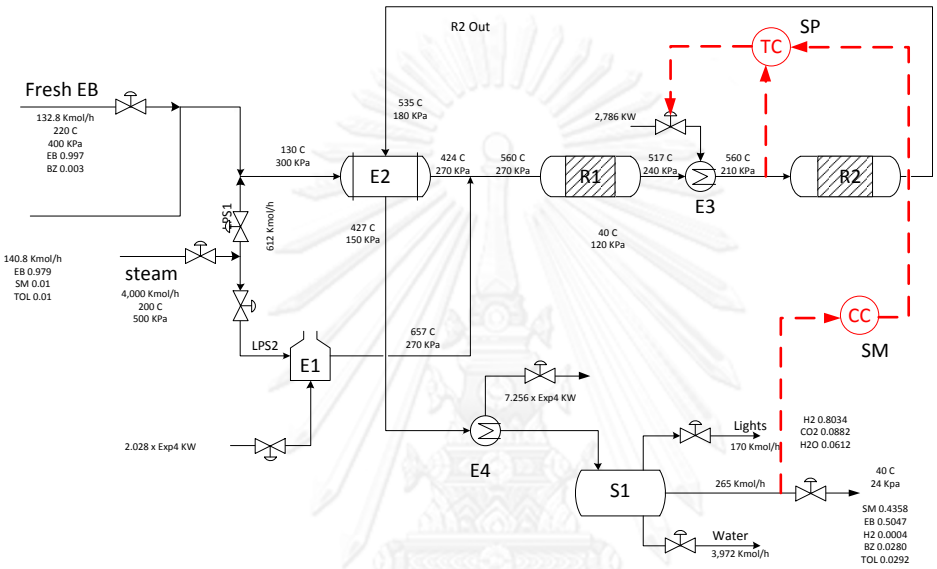


Figure 5.18 Flowsheet with controllers for step 3.5

Step 4: Disturbance management for quality control

4.1 Heat Disturbance Management

According to the analysis made in Step 2.2, the temperatures of the stream going into reactor R1, reactor R2 the stream entering the 3-phase separator must be maintained.

4.1.1 The furnace is to heat the steam to the desired reactor inlet temperature, 560°C. The reactor inlet temperature of reactor R1 is controlled by furnace duty. The heat disturbance in reactor inlet is rejected toward adjusting furnace duty. The heater E3 is to heat the reactor R2 inlet to 560°C. The disturbance in reactor R1 outlet is rejected through E3 duty. The cooler E4 is to cool 3-phase separator feed to 40°C. The disturbance in this feed is handle by E4 duty.

4.1.2 The condenser and reboiler duty of product column C1 and recycle column C2 are used to control product quality. All control loops of this step 4.1 shown in Figure 5.19.

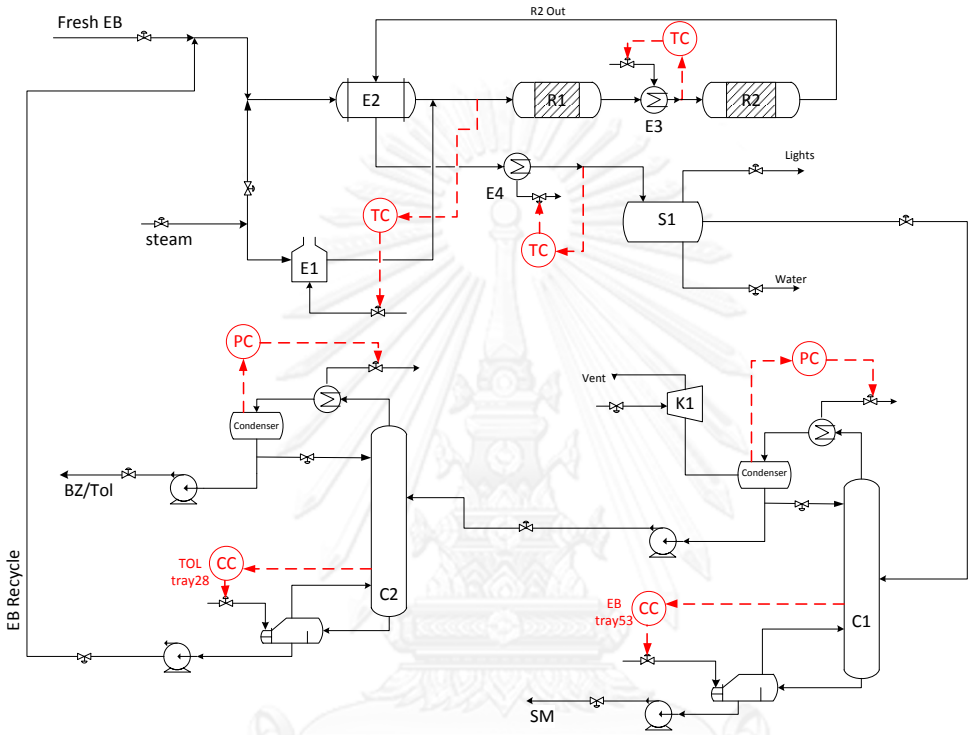


Figure 5.19 Control loops designed by step 4.1 heat disturbances management.

4.2 Material Disturbances Management.

Several disturbance tests are made to identify control structure to achieve the desired material disturbance shifting directions made in Section 2.6. Nine different control structures of distillation columns are proposed. Five structures for single end control are R, R/D, R/F, V/B and R/(R+D) and four structure of dual end control, detail of all structures are shown in Figure 5.20 and Figure 5.21. The controlled and manipulated variables of the purpose structure are shown in Table 5.4. In order to identify which structure gives the desired disturbance shifting, three disturbances: feed component (light and heavy keys) flow changes, and feed composition (light and heavy keys) changes are introduced to both columns. In

addition other two disturbances: feed temperature and feed flow changes are also tested to see which structure give the best performance on product purity. The product rates and reboiler duty are also considered.

Column C1. For SM in feed increased, all structures shift the surplus SM to the bottoms as desired. The transient responses are different. R/(R+D) and Dual3 structures are discarded in this transient response aspect while distillate flow in transient of V/B structure very fluctuate. For SM composition in feed change, see Figure 5.24, V/B and Dual3RR are better than the other in terms of bottoms composition, condenser and reboiler duties. For feed temperature and feed flow changes, R/F, R and Dual3 structures can be eliminated for dynamic response aspect. The rest control structures performance are acceptable. Dual3RR structures are selected for disturbance shifting and product purity criteria. Dual3RR structure use SM fraction tray 18 regulate reflux ratio and use EB fraction tray 53 to regulate reboiler duty.

For Column C2 the best structure of single end control is fix vapor to boil up/bottom flow ratio (V/B ratio) as it moves smoothly to steady state. The best structure of dual end control are Dubal2 but dual control give more fluctuated in a transient state. To improve the control performance new structure Dual V/B was simulated by using toluene fraction on tray 28 to adjust V/B ratio and for rectifying section used EB fraction on tray 1 to adjust reflux flow. Performance of Dual V/B better than others in term of smooth transient.

Table 5.4 Nine control structures of two distillation columns for dynamic testing

	Column C1		Column C2	
	Rectifying Section	Stripping Section	Rectifying Section	Stripping Section
Single End Control				
Fix R	Fix R	EB tray 53	Fix R	TOL tray 28
Fix RR	Fix RR	EB tray 53	Fix RR	TOL tray 28
Fix R/F	Fix R/F	EB tray 53	Fix R/F	TOL tray 28
Fix R/(R+D)	Fix R/(R+D)	EB tray 53	Fix R/(R+D)	TOL tray 28
Fix V/B	SM tray 18	Fix V/B	EB in TOP	Fix V/B
Dual End Control				
Dual1	SM tray 18, reflux	EB tray 53, bottoms	EB in TOP, reflux	TOL tray 28, bottoms
Dual2	SM tray 18, distillate	EB tray 53, Qr	EB in TOP, distillate	TOL tray 28, Qr
Dual3	SM tray 18, reflux	EB tray 53, Qr	EB in TOP, reflux	TOL tray 28, Qr
Dual4	SM tray 18, distillate	EB tray 53, bottoms	EB in TOP, distillate	TOL tray 28, bottoms
Dual3RR	SM tray 18, RR	EB tray 53, Qr		
Dual3VB			EB in TOP, reflux	TOL tray 28, V/B

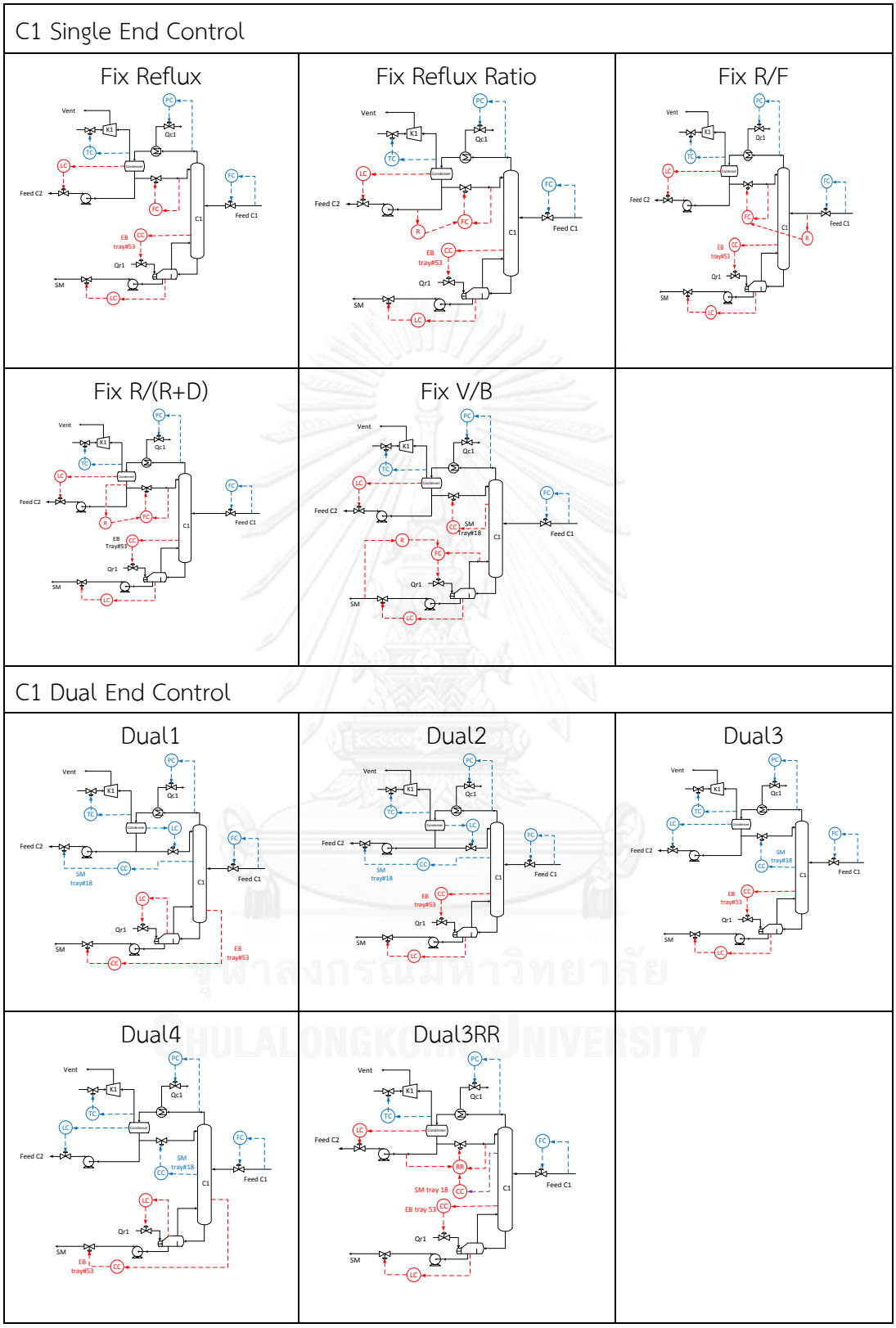


Figure 5.20 Difference control structures of distillation column C1.

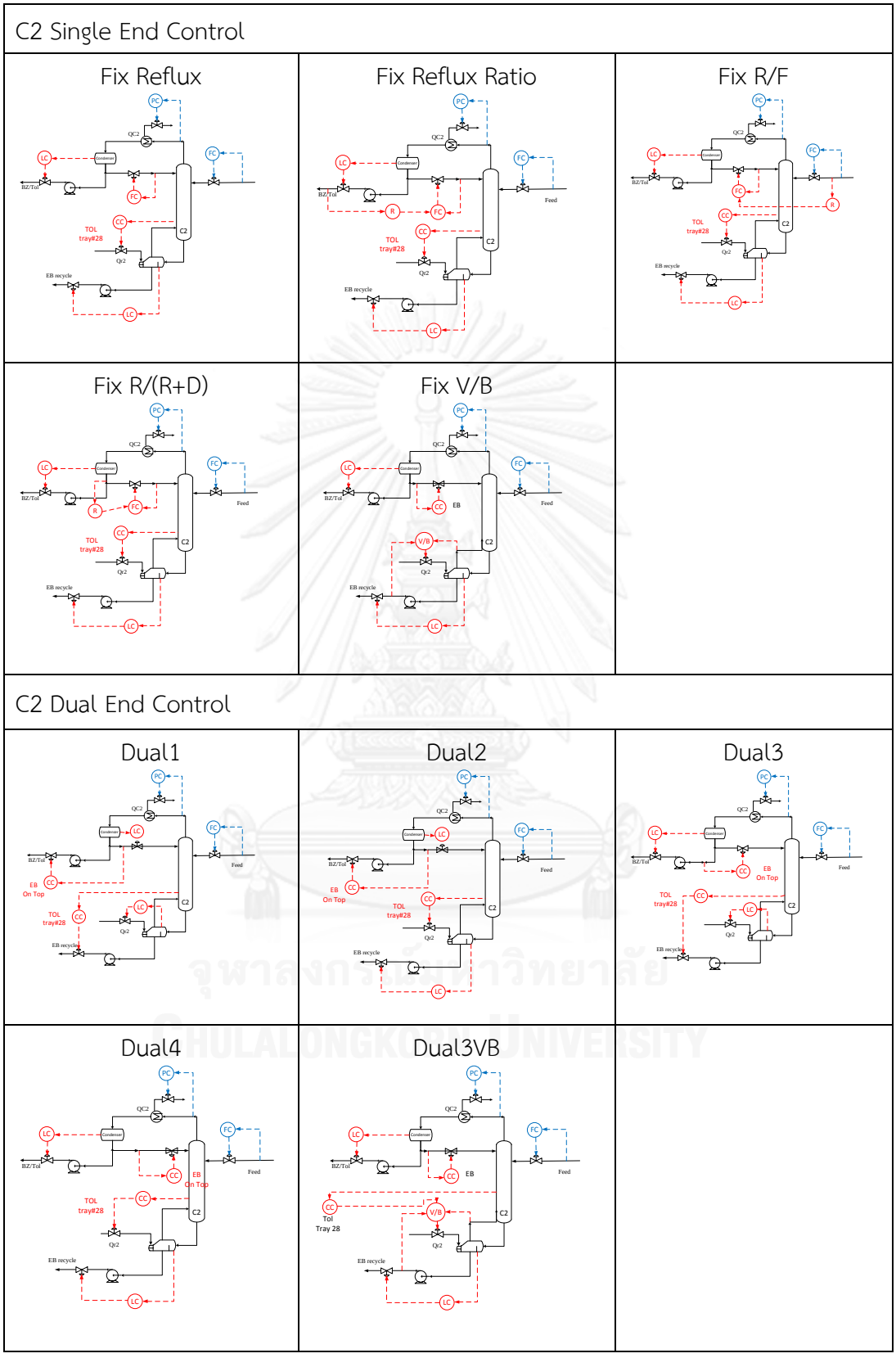


Figure 5.21 Difference control structures of distillation column C2.

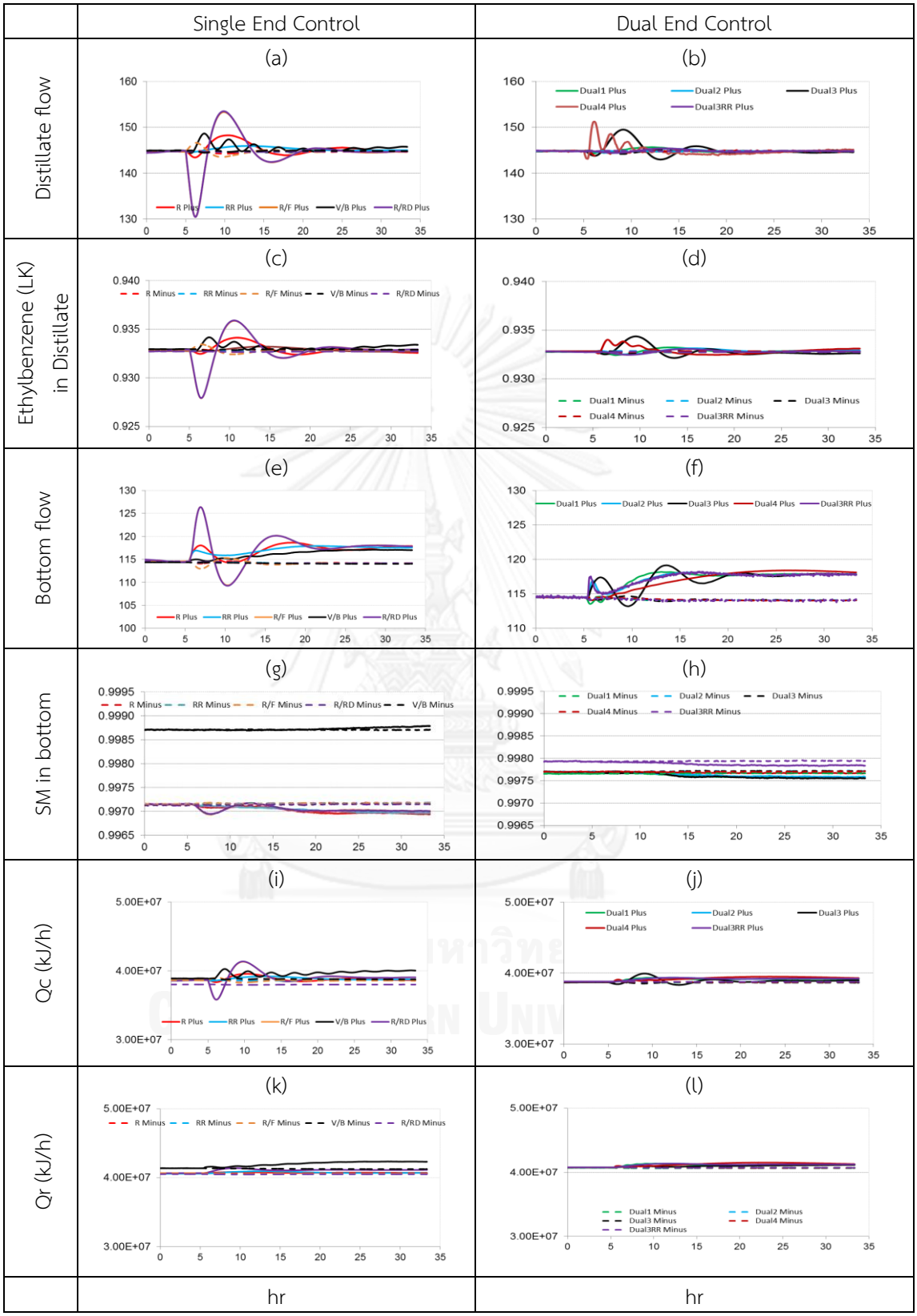


Figure 5.22 Column C1 dynamic results for component flow changes, Plus SM 118 and Minus SM 114.3 kmol/h

