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นายกิตต์ กิจนิธิกุล

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PLANTWIDE CONTROL STRUCTURE DESIGN FOR BUTANE ISOMERIZATION PROCESS

Mr. Kitt Kijnithikul

A Thesis Submitted in Partial Fulfillment of the Requirements
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By	Mr. Kitt Kijnithikul
Field of Study	Chemical Engineering
Thesis Advisor	Assistant Professor Montree Wongsri, D.Sc.

Accepted by the Faculty of Engineering, Chulalongkorn University in Partial Fulfillment of the Requirements for the Master's Degree

....Dean of the Faculty of Engineering
(Professor Bundhit Eua-arporn, Ph.D.)

THESIS COMMITTEE

....Chairman
(Assistant Professor Amornchai Arpornwichanop, D.Eng.)
....Thesis Advisor
(Assistant Professor Montree Wongsri, D.Sc.)
....Examiner
(Chutimon Satirapipathkul, D.Eng.)
....External Examiner
(Chaiyapop Siraworakun, D.Eng.)

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งานนี้จะอธิบายถึงการออกแบบการควบคุมแบบแพลนไวน์ด้วยวิธีวงศ์ศรี 2015 กับกระบวนการผลิตบิวเทนด้วยปฏิกิริยาไอโซเมอร์ไรเซชั่น ความสำคัญของวิธีนี้คือการใช้ตำแหน่งควอนทิไฟเออร์และการควบคุมในระดับโรงงาน เพื่อรักษาเสถียรภาพของการควบคุมการไหลของสาร และการรักษาคุณภาพของผลิตภัณฑ์ ให้สามารถกำจัดตัวรวบกวนทั้งหลายให้ออกไปจากกระบวนการหรือไปในเส้นทางที่ต้องการให้ไป รวมถึงการควบคุมในระดับอุปกรณ์ โดยอุปกรณ์ที่ติดตั้งและโครงสร้างการควบคุมจะถูกเรียกว่าฟิกเจอร์แพลน ซึ่งโครงสร้างใหม่จะถูกนำไปเปรียบเทียบกับการออกแบบของไลเบน

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KITT KIJNITHIKUL: PLANTWIDE CONTROL STRUCTURE DESIGN FOR BUTANE ISOMERIZATION PROCESS. ADVISOR: ASST. PROF. MONTREE WONGSRI, D.Sc., pp.

This work describes the application of the plantwide control design procedure of Wongsri (Wongsri, 2015) and the butane isomerization process. The significance of the procedure is the use of quantifiers and their handlers to design plant level quantifier (inventory) control to stabilize the material flow. Next to maintain the quality of the products, the material and heat disturbance management are used to design for plant level disturbance rejection loops for dynamic efficiency. The column control loops are designed to shift plus and minus disturbance in the desired direction. Then the unit level loop design is carried out. The plant equipped with this designed control structure is called a Fixture Plant. The new design is compared with Luyben's design.

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

Page

THAI ABSTRACT iv

ENGLISH ABSTRACT v

ACKNOWLEDGEMENTS vi

CONTENTS vii

List of Figures 1

List of Tables 1

CHAPTER 1 INTRODUCTION 1

    1.1 Importance and reasons of research 1

    1.2 Research Objectives 2

    1.3 Scopes of research 2

    1.4 Expected Contributions 2

    1.5 Research Procedures 2

    1.6 Research Framework 3

CHAPTER 2 LITERATURE REVIEW 5

    2.1 Review of plantwide control structure design 5

    2.2 Review plantwide control structure design of butane isomerization process 10

CHAPTER 3 Plantwide Control Design Principle 11

    3.1 The plantwide process control 11

    3.2 The main function of control system 13

    3.3 Integrated Process 13

    3.4 Plantwide Control Problems 16

    3.5 Reaction/Separation Section Interaction 18

    3.6 Basic Concepts of Plantwide Control 20

    3.7 Nine steps of Plantwide Process Control Design Procedure of Luyben 22

    3.8 Wongsri’s Plantwide Control Design Procedures 25

CHAPTER 4 BUTANE ISOMERIZATION PROCESS 38

    4.1 Introduction 38

    4.2 Reaction Kinetics 38

Page

4.3 Iso-butane process description....39

CHAPTER 5 CONTROL STRUCTURES DESIGN....43

5.1. Introduction....43

5.2. Control Structure Design Procedures Applied to the Butane Isomerization Process....43

Stage 1: Plant information and analysis....43

Step 1: Gather relevant plant information and control objectives, including constraints for control....43

Step 2: Perform Plant Analysis....46

Stage 2: Plant Level Design: Fixture Plant Design and Disturbance Management....70