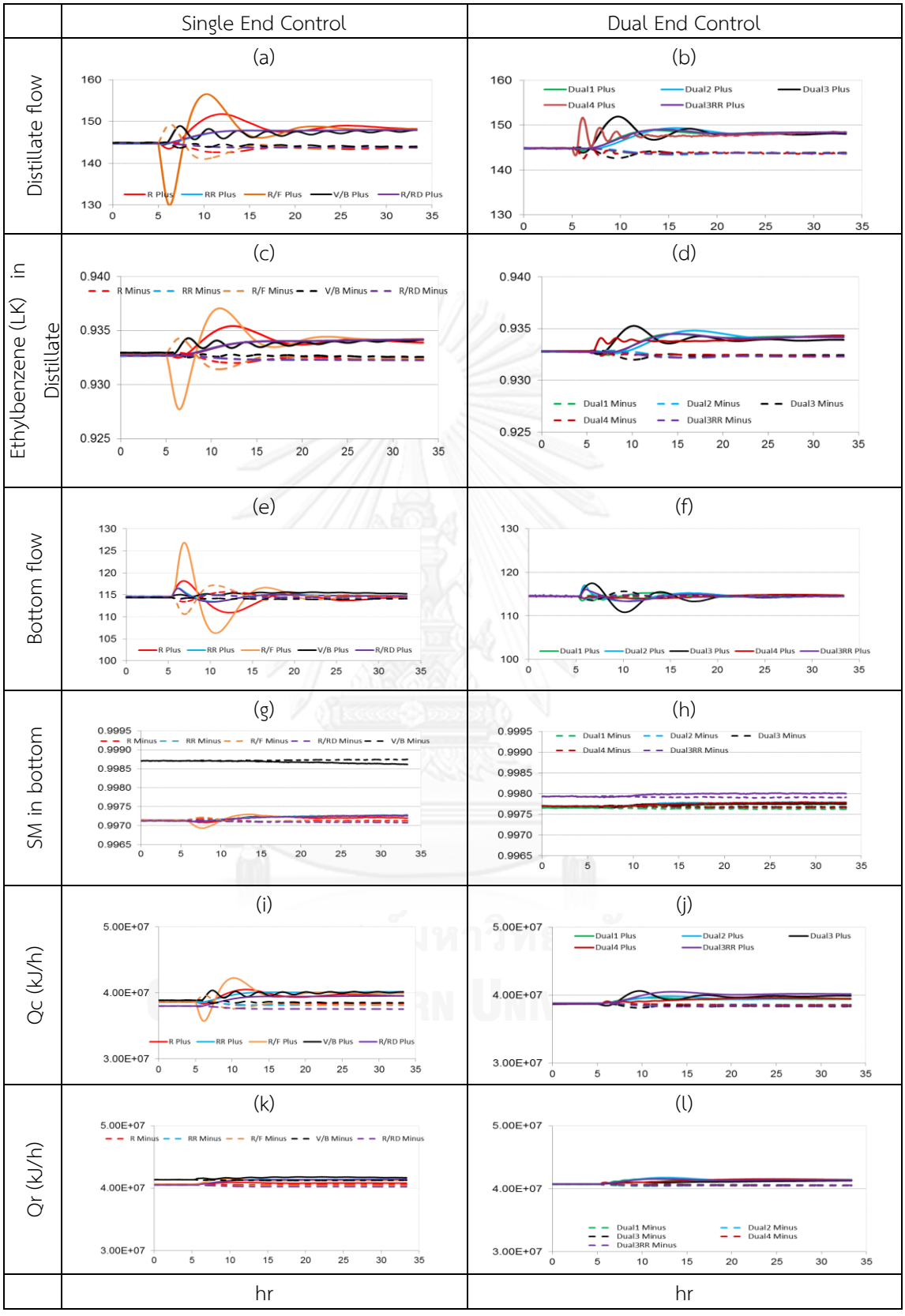


Figure 5.23 Column C1 dynamic results for component flow changes, Plus EB 139 kmol/h, Minus EB 134.5 kmol/h and base case EB 135.5 kmol/h

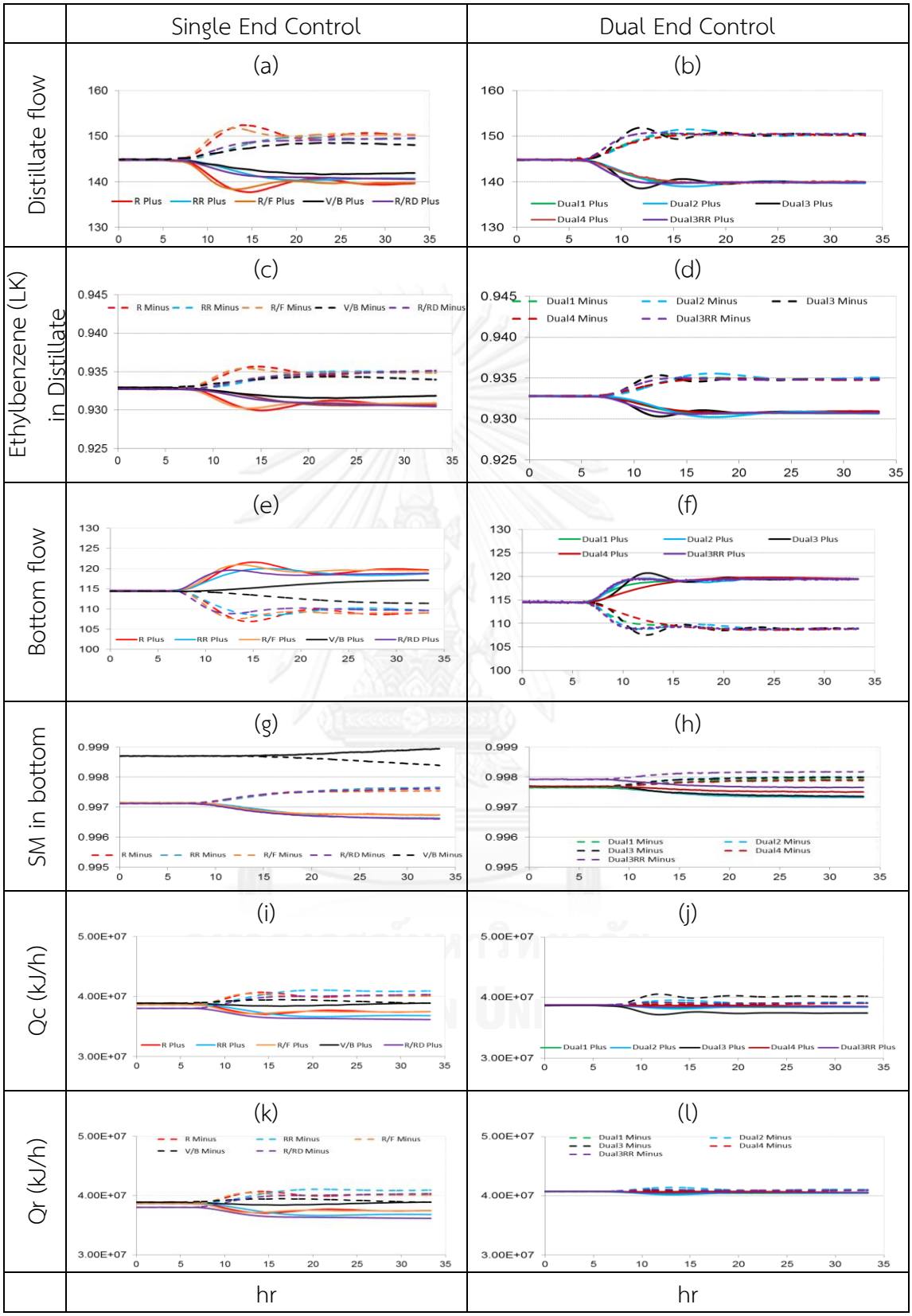


Figure 5.24 Column C1 dynamic results for composition changes, Plus SM 0.46 and Minus SM 0.40

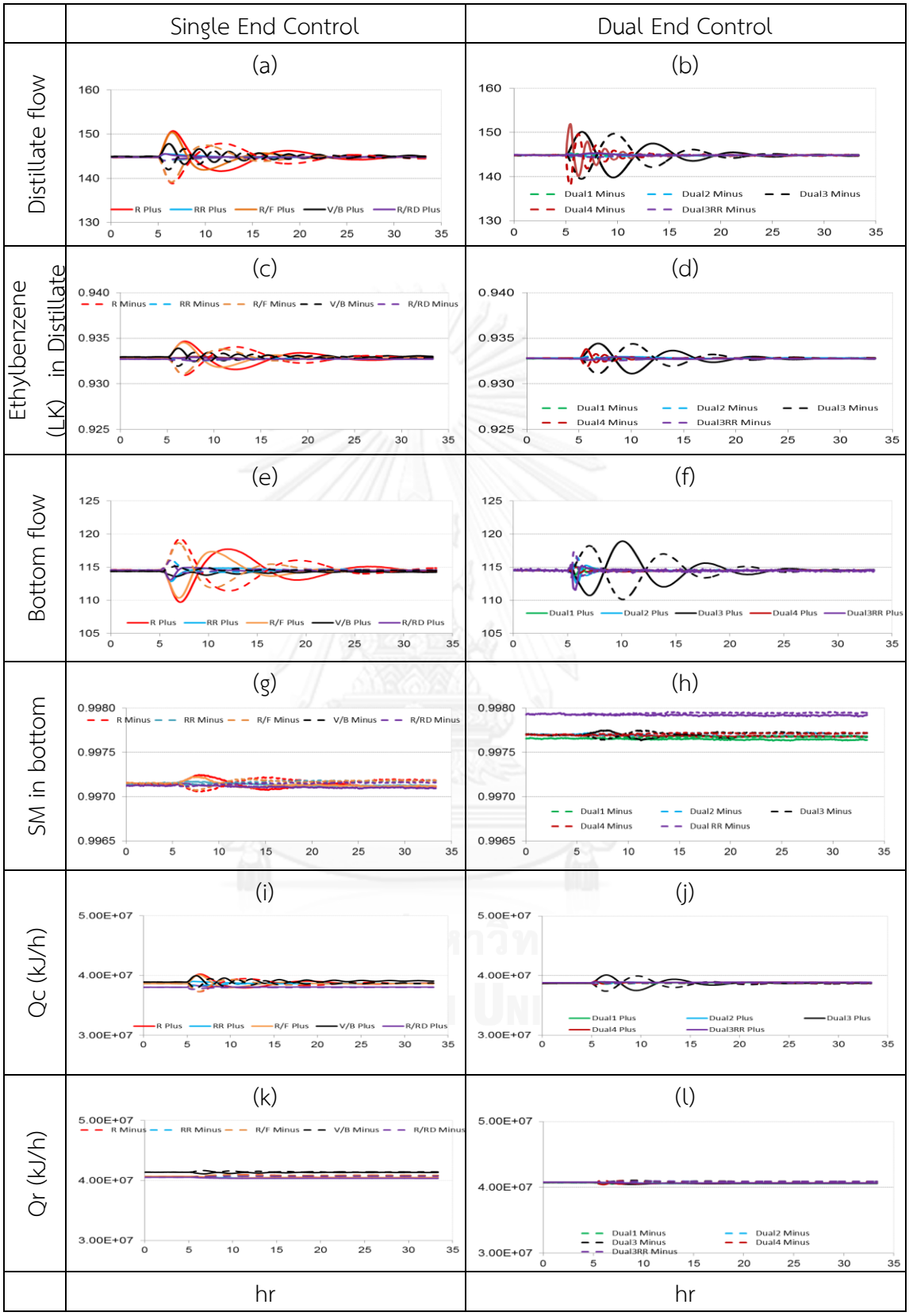


Figure 5.25 Column C1 dynamic results for temperature changes, Plus 44°C and Minus 36°C

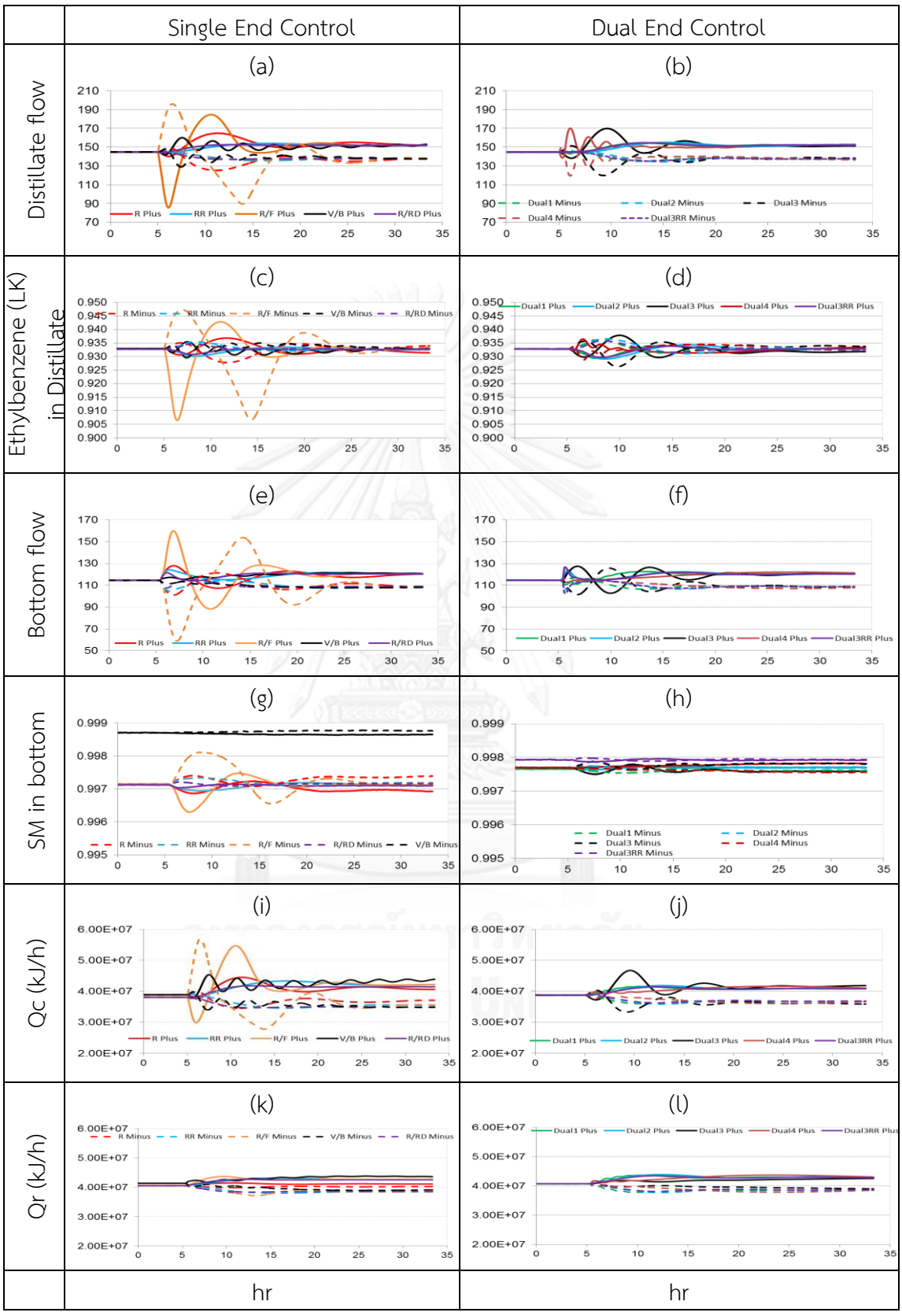


Figure 5.26 Column C1 dynamic results for total EB flow changes, Plus 273 kmol/h and Minus 247 kmol/h

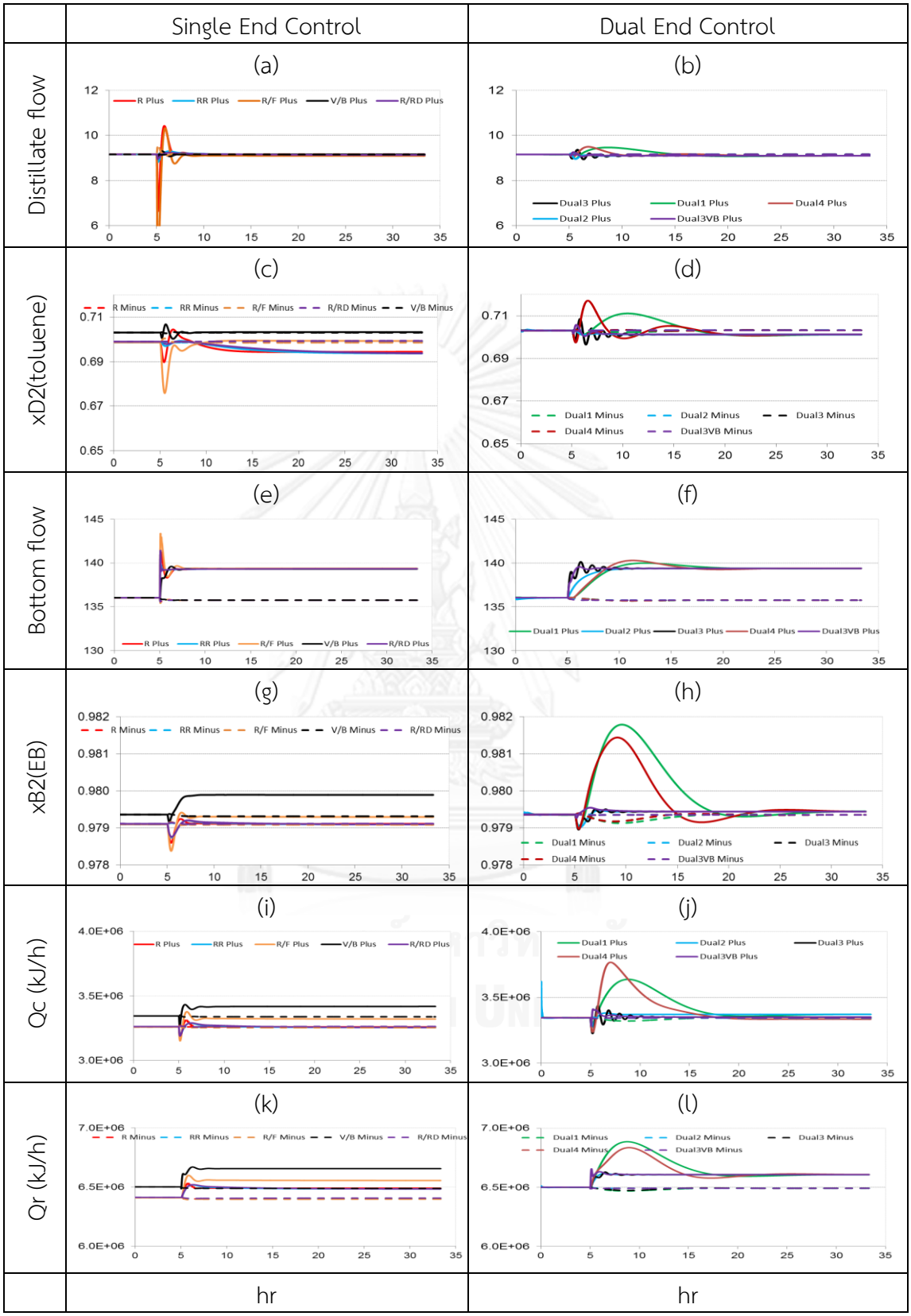


Figure 5.27 Column C2 dynamic results for component flow change, Plus EB 136.56 kmol/h and Minus EB 133 kmol/h.

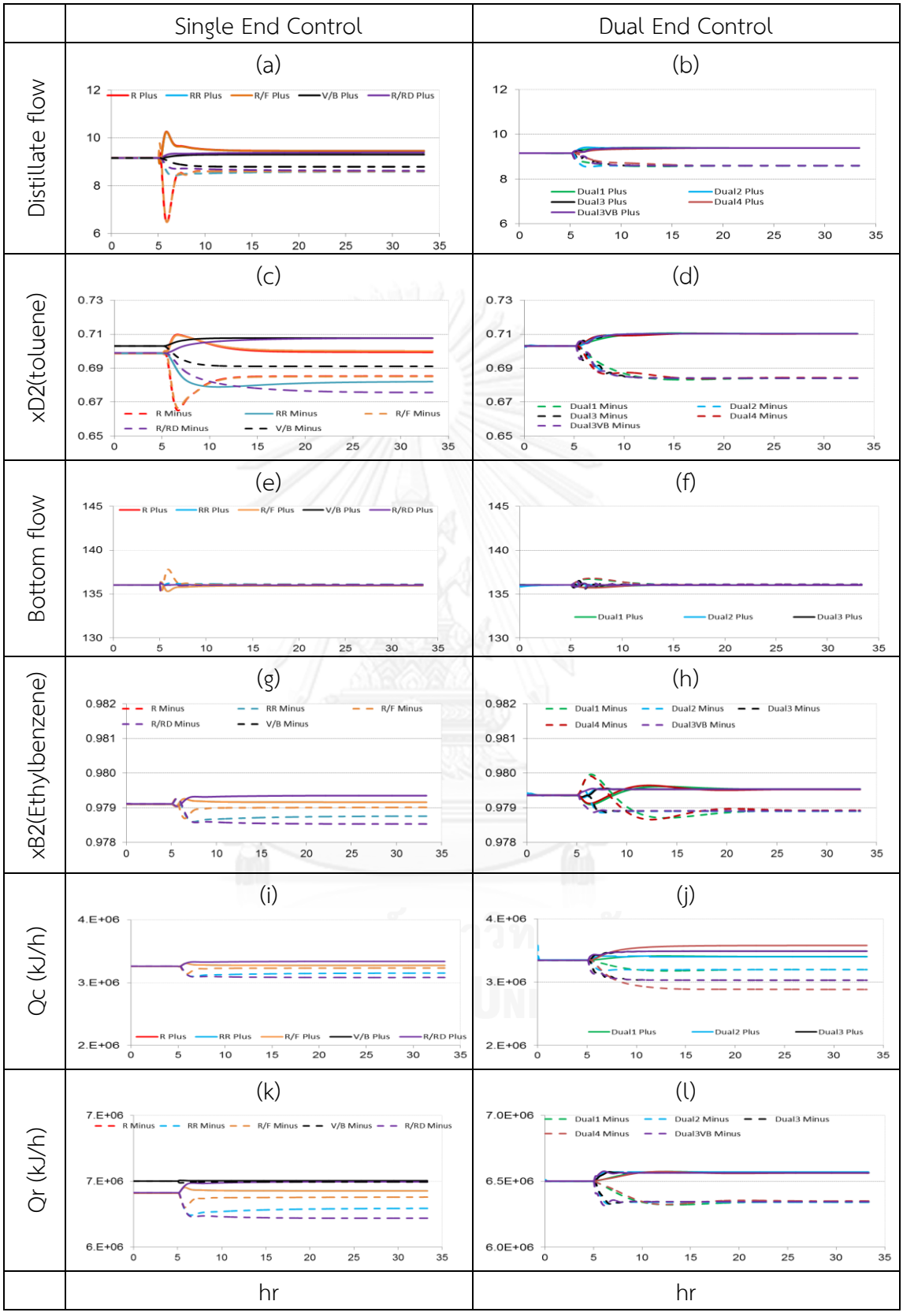


Figure 5.28 Column C2 dynamic results for component flow change, Plus toluene 8 kmol/r and Minus toluene 7.3 kmol/h

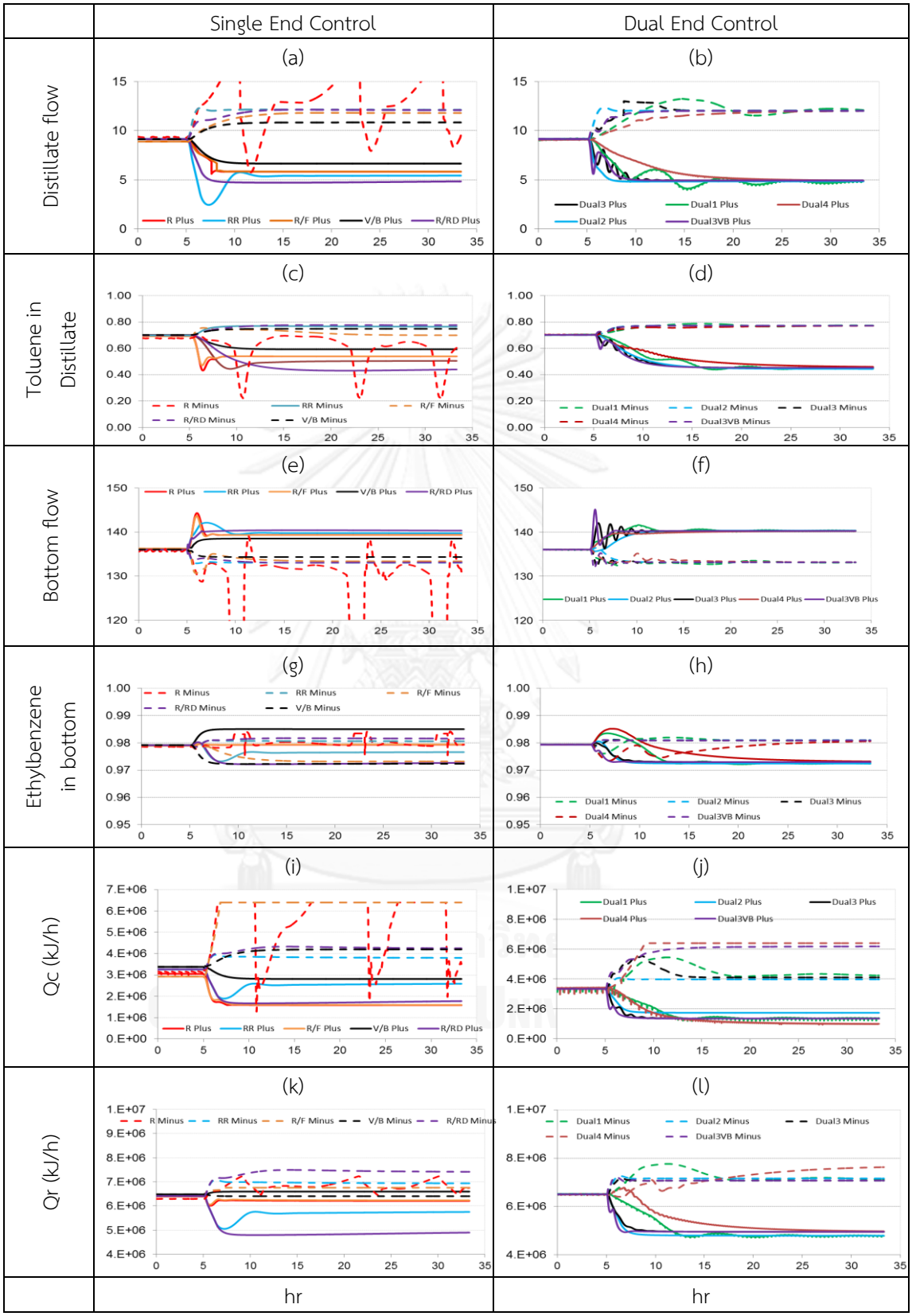


Figure 5.29 Column C2 dynamic results for composition change, Plus EB 0.94 and Minus EB 0.9

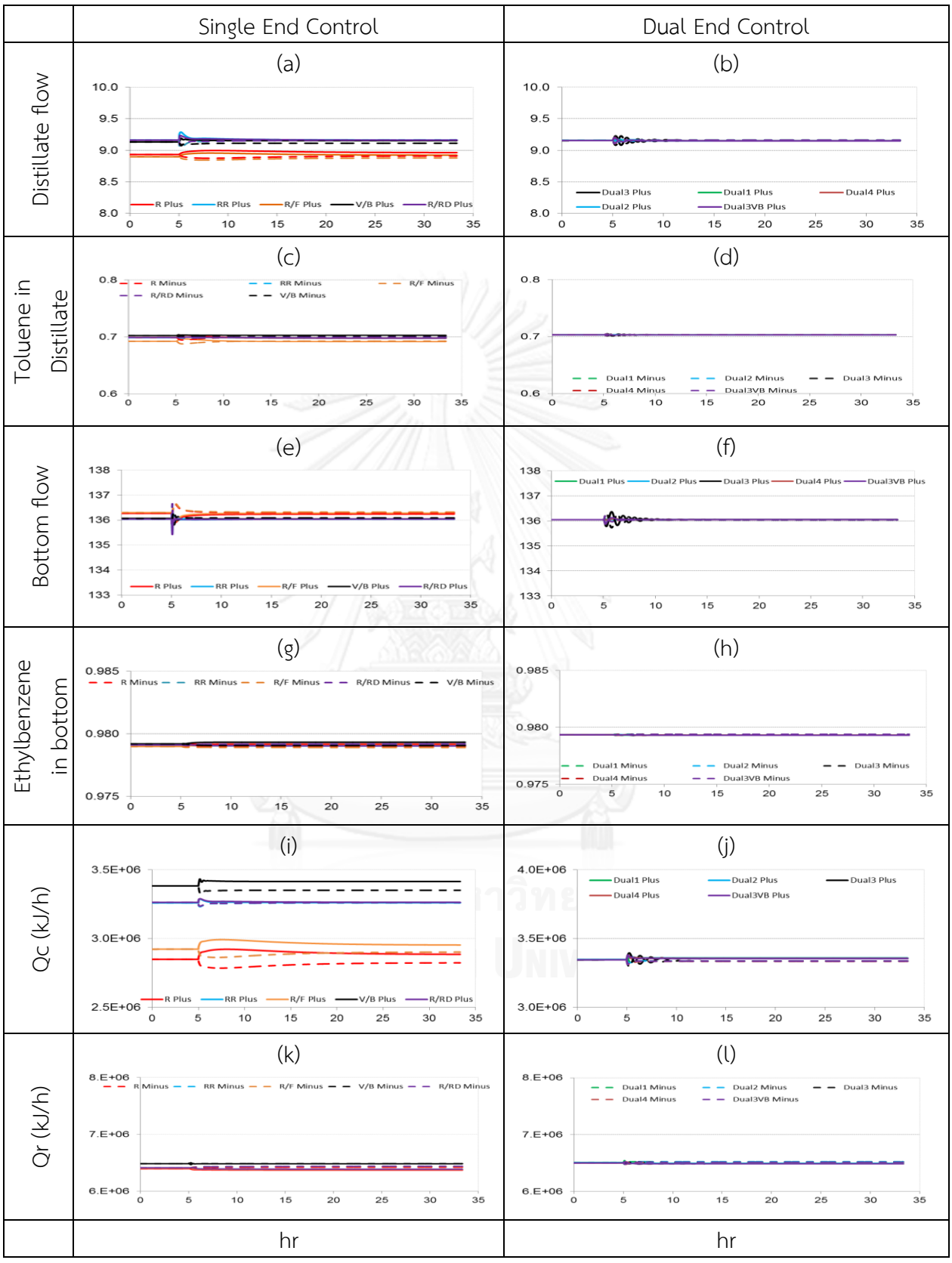


Figure 5.30 Column C2 dynamic results for temperature change, Plus 42°C and Minus 40°C

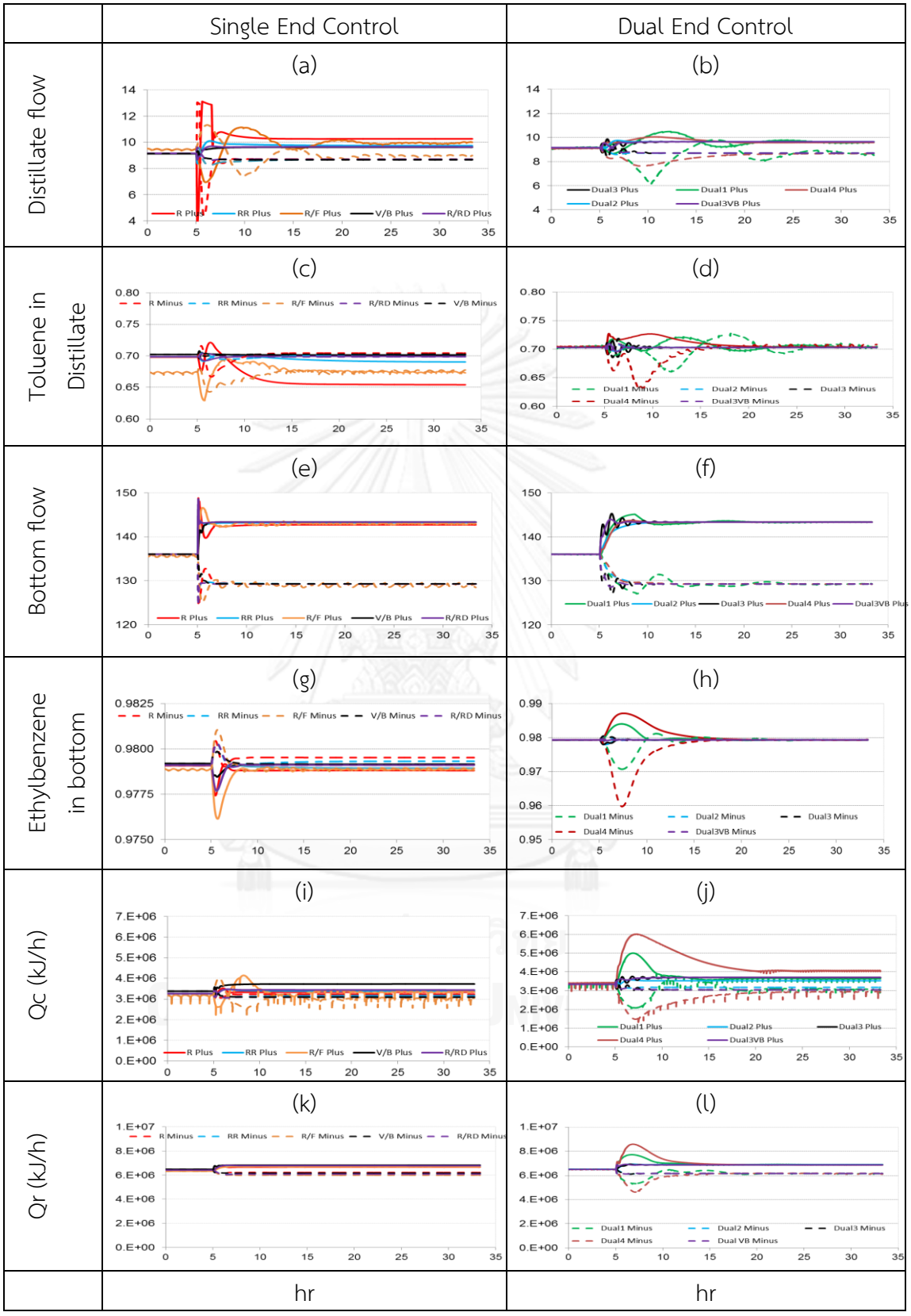


Figure 5.31 Column C2 dynamic results for total flow change, Plus 153 kmol/h and Minus 138 kmol/h

Stage 3. Unit Level Loop and Enhanced Loop Designs

Step 5: Design the rest of the control loops.

The units to be considered in this step are compressor K1 and organic level of 3-phase separator. The temperature of a stream going into the compressor K1 are controlled by manipulating the compressor duty, when condenser temperature increase meaning is heavy components leak to the top and the loss of high value components, EB in the vent stream and to prevent such situation the controller will decrease compressor duty to decrease vent flow then pressure in condenser build up and then cooler duty will increase to down pressure. Organics level of 3-phase separator is adjusting by product column C1 feed flow. For enhanced control, ratio control will be used for control total EB/LPS1 ratio. For column C1 main product styrene with purity 99.7% is the process objective, to keep on the objective a cascade loop is added to regulate EB tray 53 set point by SM product purity at bottom tray. Figure 5.32 show control loop designed by this step.

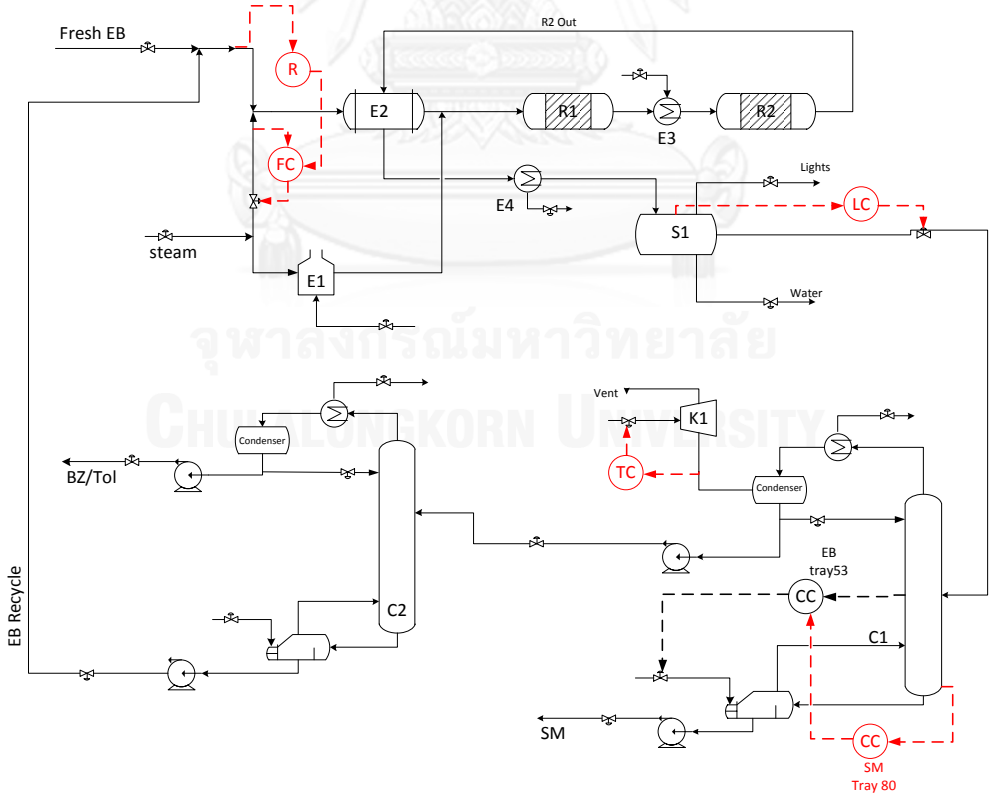


Figure 5.32 Control loop design by step 5

Stage 4. Energy Management and Optimization

There is no supplementary design in this step.

Stage 5. Design Validation

Step 8: Validate the designed control structures by rigorous dynamic simulation via HYSYS process simulation software. The performance of the design control structures is tested by difference disturbances, changes of total EB feed flow rate (TPM), EB fraction in fresh feed decrease and decrease 5% of catalyst activity. The control structure designed by Skogestad (2004) and Vasudevan (2011) are also tested in comparison. There is no temperature disturbance test due to all structure used same loops to reject heat disturbance. Figure 5.33 show new control structure which design by Wongsri's Procedure. All structure used dual end control structure for distillation column but difference detail as show in Figure 5.34 and Figure 5.35.