Step 3: Establish fixture plant....70

Step 4: Handling the disturbances....75

Stage 3. Unit Level Design....100

Step 5: Design the rest of the control loop....100

Stage 4. Enhanced control design....101

Step 6: Adding enhanced control....101

Stage 5. Energy Management and Optimization....102

Step 7: Save energy via heat exchanger networks....102

Step 8: Optimize for economics and/or improving control performance....102

Validate the designed control structures by rigorous dynamic simulation....109

5.3 Dynamic simulation results....109

5.4 Control Structure Performance Evaluation....125

CHAPTER 6 Conclusion....128

....130

REFERENCES....130

APPENDIX....134

APPENDIX A PROCESS STREAMS AND EQUIPMENTS DATA....135

APPENDIX B CONTROLLER TYPE AND TUNING PARAMETERS....140

Page

B.1 Tuning Controller....140
B.2 Tuning Parameter....140
APPENDIX C. IAE Results Data....148
VITA....154

List of Figures

Figure 3.1 Integrated Processes Flowsheet. 14

Figure 3.2 Units in series (a) Level control in direction of flow 17

Figure 3.3 Conventional Control Structure with Fixed Reactor Holdup 19

Figure 3.4 Control Structure with Variable Reactor Holdup 19

Figure 4.1 Temperature profile of Column C1 40

Figure 4.2 Composition profiles of Column C1 40

Figure 4.3 Temperature profile of Column C2 41

Figure 4.4 Composition profiles of Column C2 41

Figure 4.5 Butane Isomerization process flowsheet. 42

Figure 5.1 Heat pathways of Butane Isomerization Process 48

Figure 5.2 $n\mathrm{C}_4$ pathway and quantifier 49

Figure 5.3 $i\mathrm{C}_4$ pathway and quantifier 50

Figure 5.4 $i\mathrm{C}_5$ (heavy impurity) pathway and quantifier 50

Figure 5.5 Effect of molar flow rate disturbances on composition profile 51

Figure 5.6 Effect of temperature disturbances on composition profile 52

Figure 5.7 Effect of composition $n\mathrm{C}_4$ disturbances on composition profile 52

Figure 5.8 Effect of component $n\mathrm{C}_4$ disturbances on composition profile 53

Figure 5.9 Effect of pressure disturbances on composition profile 53

Figure 5.10 Open loop sensitivity analyses between steady stage temperature deviation and manipulate variable (reboiler duty) in the column C1 55

Figure 5.11 Open loop sensitivity analyses between steady stage temperature deviation and manipulate variable (reflux ratio) in the column C1 55

Figure 5.12 Effect of total fresh feed flow rate disturbances on the steady state temperature deviation profiles of the column C1 56

Figure 5.13 Effect of fresh feed temperature disturbances on the steady state temperature deviation profiles of the column C1 57

Figure 5.14 Effect of fresh feed light key composition disturbances on the steady state temperature deviation profiles of the column C1 57

Figure 5.15 Effect of fresh feed light key component flow rate disturbances on the steady state temperature deviation profiles of the column C1 58

Figure 5.16 Effect of fresh feed heavy key component flow rate disturbances on the steady state temperature deviation profiles of the column C1 58

Figure 5.17 Open loop sensitivity analyses between steady stage temperature deviation and manipulate variable (reboiler duty) in the column C2 61

Figure 5.18 Open loop sensitivity analyses between steady stage temperature deviation and manipulate variable (reflux ratio) in the column C1 62

Figure 5.19 Effect of total fresh feed flow rate disturbances on the steady state temperature deviation profiles of the column C2 62

Figure 5.20 Effect of fresh feed temperature disturbances on the steady state temperature deviation profiles of the column C2 63

Figure 5.21 Effect of fresh feed light key composition disturbances on the steady state temperature deviation profiles of the column C2 63

Figure 5.22 Effect of fresh feed light key component flow rate disturbances on the steady state temperature deviation profiles of the column C2 64

Figure 5.23 Effect of fresh feed heavy key component flow rate disturbances on the steady state temperature deviation profiles of the column C2 65

Figure 5.24 Flowsheet of butane isomerization process to use mode of operation on-supply mode 68

Figure 5.25 Flowsheet of butane isomerization process to use mode of operation on-demand mode 69

Figure 5.26 Flowsheet of butane isomerization process to use mode of operation on-internal mode 69

Figure 5.27 Flowsheet of butane isomerization process to keep the raw materials entered and reentered fixed 71

Figure 5.28 Flowsheet of butane isomerization process to keep combine feed.......... 72

Figure 5.29 Flowsheet of butane isomerization process for adjusts the flow of exit material streams according to their accumulation. 73

Figure 5.30 Flowsheet of butane isomerization process for regulate the production rate.. 75

Figure 5.31 Flowsheet of butane isomerization process for handle the heat disturbances.. 76