Table 5.5 Summary of disturbances

Scenario	Description
1	Total EB change from 268 kmol/h to 281.5 kmol/h
2	Total EB change from 268 kmol/h to 254.5 kmol/h
3	EB fraction in fresh feed decrease from 0.997 to 0.977
4	Decrease catalyst deactivate by 5%

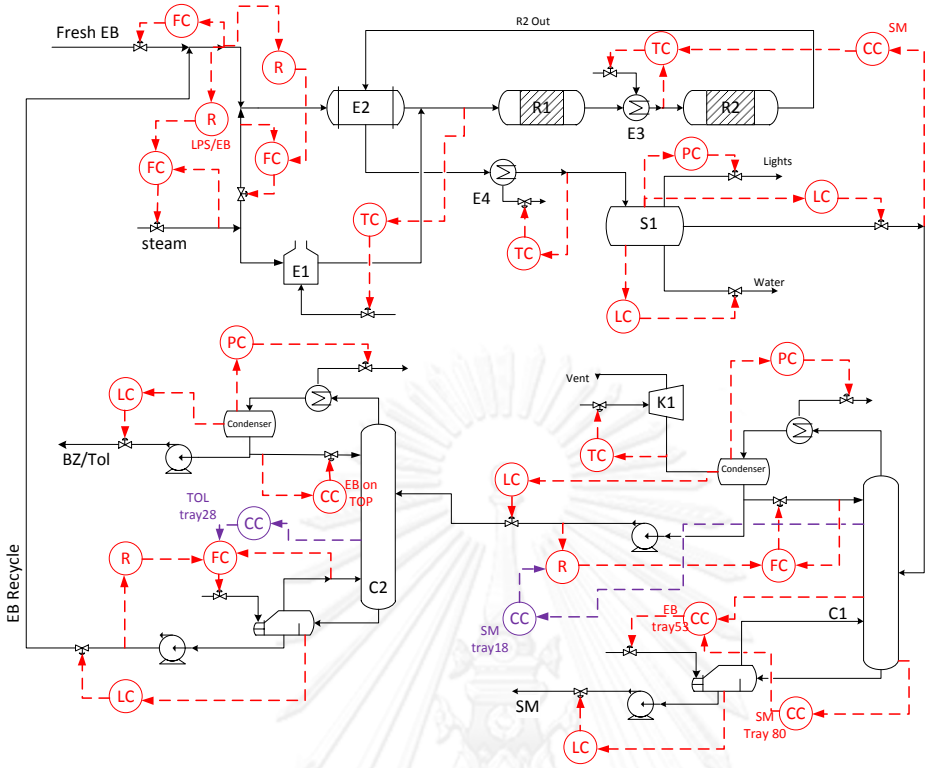


Figure 5.33 New Design Structure by Wongsri procedure

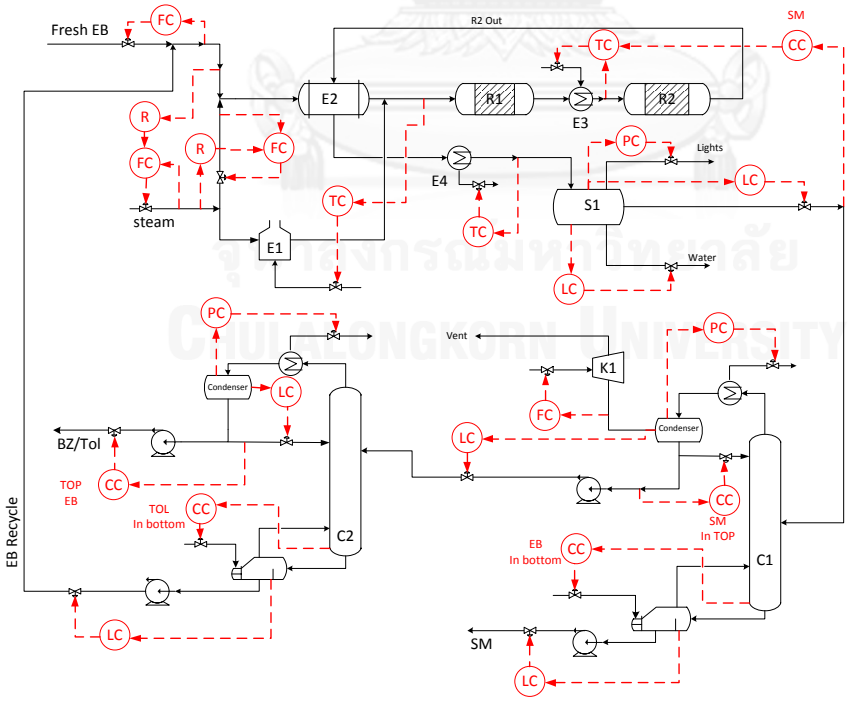


Figure 5.34 Control structure design by IFSHO procedure

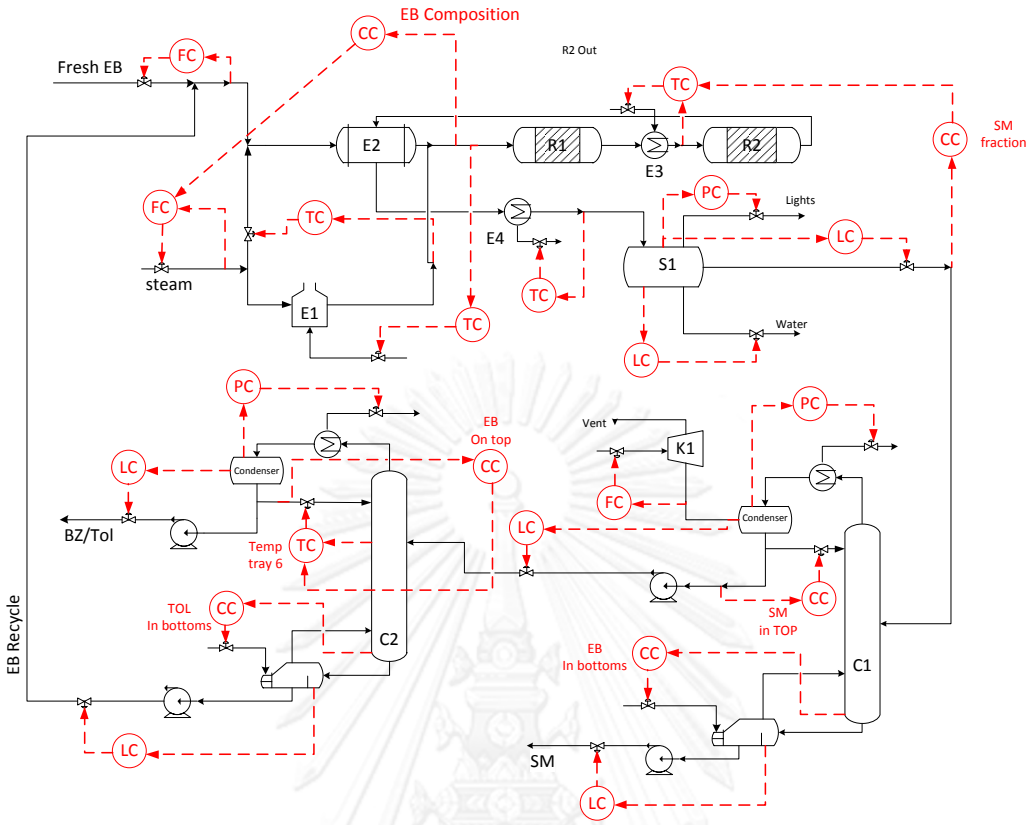


Figure 5.35 Control structure design by SOC procedure

5.2 Dynamic simulation results

The new design control structure consists of 24 control loops, involving 20 CDOFs. The details of loops and tuning parameter are shown in Table B1. The composition, temperature and ratio control loops are tuned as PI controllers using the autotuner in Aspen HYSYS. All PI controllers also use autotuner. Dead time 3 minutes are inserted into the composition loops and 1 minute for temperature loop to account for the dynamic delay of real process. Figure 5.36 gives results for ±5% changes in the total EB feed flow. The solid lines are 5% increase; the dashed lines are 5% decrease. The first column is the result of New Design Structure compare with result of IFSHO structures and SOC structure. Figure 5.37 gives the results for scenario 3, EB purity in fresh feed decrease from 0.997 to 0.977 and scenario 4, deactivation of catalyst. All disturbances start at 5 hours.

For the change in total EB, fresh EB and fresh steam has also changed immediately from the effected of ratio control for New Design Structure and IFSHO but for SOC structure found slightly slower effected by cascade loop. The cascade loop manipulates the steam flow to keep the EB fraction in feed reactor R1 at 0.06. All control structures able to keep 0.44 of SM fractions in feed C1 by adjusting reactor R2 feed temperature and able to maintain the purity of SM product purity 99.7% as specified show in Figure 5.36C. The dynamic response of $X_{D2\ (EB)}$ of the IFSHO structure is more oscillatory than other control structures.

For the change in EB composition in fresh feed by decreased from 0.997 to 0.977, fresh EB feed increased for all structures. New design structure and IFSHO structure keep fresh steam feed constant, but SOC structure decrease fresh steam flow to regulate EB fraction in feed of reactor R1 and decreases LPS1 flow for regulate furnace outlet temperature, shown in Figure 5.37A. All structure increase reactor R2 feed temperature and keep SM fraction constant in feed column C1. At column C2 of IFSHO structure, regulate EB fraction on TOP by distillate flow but it cannot keep EB fraction on TOP as specification, show in Figure 5.37D.

For the disturbance of catalyst deactivation, all structures increased of reactor R2 feed temperature and keep SM fraction in feed of column C1 at 0.44 but more

transient time to reach SM fraction control target show in Figure 5.37B. In Figure 5.37A show that LPS1 flow rate of SOC structure, LPS1 increase to keep the furnace outlet temperature at set point.

To regulate vent flow New design used compressor feed temperature when the feed temperature increases, it represents that some heavy components leak to the top, to prevent material loss, the controller will decrease vent flow by decrease duty, show in Figure 5.38A.

For Figure 5.38B show action of controller to regulate EB on top of column C2, New design used reflux flow to regulate the composition which yields good performance, but IFSHO used distillate flow to regulate the composition and yield low performance because the flow rate of distillate very small when compared with feed.

For the other criteria, IAE of major loops are calculated and found that New design structure has lower IAE than the other for all disturbances show in Table 5.6. For the overall transient of the process (DDS) New design has better performance than others, but in term of margin and yield New design performance slightly less than or equal to another.

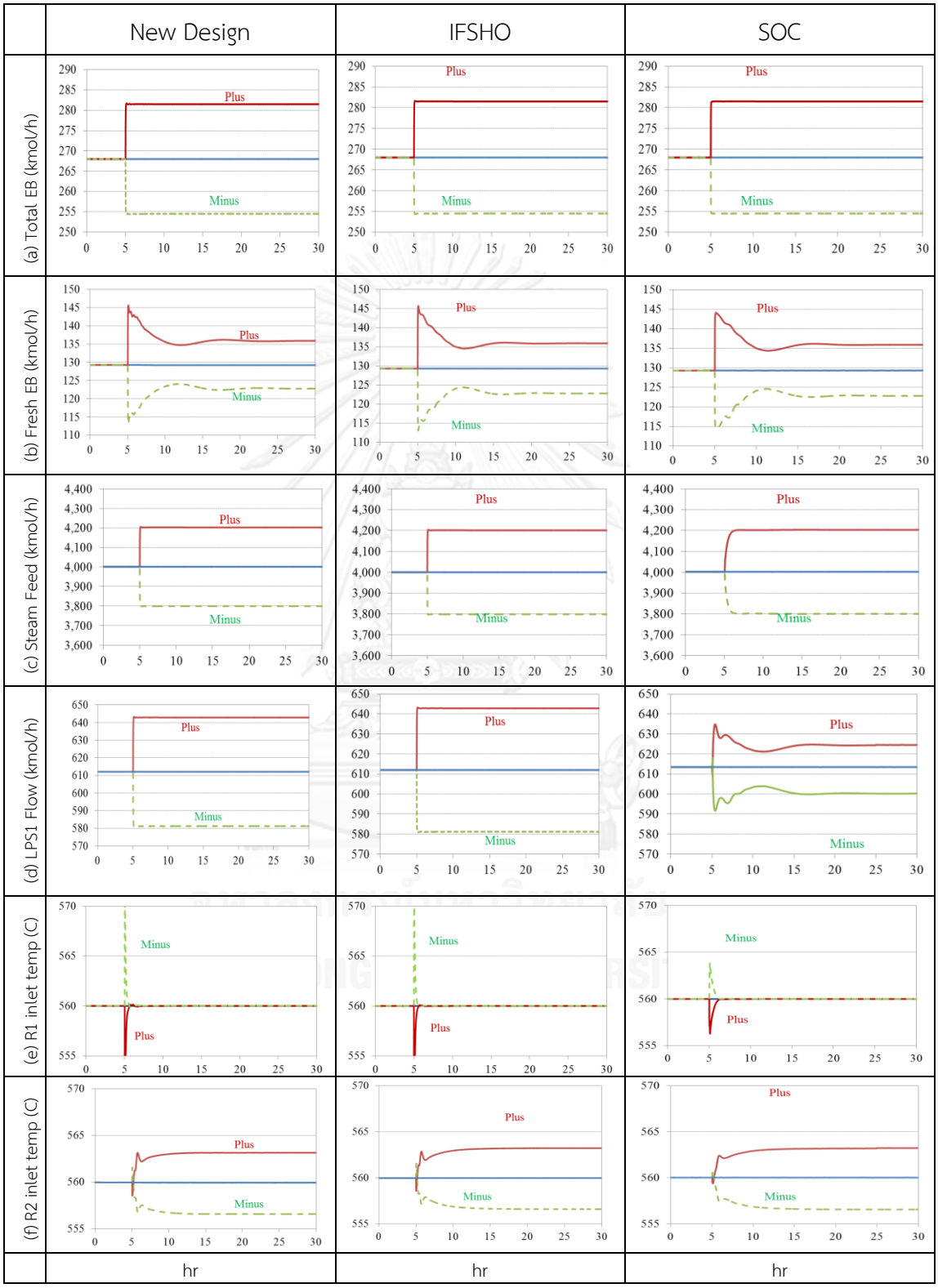


Figure 5.36A Dynamic Responses of scenario 1 and 2

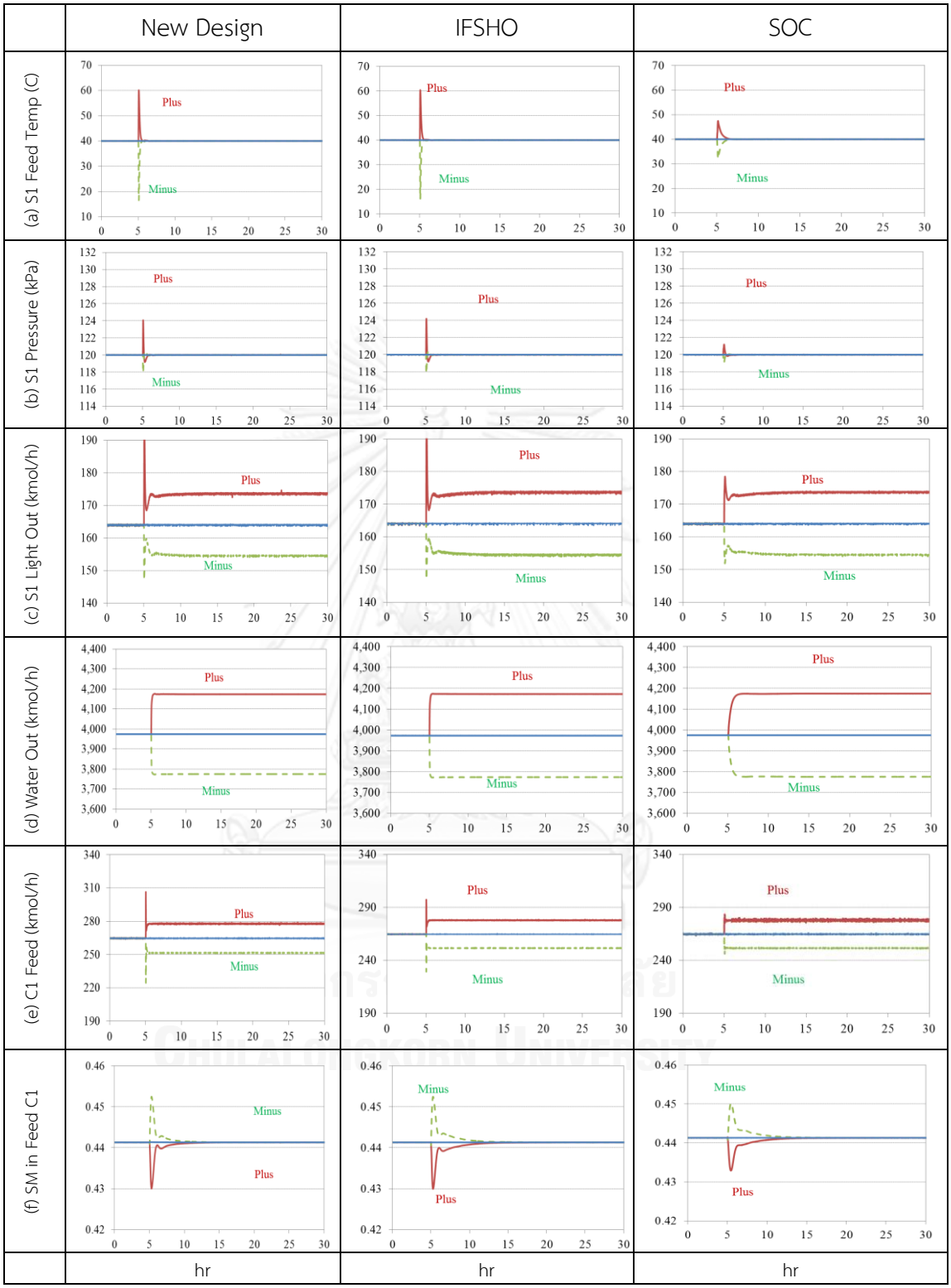


Figure 5.36B Dynamic Responses of scenario 1 and 2

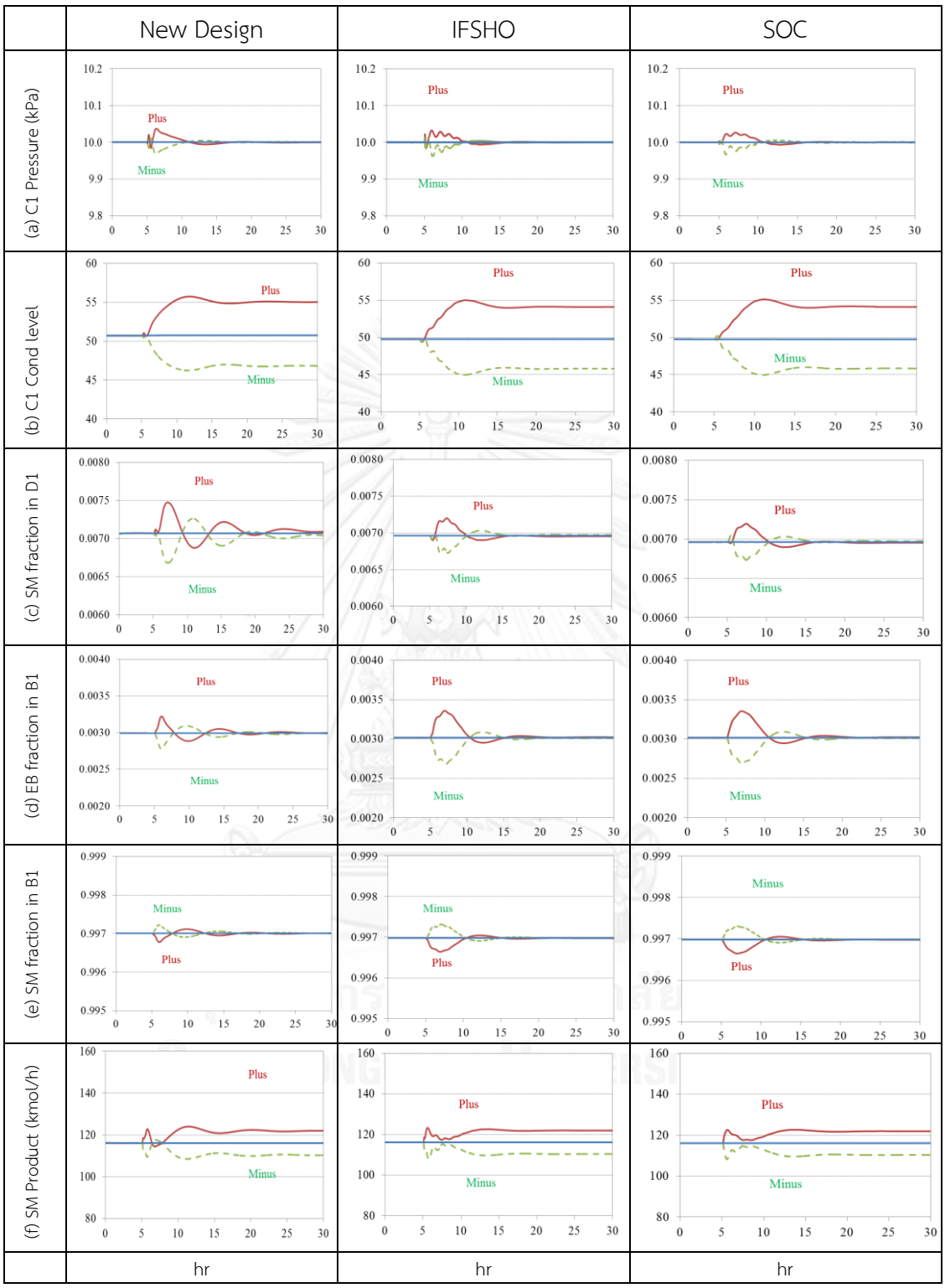


Figure 5.36C Dynamic Responses of scenario 1 and 2

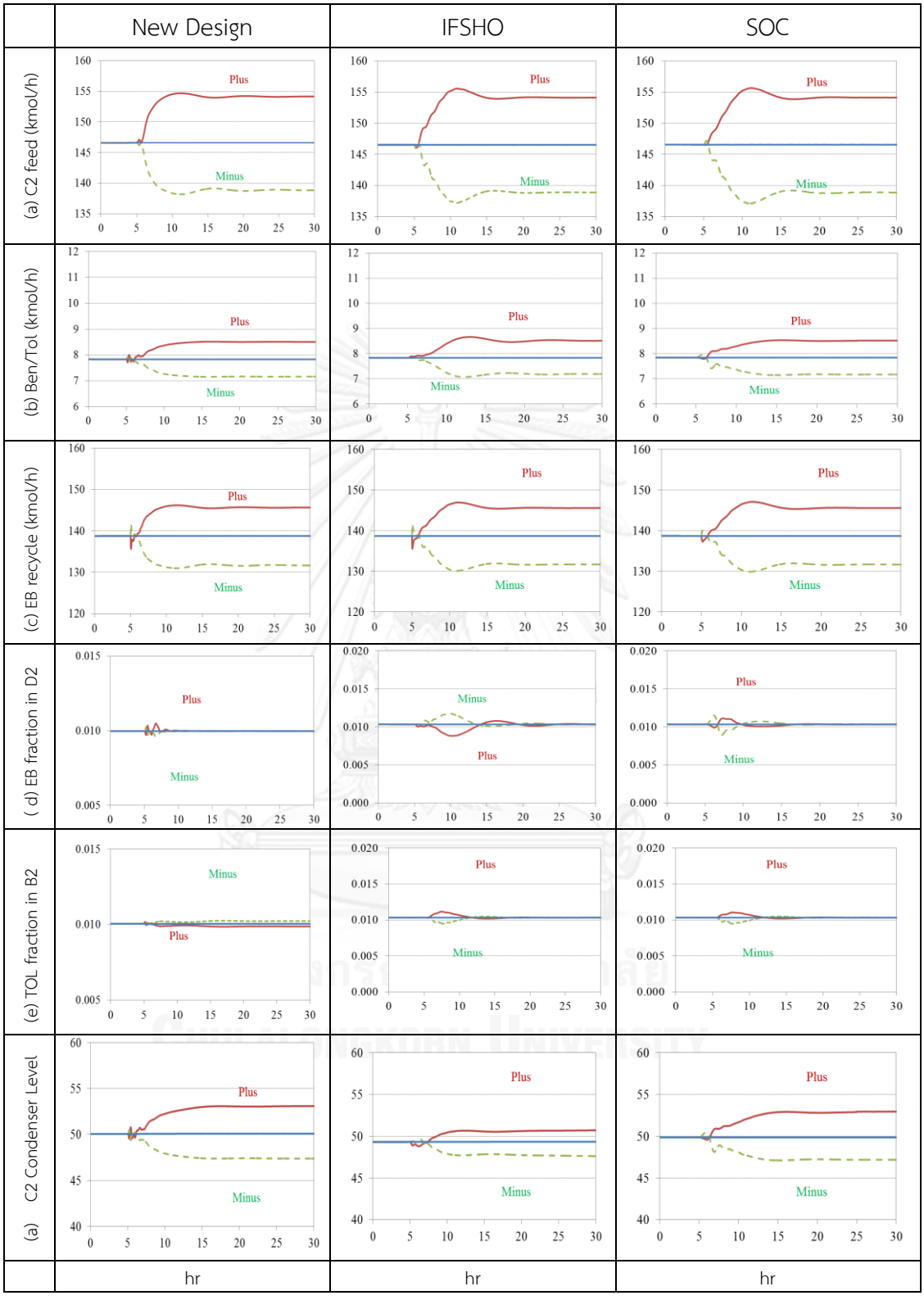


Figure 5.36D Dynamic Responses of scenario 1 and 2

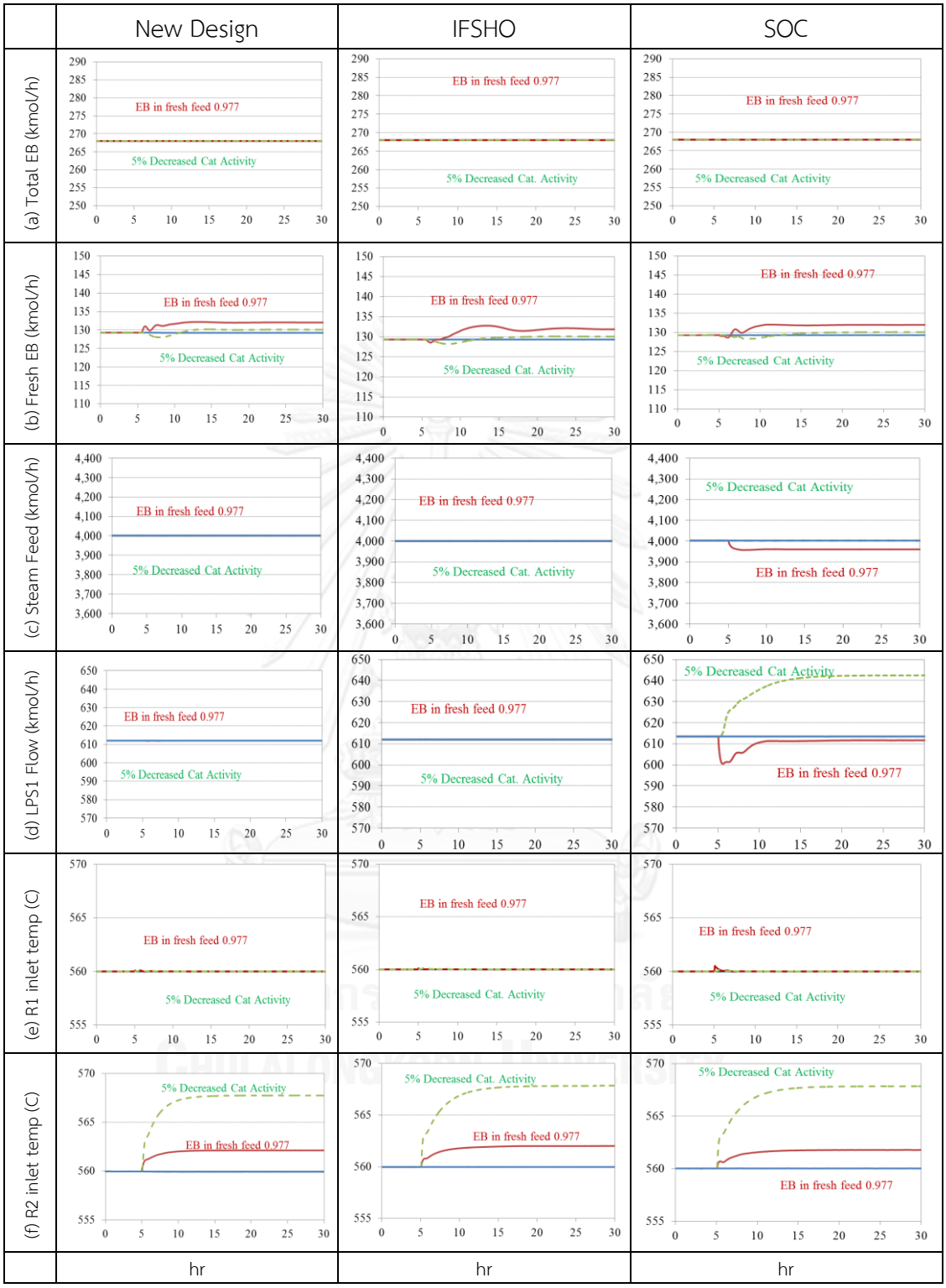


Figure 5.37A Dynamic Responses of scenario 3 and 4

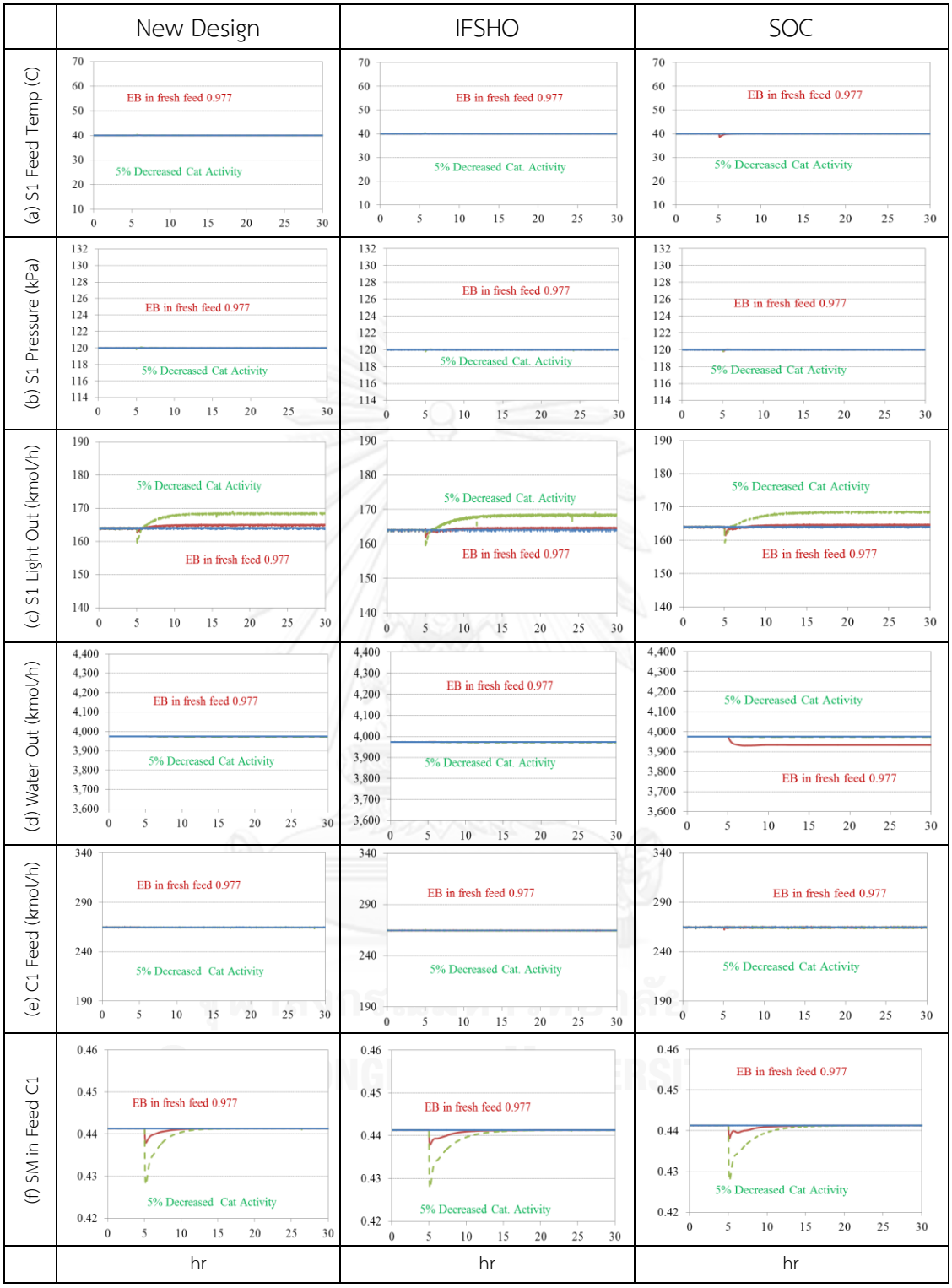


Figure 5.37B Dynamic Responses of scenario 3 and 4

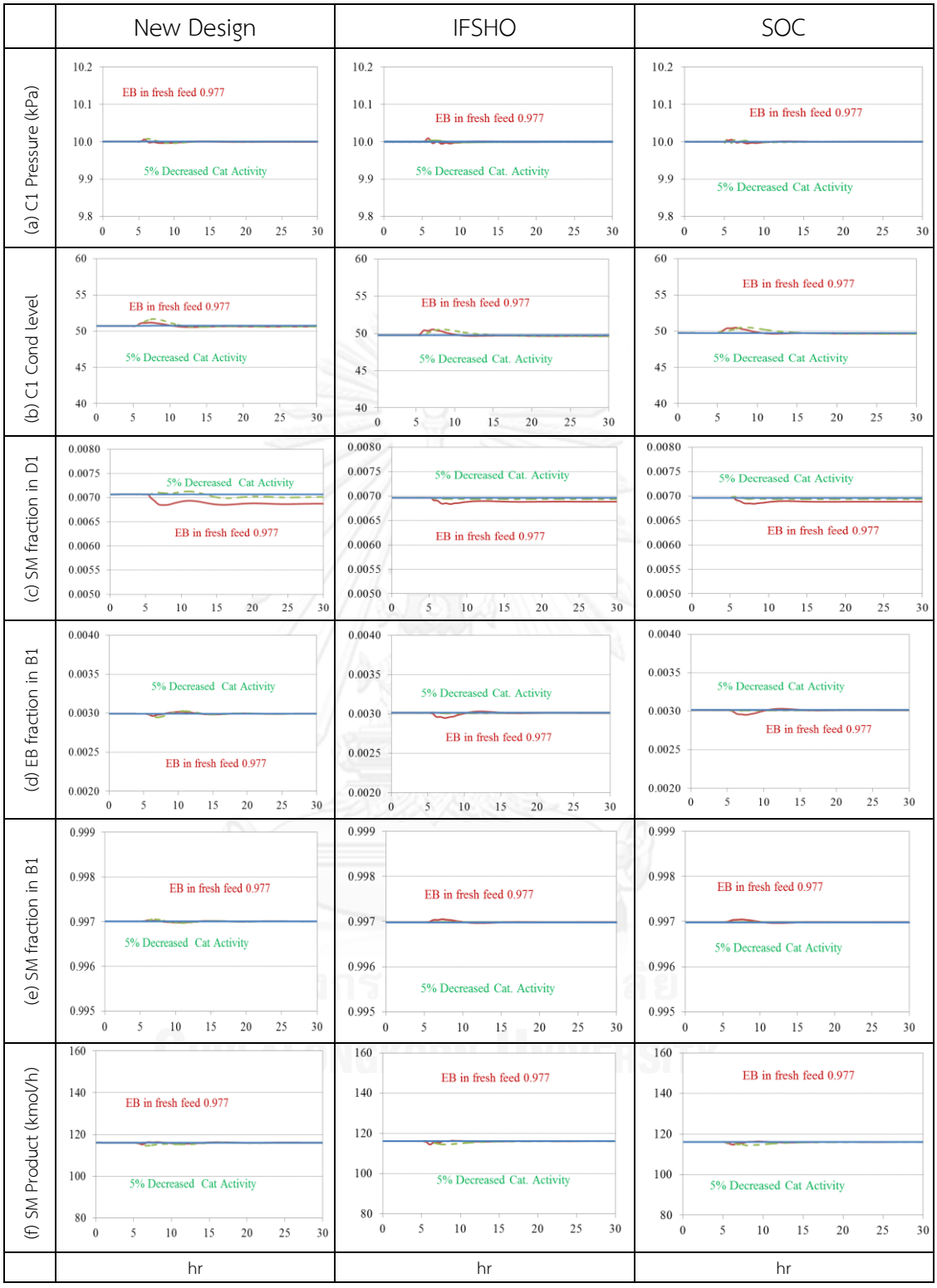


Figure 5.37C Dynamic Responses of scenario 3 and 4

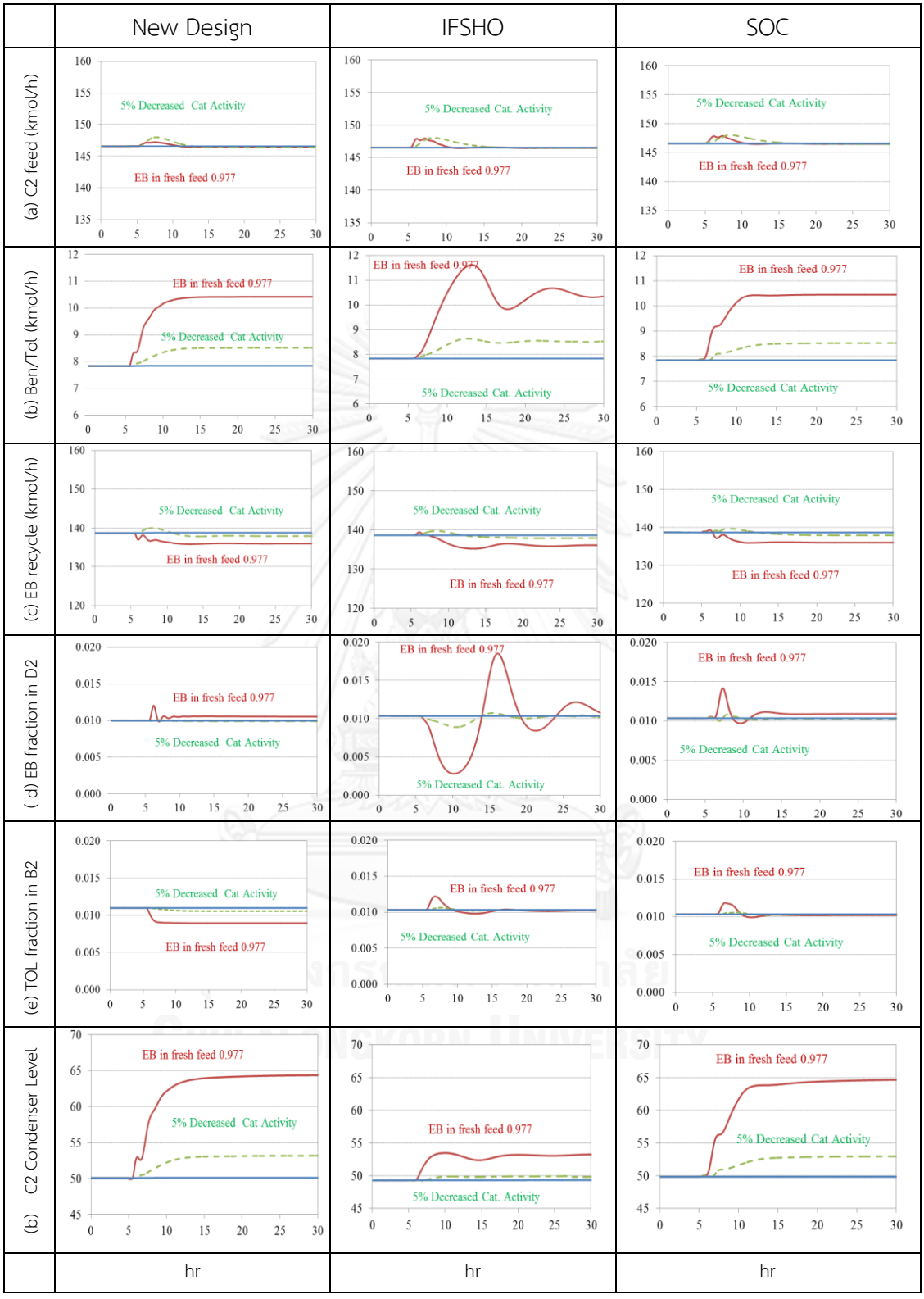


Figure 5.37D Dynamic Responses of scenario 3 and 4

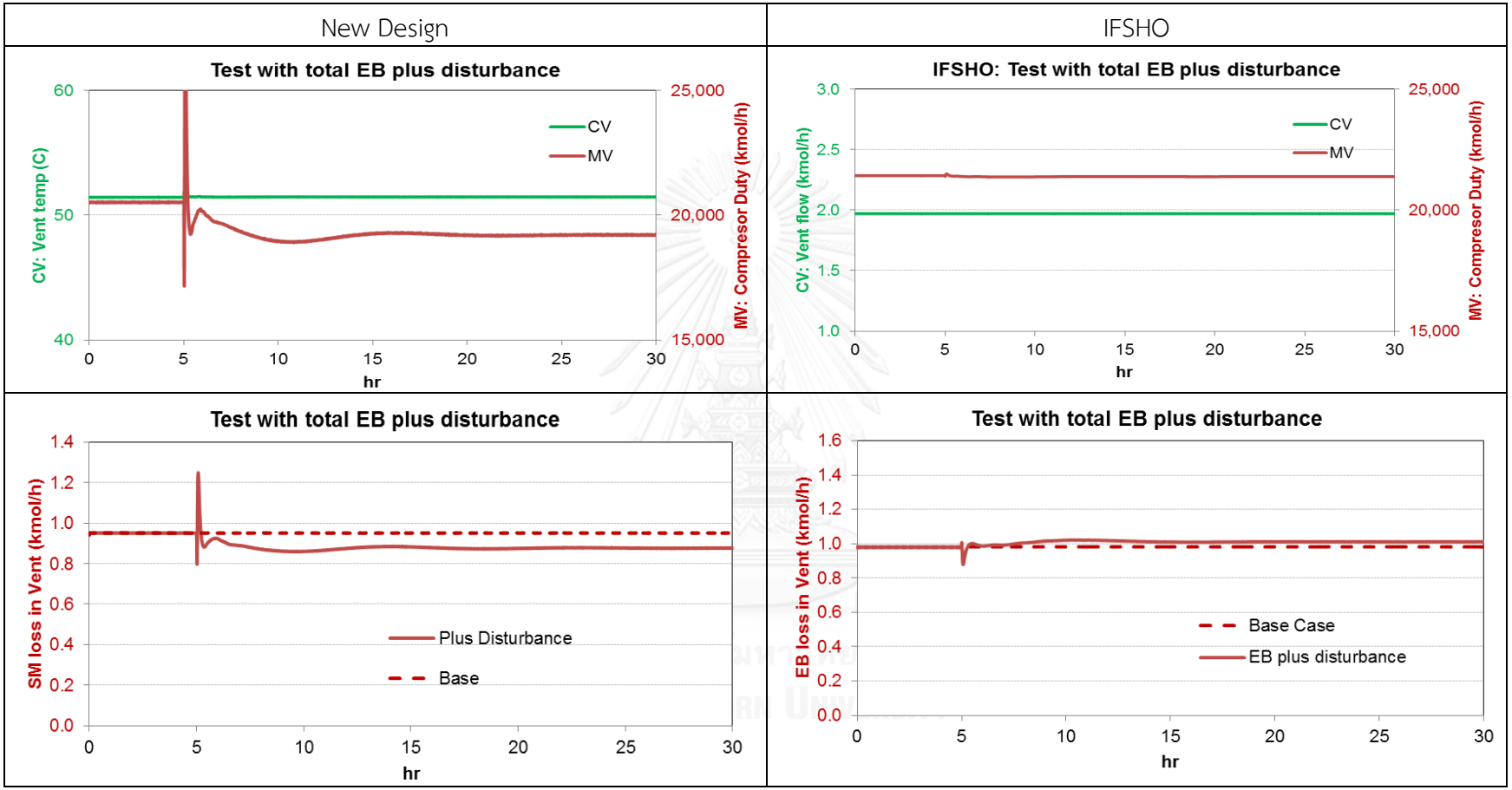


Figure 5.38A Performance of difference paring with the same MV (compressor duty)

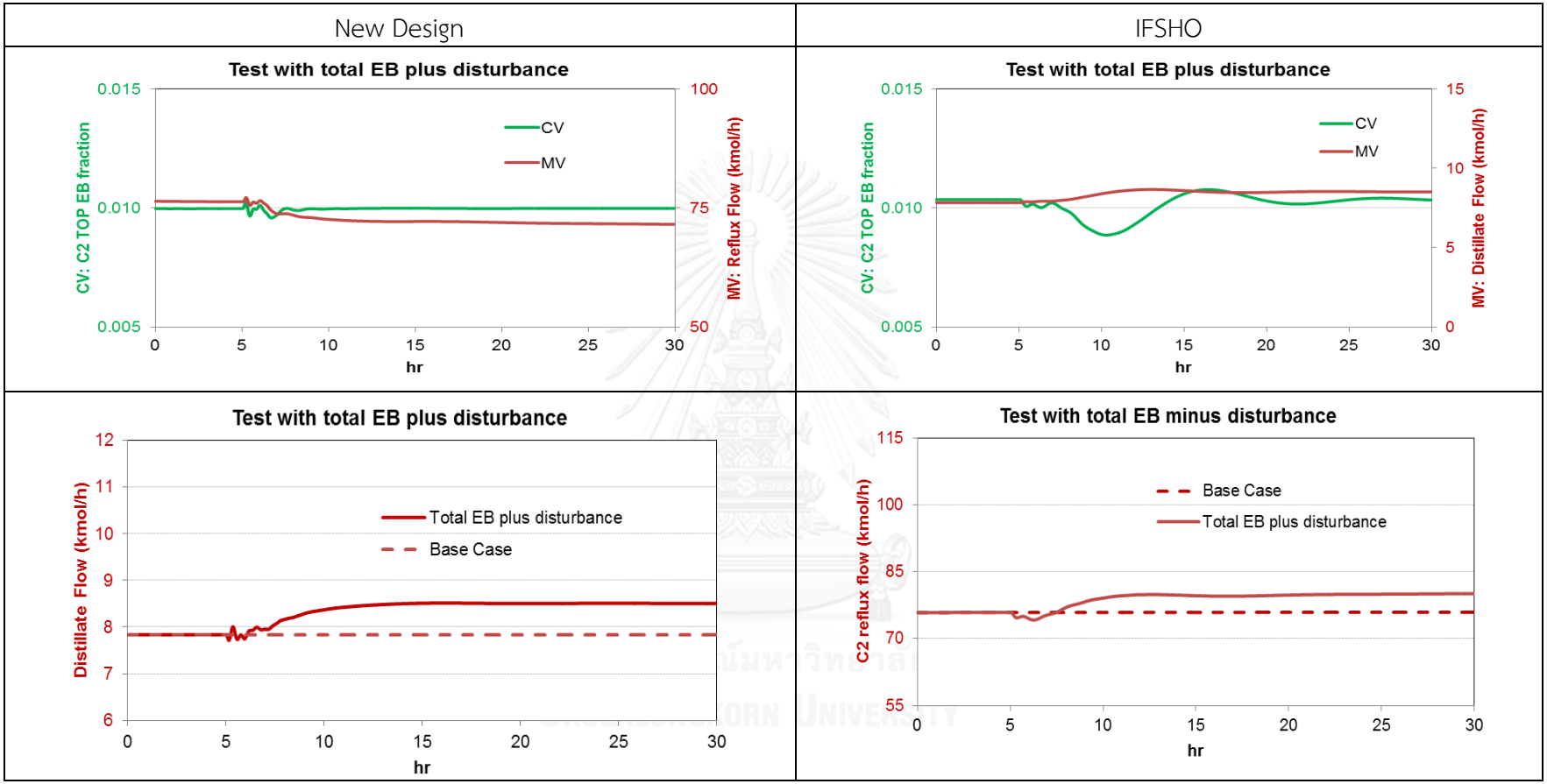


Figure 5.38B Performance of difference paring with the same CV (EB fraction on TOP of Column C2)

Table 5.6 IAE of major loops

Period 5-20 hrs	Total EB 281.5 kmol/h			Total EB 254.5 kmol/h			EB in feed -0.02 to 0.977			5% Decreased Cat Activity		
	New Design	**IFSHO**	**SOC**	**New Design**	**IFSHO**	**SOC**	**New Design**	**IFSHO**	**SOC**	**New Design**	**IFSHO**	**SOC**
Fresh EB	16.31	21.76	22.45	16.17	21.90	22.53	6.74	14.28	8.98	8.85	10.68	10.50
EB Recycle	16.35	21.71	22.50	16.21	21.85	22.55	6.72	14.27	8.96	8.83	10.68	10.50
Fresh Steam	4.23	2.15	72.08	3.82	1.94	70.23	0.58	1.79	20.75	0.38	0.29	3.68
SM Product	24.27	19.21	18.54	22.82	19.72	18.73	2.59	4.45	2.98	6.48	7.49	7.57
Sum IAE of Flow rate	**61.2**	**64.8**	**135.6**	**59.0**	**65.4**	**134.0**	**16.6**	**34.8**	**41.7**	**24.5**	**29.1**	**32.2**
R1 temperature	1.68	1.45	1.60	1.66	1.43	1.62	0.12	0.57	0.35	0.20	0.18	0.23
R2 temperature	3.54	5.01	5.60	3.61	5.05	5.64	2.80	5.82	3.75	12.25	15.84	16.05
Separator	3.39	3.14	3.28	3.38	3.13	3.40	0.15	1.02	0.70	0.22	0.22	0.23
Separator pressure	0.68	0.61	0.32	0.37	0.35	0.26	0.07	0.24	0.09	0.19	0.10	0.10
Sum IAE of Temp & Pressure	**9.3**	**10.2**	**10.8**	**9.0**	**10.0**	**10.9**	**3.1**	**7.6**	**4.9**	**12.9**	**16.3**	**16.6**
C1 SM in TOP	0.00149	0.00085	0.00087	0.00147	0.00085	0.00087	0.00045	0.00026	0.00021	0.00080	0.00004	0.00006
C1 EB in Bottom	0.00072	0.00386	0.00130	0.00068	0.00392	0.00128	0.00013	0.00520	0.00025	0.00020	0.00104	0.00005
C2 EB in TOP	0.00061	0.00656	0.00296	0.00051	0.00584	0.00501	0.00188	0.04195	0.00702	0.00047	0.00668	0.00146
C2 TOL in Bottom	0.00070	0.00310	0.00292	0.00072	0.00310	0.00324	0.00270	0.00564	0.00417	0.00153	0.00113	0.00095
Sum IAE of Distillation Part	**0.0035**	**0.0144**	**0.0080**	**0.0034**	**0.0137**	**0.0104**	**0.0052**	**0.0530**	**0.0116**	**0.0030**	**0.0089**	**0.0025**

Table 5.7 Comparison DDS

	Total EB 281.5 kmol/h			Total EB 254.5 kmol/h			EB fraction in feed 0.977			5% Decreased Cat Activity		
	New Design	IFSHO	SOC	New Design	IFSHO	SOC	New Design	IFSHO	SOC	New Design	IFSHO	SOC
DDS (kmol)	33.21	34.17	34.34	43.83	45.12	45.04	7.71	4.82	4.27	10.80	11.42	13.78
Margin ($/kg fresh EB)	0.186	0.186	0.186	0.188	0.188	0.190	0.182	0.175	0.183	0.184	0.177	0.184
Yield On production (%)	87.8	87.8	87.8	91.8	91.8	91.8	88.3	88.4	88.4	89.4	89.4	89.4
Total Fresh EB feed (5-20 hrs)	1,363	1,372	1,361	1,208	1,219	1,209	1,305	1,313	1,301	1,283	1,292	1,281
Total SM product (5-20 hrs)	1,197	1,205	1,195	1,109	1,119	1,110	1,152	1,160	1,150	1,147	1,155	1,145

CHAPTER VI
CONCLUTIONS AND RECOMMENDATIONS

6.1 Conclusion

New design structure of the styrene process using Wonsri's Design Procedure as described in Chapter III. The procedure is applied step-by-step in order to develop the regulatory control structure of the styrene process. The new step is about distillation control structure in Step 2.6, use steady-state simulation to the specific sensitive tray that use to control column and Step 4.2, uses dynamic simulation to select the appropriate distillation control structure.

The performance of the New design compared with structure of Vasudevan's IFSHO and Skogestad's SOC. The main differences of each structure summarize as follows. First, the column C1 dual end control cascade with reflux ratio and column C2 use dual end control cascade with a V/B ratio, while dual-end control is also applied in IFSHO and SOC, however the location that used to control separation are differences between IFSHO and SOC. Second, for steam feed, New design and IFSHO use ratio control with total EB flow while SOC use cascade control adjust steam flow setpoint by EB fraction in feed reactor R1. Third, New design control LPS1 flow by ratio with total EB feed and IFSHO ratio it with total steam flow while SOC cascade with furnace outlet temperature. Forth, IFSHO regulate EB fraction on top section by distillate flow while New design and SOC use reflux flow to regulate EB fraction on top. Finally, New design uses compressor feed temperature to regulate duty while IFSHO and SOC regulate compressor duty by vent flow rate.

The performance of control structure is subject to the expected disturbances. Four material disturbances are total EB flow ±5%, fresh EB feed composition decreased from 0.997 to 0.977 and catalyst deactivation, but does not apply heat disturbance because all control structures have same loops for control heat disturbance. There are no disturbance of extreme case such as total EB flow -20% due to new setpoint is needed to evaluate.

The performance results for all three alternative control structures are quantified in term of DDS and IAE. DDS developed by Konda and Rangaish, is a dynamic performance indicator in transient state and better performance is indicated by the lower value of DDS are shown in Table 5.7. For IAE, the important process parameters are select follow fixture plant concepts such as production rate of styrene and fresh EB feed flow, which IAE of each parameter show in Table 5.6. Both DDS and IAE are measures based on the transient state.

The result in Table 5.7 indicates that the DDS values are slightly lower for a New design, the DDS calculates from the mole balance of each component, while dynamic is running. In Table 5.6 show summary IAE of important parameters of the process of New design less than others.

The overall performance of New Design control structure developed in this work has good performance. However control engineers may probably give more weight to certain disturbances that more likely to occur and manager may give more weight to IAE of production rate because it is an economic performance indicator.

6.2 Recommendations

Study and design the control structure of the other process in plantwide control via the new design procedure of Wongsri (2014).

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APPENDIX

APPENDIX A

EQUIPMENT DATA AND STREAMS CONDITION

Table A1 Equipment Data

Units	Properties	Value
PFR Reactors	Diameter (m)	3.3
	Length (m)	8
	Void fraction	0.44
3-Phase Separator	Volume (m^2)	24
FEHE (E2)	UA (kJ/h-^{O}C)	3.0×10^5
Column C1	Number of stages	80
	Feed stage	35
	Top Pressure (kPa)	10
	Bottom Pressure (kPa)	50
	Diameter (m)	4.57
	Reflux Ratio	4.9
	Reboiler Volume (m^3)	24.49
	Condenser Volume (m^3)	19.43
Column C2	Number of stages	36
	Feed stage	16
	Top Pressure (kPa)	120
	Bottom Pressure (kPa)	138
	Diameter (m)	1.74
	Reflux Ratio	11
	Reboiler Volume (m^3)	7.88
	Condenser Volume (m^3)	2.38

Table A2 Stream conditions

Name	Fresh-EB	Total-EB	Cin	V3-out	LPS2	LPS1	LPS	Steam	C-hoted	R2cooled
Vapor Fraction	1.0000	0.4968	0.9846	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature (°C)	220.0	181.7	129.2	197.2	198.6	198.6	198.6	200.0	426.8	427.7
Pressure (kPa)	400	300	300	300	400	400	400	500	270	150
Molar Flow (kmol/h)	132.8	263.6	875.6	612.0	3,388.0	612.0	4,000.0	4,000.0	875.6	4,401.1
Mass Flow (kg/h)	14,088	27,965	38,991	11,025	61,035	11,025	72,060	72,060	38,991	100,026
Components Fraction										
Styrene	0.0000	0.0053	0.0016	0.0000	0.0000	0.0000	0.0000	0.0000	0.0016	0.0273
Ethylbenzene	0.9970	0.9906	0.2982	0.0000	0.0000	0.0000	0.0000	0.0000	0.2982	0.0303
Hydrogen	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0328
Benzene	0.0030	0.0015	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0007
Ethylene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Toluene	0.0000	0.0026	0.0008	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0017
Methane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005
H2O	0.0000	0.0000	0.6989	1.0000	1.0000	1.0000	1.0000	1.0000	0.6989	0.9046
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021

Table A2 (Continuous) Stream conditions

Name	R1_inlet	LPS2Out	R1 outlet	R2 inlet	R2 outlet	to separator	Light	Organics	Water	Light Out
Vapor Fraction	1.0000	1.0000	1.0000	1.0000	1.0000	0.0387	1.0000	0.0000	0.0000	1.0000
Temperature (^{o}C)	560.0	654.4	517.4	560.0	535.5	40.0	40.0	40.0	40.0	40.0
Pressure (kPa)	270	270	240	210	180	120	120	120	120	100
Molar Flow (kmol/h)	4,263.6	3,388.0	4,349.1	4,349.1	4,401.1	4,401.1	170.4	259.8	3,970.9	170.4
Mass Flow (kg/h)	100,026	61,035	100,026	100,026	100,026	100,026	1,354	27,134	71,538	1,354
Components Fraction										
Styrene	0.0003	0.0000	0.0180	0.0180	0.0273	0.0273	0.0081	0.4575	0.0000	0.0081
Ethylbenzene	0.0613	0.0000	0.0414	0.0414	0.0303	0.0303	0.0131	0.5039	0.0000	0.0131
Hydrogen	0.0000	0.0000	0.0203	0.0203	0.0328	0.0328	0.8466	0.0004	0.0000	0.8466
Benzene	0.0001	0.0000	0.0004	0.0004	0.0007	0.0007	0.0022	0.0099	0.0000	0.0022
Ethylene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0002
Toluene	0.0002	0.0000	0.0008	0.0008	0.0017	0.0017	0.0019	0.0268	0.0000	0.0019
Methane	0.0000	0.0000	0.0003	0.0003	0.0005	0.0005	0.0135	0.0001	0.0000	0.0135
H2O	0.9382	1.0000	0.9178	0.9178	0.9046	0.9046	0.0612	0.0008	1.0000	0.0612
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0000	0.0000	0.0005
CO2	0.0000	0.0000	0.0010	0.0010	0.0021	0.0021	0.0527	0.0006	0.0000	0.0527

Table A2 (Continuous) Stream conditions

Name	SM	Vent	P5 out	Tol/Ben	P6 out	V11 out	P1 out	P out	EB recycle	V1 Out
Vapor Fraction	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000
Temperature (°C)	120.7	145.3	86.3	86.4	148.0	148.0	40.2	40.0	148.0	218.6
Pressure (kPa)	100	150	300	200	400	300	524	300	300	300
Molar Flow (kmol/h)	117.7	1.9	8.6	8.6	131.6	131.6	259.8	3,970.9	130.8	132.8
Mass Flow (kg/h)	12,264	154	760	760	13,957	13,957	27,134	71,538	13,878	14,088
Components Fraction										
Styrene	0.9975	0.0045	0.0000	0.0000	0.0107	0.0107	0.4575	0.0000	0.0107	0.0000
Ethylbenzene	0.0025	0.5875	0.0100	0.0100	0.9837	0.9837	0.5039	0.0000	0.9841	0.9970
Hydrogen	0.0000	0.0539	0.0000	0.0000	0.0000	0.0000	0.0004	0.0000	0.0000	0.0000
Benzene	0.0000	0.0808	0.2820	0.2820	0.0000	0.0000	0.0099	0.0000	0.0000	0.0030
Ethylene	0.0000	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Toluene	0.0000	0.0784	0.7051	0.7051	0.0057	0.0057	0.0268	0.0000	0.0052	0.0000
Methane	0.0000	0.0068	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000
H2O	0.0000	0.1079	0.0017	0.0017	0.0000	0.0000	0.0008	1.0000	0.0000	0.0000
CO	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO2	0.0000	0.0796	0.0011	0.0011	0.0000	0.0000	0.0006	0.0000	0.0000	0.0000

APPENDIX B
TUNING OF CONTROL STRUCTURES

B.1 Tuning Controllers

Notice throughout this work uses several types of controllers such as P and PI but not use PID because It has best performance in program simulation but hard to achieve in real process. Controller type depends 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 care 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 better 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 τ_I = 0.3 minutes work in most controllers. The value of controller gain should be kept modest because the flow measurement signal is 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 flow rate of gas flowing through the system. Setting the integral time equal to about 2 to 4 times the process time constant and using a reasonable controller gain usually gives satisfactory pressure control. Typical pressure controller tuning constants for columns and tanks are K_C = 2 and τ_I = 10 minutes.

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.

$$K_U = \frac{4h}{a\pi} \tag{1}$$

The period of the output PV curve is the ultimate period, PU 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:

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

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

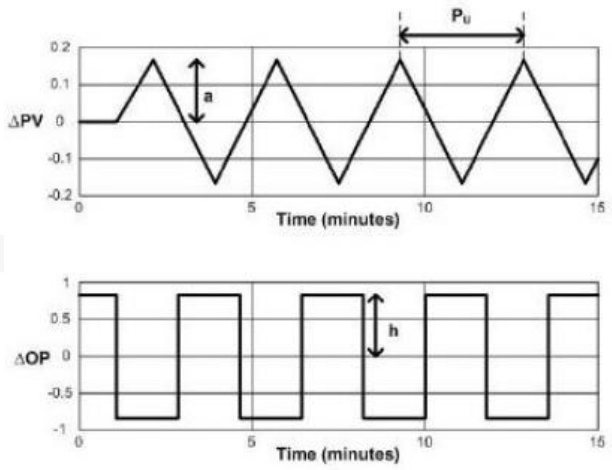

Figure B1 Input and Output from Relay-Feedback Test (Luyben, W., Plantwide Dynamic Simulations in Chemical Processing and Control, p. 30, 1998.)

Table B1 Type of controllers and Tuning parameters of design control structure

MV	CV	Type	Action	Normal Value	Range	Kc	Ti
Fresh EB flow	Total EB flow	PI	Reverse	268 kmol/h	175-360	0.5	0.3
Steam flow	Ratio to total EB	PI	Reverse	4,000 kmol/h	3,500-4,500	0.098	0.0271
LPS1 flow	Ratio with total EB	PI	Reverse	612 kmol/h	512-712	0.243	0.00253
Furnace duty	R1 inlet temp	PI	Reverse	560°C	500-620	0.0033	1.15
Inter-heater duty	R2 inlet temp	PI	Reverse	560°C	500-620	0.248	1.12
R2 inlet temp SP (cascade)	SM in feed C1	PI	Reverse	0.4413	0.35 - 0.55	0.261	27.5
cooler duty	S1 feed temp	PI	Direct	40°C	0 - 80	0.0228	2.19
Light flow	S1 Pressure	PI	Direct	120 kPa	100 - 140	2	10
Water flow	S1 Water level	P	Direct	0.3 m	0 - 0.6	2	
C1 feed flow	S1 Organic level	P	Direct	20%	0 - 40	2	
C1 condenser duty	C1 pressure	PI	Direct	10 kPa	5 - 15	2	10
C1 Distillate flow	C1 condenser level	P	Direct	50%	0 - 100	2	
C1 Reflux flow	C1 Reflux Ratio	PI	Reverse	6.324	3.6 - 9.6	0.114	0.00682
C1 RR SP (cascade)	C1 SM tray 18	PI	Direct	0.166	0 - 0.3	0.334	15.1
C1 reboiler duty	C1 EB tray 53	PI	Direct	0.2152	0 - 0.4	0.173	31
C1 reboiler duty (cascade)	C1 SM tray 80	PI	Reverse	0.9963	0.994 - 1.0	0.331	32.5
SM product flow	C1 bottom level	P	Direct	50%	0 - 100	2	
K1 duty	K1 inlet temp	PI	Reverse	51.5°C	0 - 100	7.59	0.301
C2 condenser duty	C2 pressure	PI	Direct	120 kPa	70 - 170	2	10
Bz/tol flow	C2 condenser level	P	Direct	50%	0 - 100	2	
C2 Reflux flow	C2 EB on top	PI	Direct	0.0241	0 - 0.03	0.114	10
C2 reboiler duty	C2 VB ratio	PI	Reverse	1.26	0 - 2.6	1.3	0.087
C2 boiler duty (cascade)	Toluene tray 28	PI	Direct	0.052	0 - 0.13	1.01	3.25
EB recycle flow	C2 bottom level	P	Direct	50%	0 - 100	2	

Table B2 Compare control structure and tuning parameters

MVs	Wongsri			IFSHO			SOC		
	CV	Kc	T	CV	Kc	T	CV	Kc	T
Fresh EB flow	Total EB flow	0.5	0.3	Total EB flow	0.5	0.3	Total EB flow	0.5	0.3
Steam flow	Ratio to total EB	0.098	0.0271	Ratio to total EB	0.099	0.027	EB fraction in feed R1	0.257	3.61
LPS1 flow	Ratio with total EB	0.243	0.00253	Ratio with fresh steam	0.0986	0.0248	furnace outlet temp	0.198	1.12
Furnace duty	R1 inlet temp	0.0033	1.15	R1 inlet temp	0.033	1.13	R1 inlet temp	0.033	1.13
Inter-heater duty	R2 inlet temp	0.248	1.12	R2 inlet temp	0.255	1.15	R2 inlet temp	0.255	1.15
R2 inlet temp SP (cascade)	SM in feed C1	0.261	27.5	SM in feed C1	0.289	39.4	SM in feed C1	0.289	39.4
cooler duty	S1 feed temp	0.0228	2.19	S1 feed temp	0.0226	2.13	S1 feed temp	0.0226	2.13
Light flow	S1 Pressure	2	10	S1 Pressure	2	10	S1 Pressure	2	10
Water flow	S1 Water level	2		S1 Water level	2		S1 Water level	2	
C1 feed flow	S1 Organic level	2		S1 Organic level	2		S1 Organic level	2	
C1 condenser duty	C1 pressure	2	10	C1 pressure	2	10	C1 pressure	2	10
C1 Distillate flow	C1 condenser level	2		C1 condenser level	2		C1 condenser level	2	
C1 Reflux flow	C1 Reflux Ratio	0.114	0.00682	C1 SM on top	0.5	27.3	C1 SM on top	0.5	27.3
C1 RR SP (cascade)	C1 SM tray 18	0.334	15.1						
C1 reboiler duty	C1 EB tray 53	0.173	31	C1 EB in bottoms	0.23	54	C1 EB in bottoms	0.23	54
C1 reboiler duty (cascade)	C1 SM tray 80	0.331	32.5						
SM product flow	C1 bottom level	2		C1 bottom level	2.2		C1 bottom level	2.2	
K1 duty	K1 inlet temp	7.59	0.301	Vent flow	0.236	0.0173	Vent flow	0.236	0.0173
C2 condenser duty	C2 pressure	2	10	C2 pressure	2	10	C2 pressure	2	10
Bz/tol flow	C2 condenser level	2		C2 EB on top	0.212	56.2	C2 condenser level	2	
C2 Reflux flow	C2 EB on top	0.114	10	C2 Condenser level	2		C2 temp tray 6	0.0742	14.2
C2 temp tray 6 SP (cascade)							C2 EB on top	0.135	26.3
C2 reboiler duty	C2 VB ratio	1.3	0.087	C2 toluene in bottom	0.071	16.6	C2 toluene in bottom	0.071	16.6
C2 boiler duty (cascade)	Toluene tray 28	1.01	3.25						
EB recycle flow	C2 bottom level	2		C2 bottom level	2		C2 bottom level	2	

APPENDIX C

DETAIL OF DISTILLATION CONTROL STRUCTURE

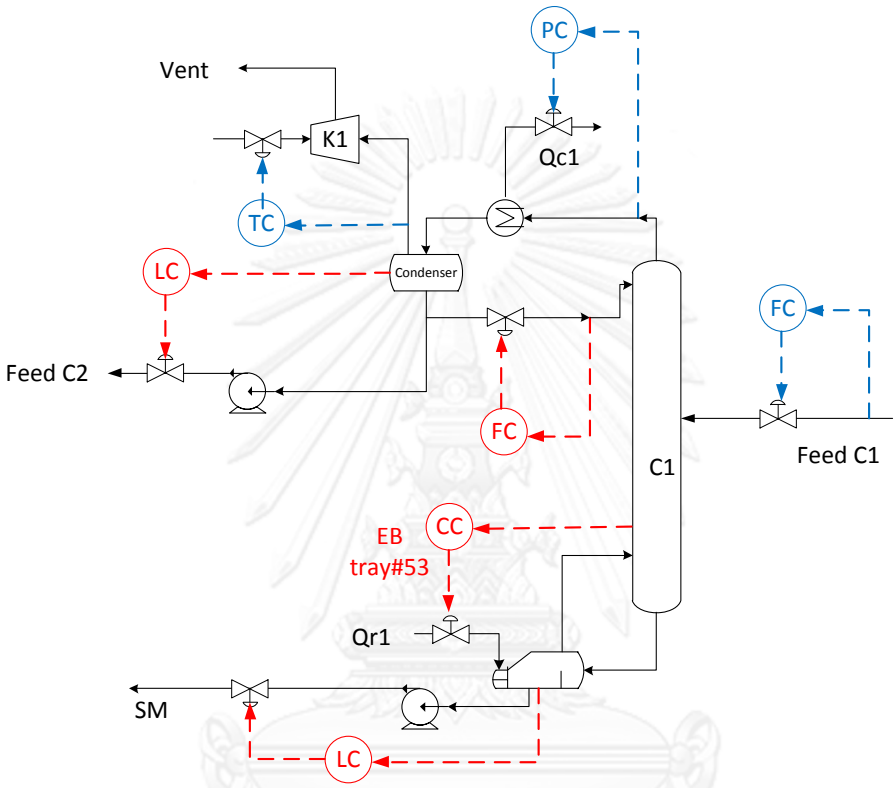

Figure C1 Column C1 Fix Reflux Flow

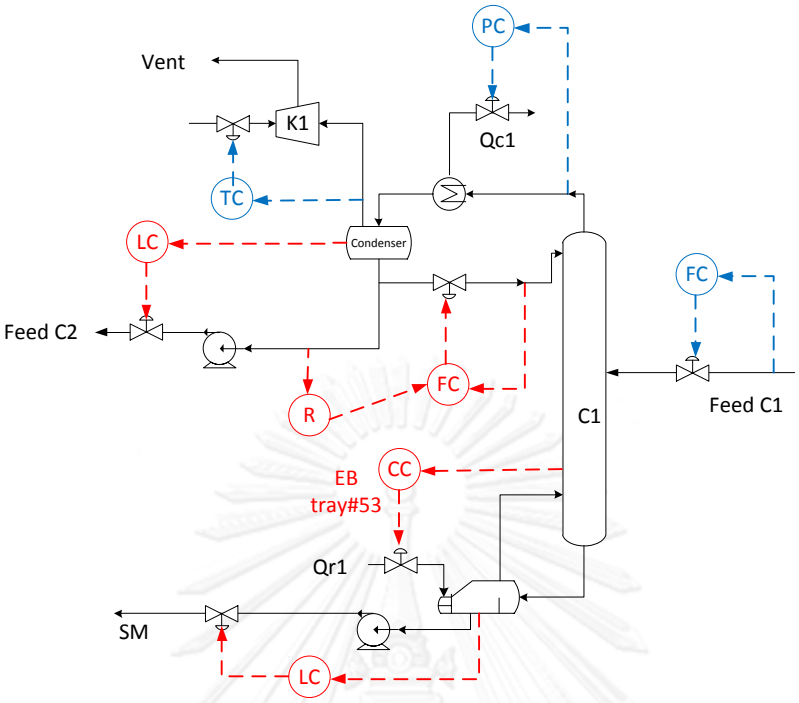

Figure C2 Column C1 Fix Reflux Ratio

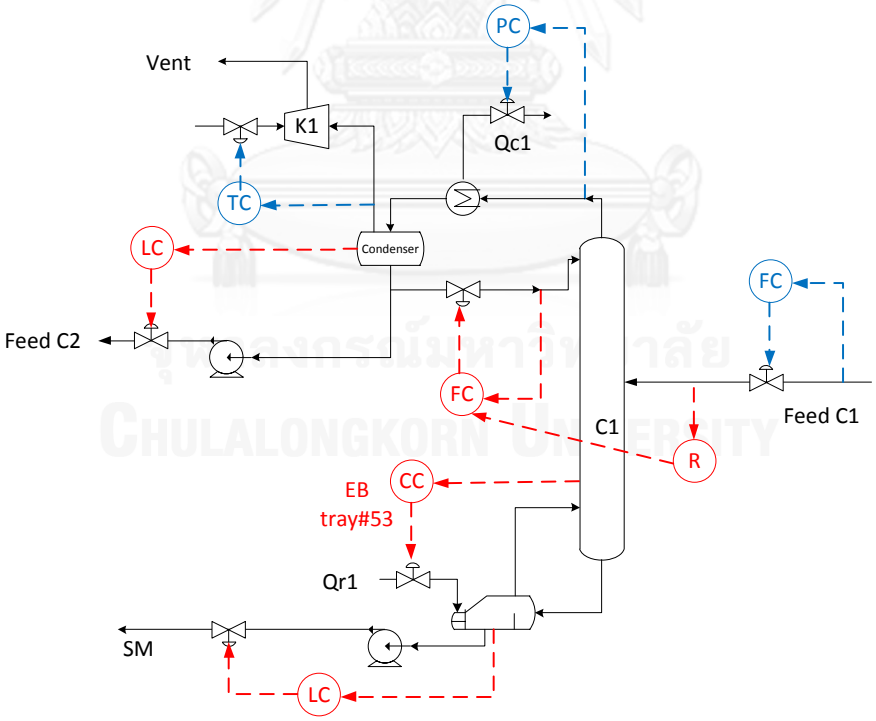

Figure C3 Column C1 Fix Reflux to Feed Ratio

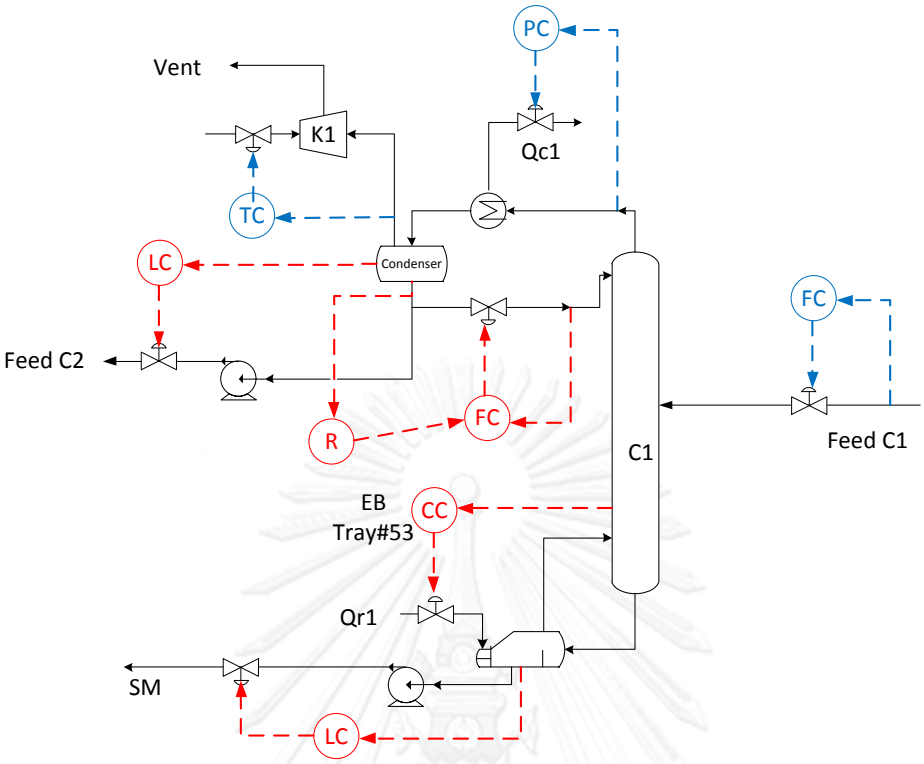

Figure C4 Column C1 Fix R/(R+D)

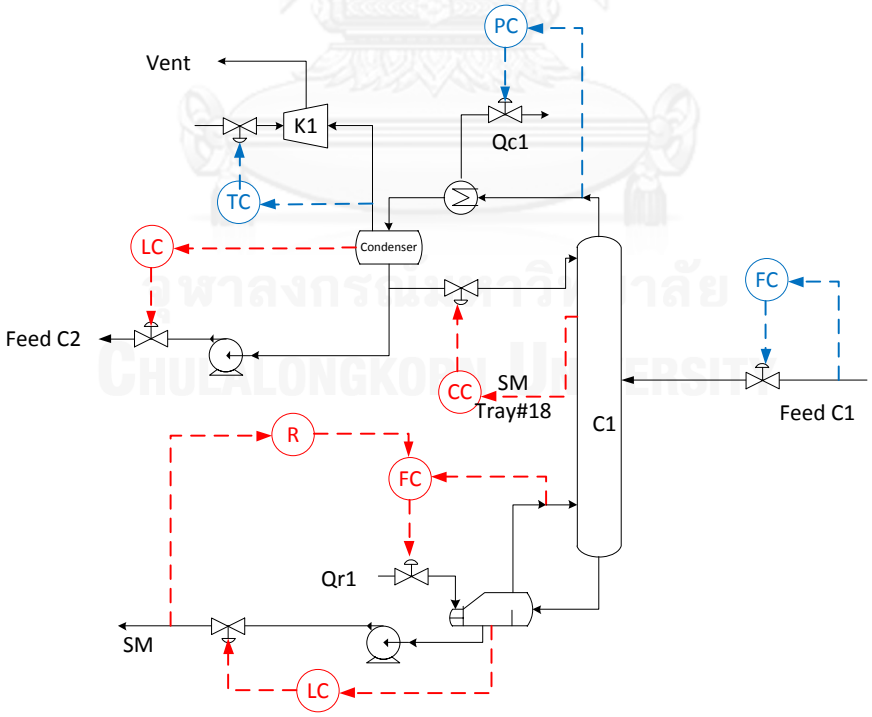

Figure C5 Column C1 Fix V/B

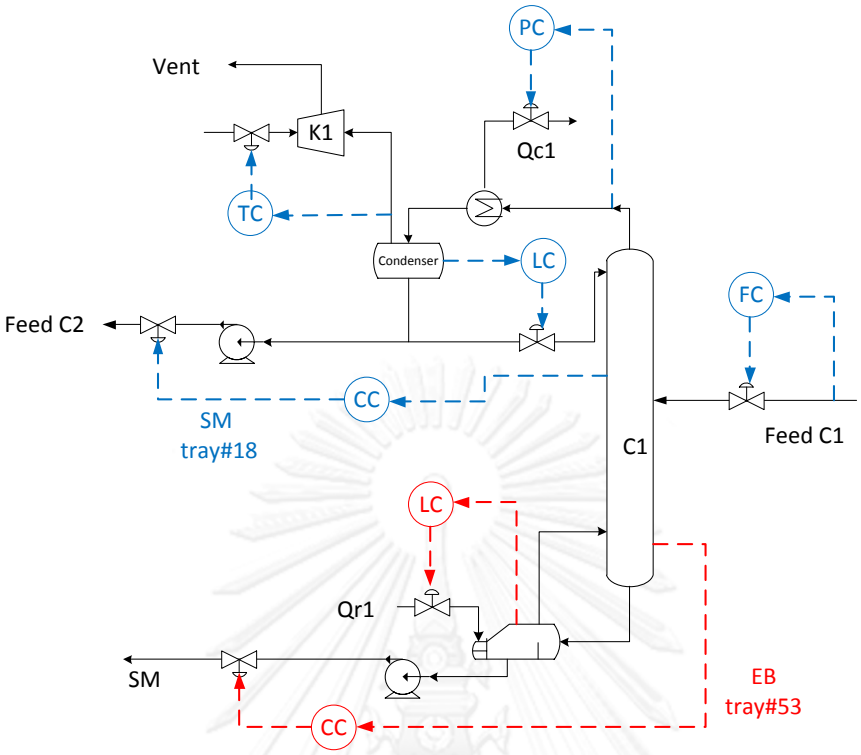

Figure C6 Column C1 Dual1

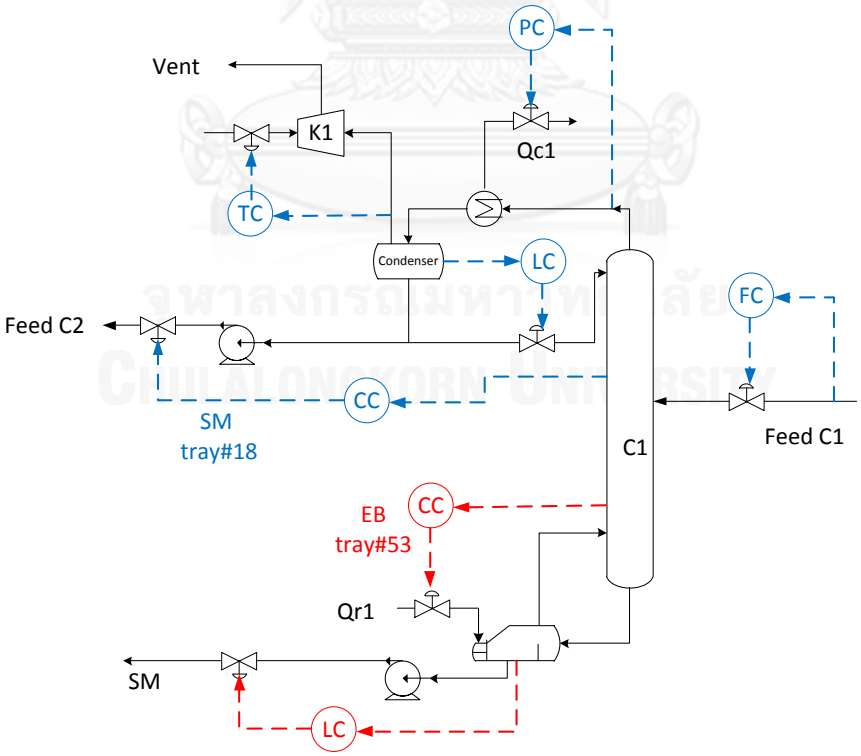

Figure C7 Column C1 Dual2

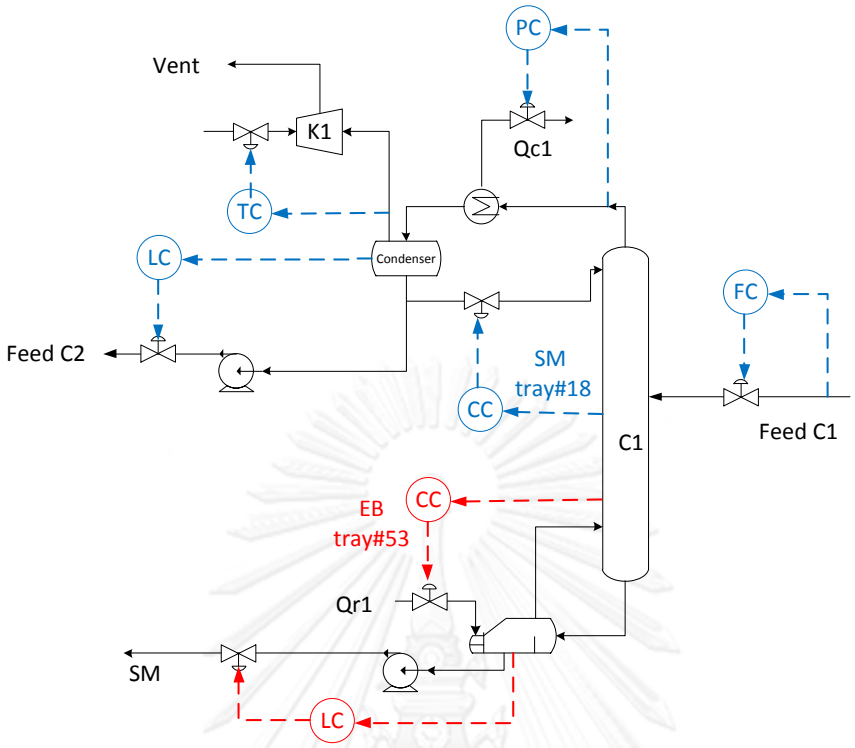

Figure C8 Column C1 Dual3

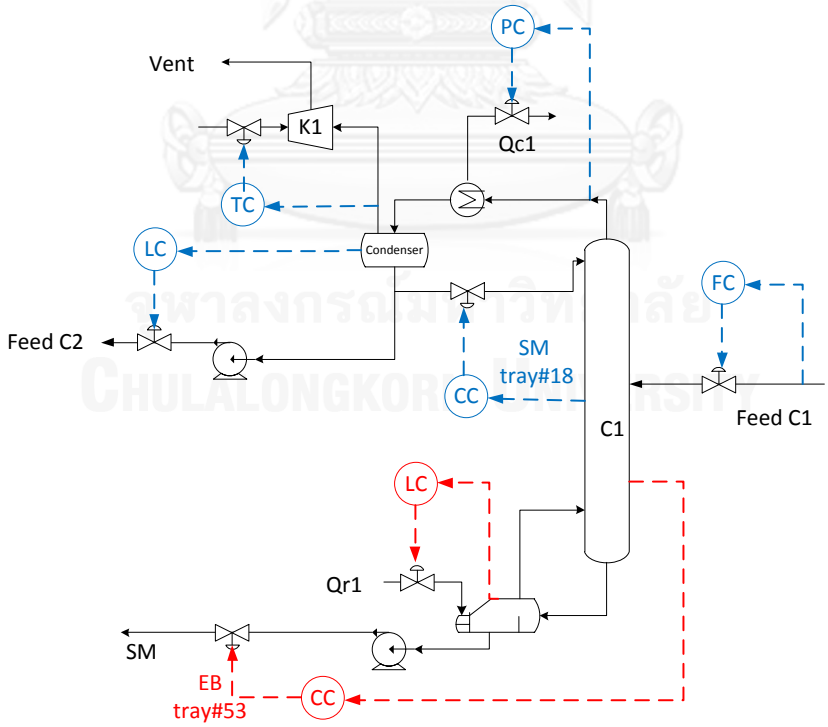

Figure C9 Column C1 Dual4

APPENDIX D
RESEARCH REVIEW OF WONGSRI

Boonserm Sophonudomsub, *Plantwide Control Structure Design For Ammonia Production Process*, 2012

The thesis presents the application of the plantwide control design procedure of Wongsri (2012) to the ammonia production process. Two new control structures obtained are evaluated and compared with the control structure of Skogestad (2008). The disturbances used in the evaluation are gas feed flow rate, feed temperature, and methane composition in feed changes. The new control structure, CS2, resulted in good dynamic performance for the feed flow rate and feed temperature changes. For methane composition in feed change, all control structures give comparable good performances.

Kantarakorn Katawetitathum, *Plantwide control structure design of the methoxy-methyl-heptane process*, 2012

The design of plantwide control structure for the methoxy-methyl-heptane process using Wongsri's design procedure is studied. The eight steps of the design procedure focus on plantwide level design which is establishing a fixture plant and thermal and material disturbance managements for quality control. The process involves the reaction of methanol with 2-methyl-1-heptene to form 2-methoxy-2 methylheptane. The combined feed of fresh and recycled 2-methyl-1-heptene is regulated and is ratioed to the fresh methanol. 2-methyl-1-heptene disturbance is pushed away via the distillate of the second column and recycled to maintain 2-methoxy-2-methylheptane purity in the third column. All components are handled at their identified quantifiers. The designed plantwide control structure performance is compared with the Luyben's design.

Napaporn Plonprasert, *Plantwide control structure design of ethyl benzene using fixture point theorem*, 2012

This research is applied "Fixture Point Theorem" to ethyl benzene process to design a new plantwide control structures that be compared with Luyben (2002). The principle of the fixture point theorem is the most disturbed variables were affected by manipulated variables or disturbances change should be precedence consideration. We selected two sets of controlled variables and three control

structures were designed and compared. The dynamic behaviors of designed control structures were illustrated whether manipulated variables change and the performance were presented in IAE value. The design control structure has a good performance and energy use reduction in the process. This research established that the Wongsri's procedure lead to a good performance plantwide control system.

Pira Kanchanawong, *Plantwide control structure design for modified ethyl benzene process*, 2012

In this thesis, plantwide control structure design of ethylbenzene (EB) process has been investigated. The ethylbenzene process incorporates the reaction of benzene with ethylene to form the desired EB product and the undesired di-ethyl benzene (DEB) by-product. The DEB is continuously recycled with no exit or recycled to "extinction" (no net DEB product produced). The optimum flowsheet is obtained from Luyben (2010). Two effective plantwide control structures are developed using Wongsri's procedure. The main emphases of the procedure are establishing a fixture plant (material-balanced control and fixed process inputs) and disturbance management. The results show that the design structures have better control performances compared to Luyben (2010).

Thitima Tapaneeyapong, *Plantwide control structure design of tert-amyl methyl ether (TAME) process*, 2012

The design of plantwide control structure for tert-amyl methyl ether process using Wingsir's design procedure is reported. The design procedure consists of 8 steps which emphasis on plantwide level design: establishing material-balanced plant, disturbance management for quality control, fixing process stream inflows. The combined feed of fresh and recycled methanol is regulated and is ratioed to isoamylene flow. Each component is handled at their quantifiers. The designed plantwide control structure performance is comparable to the Luyben's design.

Choksakunt Arrayasinlapathorn, *Plantwide control structures design of methyl acetate process*, 2011

Methyl acetate is a fast-evaporating chemical component with many uses as solvent. For its production, methanol is first dehydrated to dimethyl ether and then carbonylated to methyl acetate. The two sections of the plant, dehydration and carbonylation section are separately considered. Therefore, the plantwide control strategy is considered for the entire methyl acetate process including two sections

above. The control design methodology, Wongsri is applied to the plantwide control structures design for reach the control objectives and effectively operating large disturbances in production rate and fresh feed compositions. A commercial process simulator achieves the dynamics simulation of this research, both of steady-state and dynamic conditions.

Kanika Phetyodsri, *Plantwide control structures for methanol process*, 2011

This research focused on the development control structures for methanol process by using Wongsri's procedure (2009). Three control structures (CS1 to CS3) are designed and evaluated dynamic performance to compare with the proposed control structure of Luyben (2010) by using material and thermal disturbances. The result shows that the dynamic responses of the designed control structures and base control structure are similar as compared by integral absolute error (IAE). All designed control structures with heat integration can reduce the energy consumption, can reject disturbances, and can maintain product quality close to their specified values.

Keadsuda Machuay, *Plantwide control structure design of styrene process*, 2011

Plantwide control procedure of Wongsri (2009) is proposed and applied to the styrene process. Four plantwide control structures (CS1 to CS4), all designed control structures have evaluated the dynamic 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 better quality product 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.

Panisara Khamanarm, *Control structures design applied to alkylation process plantwide control*, 2011

The study is using plantwide control strategies for designing control structure of isobutene process control. This structure process design is to achieve the efficiency in control and reduce disturbances in process by using the new process structure design by Wongsri (2009) compare to the basic heuristic structure design by Luyben (2002). Disturbances are set up by two categories: flow rate of substances and temperature changed in fresh feed. The result shows that control structure designs based on Wongsri heuristic are better performance than Luyben in both sustain in quality and minimize the power efficiency.

Pantita Laklert, *Plantwide control structures design of a methyl amines process*, 2011

The methyl amines process is a complex chemical process with reaction and separation sections connected by several recycle streams. There are three consecutive reversible reactions that produce three different chemical components. In this thesis, the plantwide control strategy is considered for the methyl amines process. Fixture point theorem is applied to the plantwide control structures design for selecting appropriated set of controlled variables. A commercial process simulation software achieves the dynamics simulation of this research, both of steady-state and dynamics conditions.

Safiya Benchavichien, *Plantwide control structure design for an auto-refrigerated alkylation process*, 2011

The alkylation process is widely used in oil refinery as a process to produce an alkylate product which is very useful for internal combustion engines. The process operated by reacting isobutene with olefins (primarily mixture of butane-butylene) and adding the sulfuric acid catalyst. The process is carried out in exothermic reactions in a series of agitated reactors. The separation sections with two distillations are then installed to extract the desirable products and return the useful remnant back to the recycled stream. This research has been developed by using plantwide control procedure of Wingsri (2012) to improve the control structure of an auto-refrigerated alkylation process which can be referred as multi-unit process containing several unit operations. The procedure used heuristics method to find the fixture plant which is appropriate in handing material and heat disturbances entering the process. Wherefore, a commercial dynamics simulator is used to design and simulate the alkylation process at steady state and dynamic conditions.

Saowarat Thongkam, *Plantwide control structure design for acetone process via dehydrogenation of 2- propanal*, 2011

Designed a control structure of the acetone process by using the heuristics procedure of Wongsri (2012). The procedure takes the fixture plant into consideration to provide a reasonable control structure which is straightforward to understand. The control structure is then designed at steady state and dynamic conditions through a commercial process dynamics simulator. Consequently, the control structure is well handled with both material and thermal disturbances.

Nattaphol Srithong, *Plantwide control design of biodiesel production process with alkali-catalyst*, 2009

This study is using plantwide control strategies for designing the control structures of an alkali-catalyzed process to produce biodiesel from palm oil that are designed to achieve the control objective. Plantwide control strategies using Luyben's heuristics method and Fixture point of Wongsri were adapted for developed control loops in the biodiesel production plant. The control structure for biodiesel production plant showed purity of biodiesel and glycerol by-product greater than ASTM.

Saowani Detjareansri, *Plantwide control structures design for alkylation process*, 2009

The research used plantwide control procedure of Wongsri (2009) to develop the control structures for alkylation process. Eight plantwide control structures (CS1 to CS8) for alkylation process was dedigned by Wongsri (2009) procedure and evaluate the dynamic performance of the designed control structures compare with the base case control structure (Luyben, 2002) 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 integral absolute error (IAE) and total energy use low. Therefore, this research establishes that the Wongsri's procedure, which combines heuristics, analytical method and dynamic simulation, a useful design procedure that leads to a good-performance plantwide control system.

Chaiyapop Siraworakun, *Synthesis of the plantwide control structure*, 2008

The combined mathematic and heuristic based approach is proposed for establishing the plantwide control structure in this dissertation. The proposed approach takes advantages of both heuristic based and mathematic based approaches to develop the appropriate plantwide control structures. In heuristic part, sets of controlled variables and manipulated variables are selected followed by establishing of the obvious control loops. In mathematic part, the dynamic performance-based optimization is proposed for establishing the plantwide control structures. The optimization problem is formulated as a mixed integer nonlinear programming (MINLP). The objective function is presented as an integral of the time-weighted absolute error (ITAE) of all measurements and manipulated variables in the face of disturbances. A discrete state-space model is used as the process model in

the optimization problem. The proposed approach is investigated on the Tennessee Eastman (TE) process. The obtained plantwide control structures are compared with the earlier work given by Luyben et al. (1999). In dynamic simulation, the performance of the plantwide control structures is evaluated in the face of disturbances and setpoint changing. The simulation results show that the obtained plantwide control structures give the appropriate responses compares with those of Luyben et al. (1999).

Kasin Pronpitakthum, *Design of heat-integrated process structures for HDA plant*, 2008

In this research, modification of the separation section of HDA process and the use of heat exchanger network (HEN) to save energy are proposed. The traditional HDA distillation train is replaced by a new design. Furthermore, four new heat exchanger networks (HENs) are developed. The plantwide control structures are designed using the disturbance load propagation method (Wongsri, M., 1990) and heat pathway heuristics (Wongsri, M. and Hermawan Y.D., 2005), respectively. Two kinds of disturbances: thermal and material disturbances are used in evaluation of the plantwide control structures. The performances of the heat integrated plants (HIPs) and the control structures evaluated dynamically by commercial software HYSYS.

Suchada Suntisrikomol, *Plantwide control structures design procedure applied to the hydrodealkylation process using fixture point theorem*, 2008

The thesis present the “Fixture point theorem” to select appropriate the controlled variables from a large number of candidate output. The fixture point control theorem states that the most disturbed points must be satisfactorily controlled by giving them consideration before other controlled variables. The maximum (scaled) gain is used to select and paring controlled variables with manipulated variables. In this study, the set of the first rank of controlled variables is the same as Luyben (1998). We selected three sets of controlled variables (second and third rank from fixture point) and five control structures were designed and compared. In order to illustrate the dynamic behaviors of the control structures when economic disturbance load occur, such as change in methane composition in fresh feed gas and quencher outlet temperature. The performance of designed control structures were presented in IAE value and compared with reference structure. The designed structures are fast response and the most effective on

compared with reference structure 1 (Araujo et al, 2006) and same reference structure 2 (Luyben, 1998)

Teerapan rujirachun, *Plantwide control structure design for an acid-catalyzed process to produce biodiesel from used palm oil*, 2008

The study is to use plantwide control strategies for designing a control structure of an acid-catalyzed process to produce biodiesel from used palm oil. Three control structures are proposed, and their performances for withstanding disturbances that cause production rate change are evaluated. The disturbances consisting of the amount of methanol and feed temperature before entering the reactor are introduced. The first control structure uses a product flow rate to control the quality of products. The result shows that the product purity is quite steady but the product flow rate is fluctuated. The second control structure measures total methanol flow rate in the process, and adjusts the fresh methanol feed rate accordingly. This structure shows a faster dynamic response than that of the first control structure. Moreover, the heat load used to handle disturbances is the lowest. The product purity and flow rate of this structure are more fluctuated than the first control structure. In the third control structure, a cascade control is introduced to the second control structure for controlling the product purity. The product flow rate control is also added. This control structure has dynamic response similar to those obtained by the second control structure. The second control structure is the best control structure to handle disturbances due to it gives better control performances and lower heat load required.

Chaiwat Chuliwanlee, *Design of control configuration for highly heat-integrated HAD process*, 2007

The design plantwide control structures for a highly heat integrated plant is quite difficult task since the highly heat integrated plant has a few utility unit (i.e. heater and cooler) to absorb the thermal disturbance load. This problem can be solved by adding auxiliary utility unit. However, more auxiliary utility unit will be increase capital operating and maintenance costs. In this research, we propose the strategy to design the workable highly heat integrated plant like alternatives 5 and 6 of hydrodealkylation of toluene (HDA) process with minimum auxiliary utility unit. It starts with specifying the disturbances and their magnitudes, next designing the worst case condition, and then designing the heat pathway. Finally, the minimum auxiliary utility units are evaluated. We can solve the control difficulties associated with

alternatives 5 and 6 by adding an auxiliary utility unit to the process instead of three and four as suggested by Luyben (1999), respectively. The three new control structures are proposed and their performances are evaluated. As shown in dynamic simulation study, the control performance for the highly heat integrated plant with minimum auxiliary utility unit is same with the highly heat integrated plant with full auxiliary utility units. CS1 is the best control structure for handle disturbances due to it gives better control performances. In this control structure, the selective controller with low selector switch (LSS) is employed to achieve dynamic maximum energy recovery that can save energy about 0.45% for change in the heat load disturbance of cold steam. Besides, the inlet hot temperature at entrance of reboiler is maintained to prevent the propagation of thermal disturbance to separation section.

Busara Kunajitpimol, *Design of heat exchanger networks and plantwide control structure of butane isomerization plant***,** 2006

In this work, the resilient heat exchanger networks to achieve dynamic maximum energy recovery and plantwide control structures and strategies are designed for Butane Isomerization plant. The control difficulties associated with heat integration are solved by adding auxiliary utilities which is kept minimal. Two alternatives of heat exchanger networks (HEN) designs of the Butane Iso-merization plant are proposed. Both use the heat from the reactor effluent stream to provide the heat for the column reboiler. The energy saved is 24.88% from the design without heat integration, but the capital cost raised is about 0.67% due to adding of a process to process exchanger and an auxiliary utility exchanger to the process. Four plantwide control configurations of heat-integrated plant are designed following Luyben's heuristic method. The result shows the fourth control structure can reject disturbances better than other control structures. In general, the control systems for CS1 to CS4 in the butane isomerization process alternative 1 are better than that in alternative 2. However, the control systems for CS1 to CS4 in the butane isomerization process without energy integration are the most effective one compared with the other two alternaives. Various heat pathways throughout the network designed using Wongsri's disturbance propagation method to achieve DMER. The designed control structure is evaluated based on the rigorous dynamic simulation using the commercial software HYSYS.

Piyaporn Poothanakul, *Plantwide control design for a butane isomerization process*, 2002

In general, a chemical plant is composed of two sections: reaction section and separation section. Both sections need a control system so that the plant can operate economically and safely. Plantwide control strategies was adapted for developing control loops in the plant since its approach is holistic which could provide a better control system. This thesis paper applied plantwide control strategies for designing control structures of a butane isomerization process to achieve impurity of normal butane in product not more than two mole percent and also achieve the desired production rate. Three control structures were proposed, tested and compared with the control structure based on plantwide process control book, Luyben 1998.

Sayfon Kietawarin, *Control structure design applied to hydrodealkylation process plantwide control problem*, 2002

The thesis presents a comparison among 4 control structures designed for withstanding disturbances that cause production rate change. In the study, the changes have been introduced to the amount of toluene and feed temperature before entering the reactor. Compared with the reference control structure using a level control to control toluene quantity in the system, the first control scheme measured toluene flow rate in the process and adjusted the fresh toluene feed rate accordingly. This structure resulted in faster dynamic response than the reference structure. The second control scheme was modified from the first scheme by adding a cooling unit to control the outlet temperature from the reactor, instead of using internal process flow. The result is to reduce material and separation ratio fluctuations within the process. The product purity is also quite steadily. In the third control scheme, a ratio control was introduced to the second control scheme for controlling the ratio of hydrogen and toluene within the process. This scheme showed that it can withstand large disturbances. Dynamic study shows that the control structure has significant effect on process behavior. A good control system should quickly respond to disturbances and adjust itself to steady state while minimizing the deviation of the product quality.

VITA

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