Figure 5.32 The control structure for handle disturbances in reactor section.............78
Figure 5.33 The control structure for handle disturbances in temperature section.....80
Figure 5.34 Flowsheet of butane isomerization process for handle the material disturbances..99
Figure 5.35 Flowsheet of butane isomerization process for the rest of the control variables. 101
Figure 5.36 Flowsheet of butane isomerization process for energy management. ...102
Figure 5.37 butane isomerization plantwide control structure BC 103
Figure 5.38 butane isomerization plantwide control structure CS1..........................105
Figure 5.39 butane isomerization plantwide control structure CS2..........................107
Figure A.1 Butane isomerization process flow diagram for steady state simulation139
Figure B.1 Input and Output from Relay-Feedback Test. 142

List of Tables

Table 2.1 Previous thesis using Wongsri's procedure 9

Table 3.1 Wongsri's plantwide control design procedure 27

Table 4.1 Butane isomerization Reaction Kinetics 38

Table 5.1 Condition for the butane isomerization process 44

Table 5.2 Steady-State Degrees of Freedom for the Butane Isomerization Process .. 46

Table 5.3 Result of material shifting test for C1 60

Table 5.4 Result of material shifting test for C2 66

Table 5.5 Locate the quantifiers of the components 74

Table 5.6 The dynamic responses for total feed flow rate disturbances of reactor. ... 83

Table 5.7 The dynamic responses for temperature disturbances of reactor. 84

Table 5.8 The dynamic responses for composition disturbances of reactor. 85

Table 5.9 The dynamic responses for component flow rate disturbances of reactor. 86

Table 5.10 The dynamic responses for fresh feed flow rate changes, plus 259.5 and minus 234.7 kmol/h of column C1. 88

Table 5.11 The dynamic responses for temperature changes, plus 37.21°C and minus 27.21°C of column C1. 89

Table 5.12 The dynamic responses for LK composition changes, plus iC_4 0.2635 and minus iC_4 0.2165 of column C1. 90

Table 5.13 The dynamic responses for LK component flow rate changes, plus iC_4 62.265 and minus iC_4 56.340 kmol/h of column C1 91

Table 5.14 The dynamic responses for HK component flow rate changes, plus nC_4 179 and minus nC_4 161.97 kmol/h of column C1 for CS1. 92

Table 5.15 The dynamic responses for total feed flow rate changes, plus 438.0 and minus 396.2 kmol/h of column C2. 94

Table 5.16 The dynamic responses for temperature changes, plus 37.21°C and minus 27.21°C of column C2. 95

Table 5.17 The dynamic responses for LK composition changes, plus nC_4 0.7941 and minus nC_4 of column C2. 96

Table 5.18 The dynamic responses for LK component flow rate changes, plus nC_4 343.42 and minus nC_4 310.72 kmol/h of column C2. 97

Table 5.19 The dynamic responses for HK component flow rate changes, plus $iC_5$14.537 and minus $iC_5$13.153 kmol/h of column C2. 98

Table 5.20 Remaining control variable 100

Table 5.21 Control structure lists of BC 104

Table 5.22 Control structure lists of CS1 106

Table 5.23 Control structure lists of CS2 108

Table 5.24 Setpoint changes and disturbances 109

Table 5.25 Dynamic test results for TPM flowrate changes. 113

Table 5.26 Dynamic test results for increase in iC_5 composition in fresh feed. 117

Table 5.27 Dynamic test results for nC_4 composition changes in fresh feed (±0.069). 120

Table 5.28 Dynamic test results for decrease in catalyst activity. 123

Table 5.29 IAE position for handle disturbance 126

Table 5.30 Summation of the IAE results for handle disturbance to the change in TPM flowrate. 126

Table 5.31 Summation of the IAE results for handle disturbance to the change in feed heavy impurity. 126

Table 5.32 Summation of the IAE results for handle disturbance to the change in reactant feed composition. 127

Table 5.33 Summation of the IAE results for handle disturbance to the change in catalyst activity. 127

Table A.1 Equipment data of butane isomerization process 135

Table A.2 Duty table of steady state operating for butane isomerization process 136

Table A.3 Stream table for steady state operating of butane isomerization process 137

Table B.1 Formulas for the controller parameters in the Ziegler-Nichols method 143

Table B.2 Formulas for the controller parameters in the Tyerus-Luyben method 144

Table B.3 Parameter tuning for Butane isomerization process of BC 145

Table B.4 Parameter tuning for Butane isomerization process of CS1 146

Table B.5 Parameter tuning for Butane isomerization process of CS2 147

Table C.1 The IAE of temperature Results for handle disturbance to the change in TPM flowrate 148

Table C.2 The IAE of pressure Results for handle disturbance to the change in TPM flowrate 149

Table C.3 The IAE of composition Results for handle disturbance to the change in TPM flowrate 149

Table C.4 The IAE of temperature Results for handle disturbance to the change in feed heavy impurity. 150

Table C.5 The IAE of pressure Results for handle disturbance to the change in feed heavy impurity. 150

Table C.6 The IAE of composition Results for handle disturbance to the change in feed heavy impurity. 150

Table C.7 The IAE of temperature Results for handle disturbance to the change in reactant feed composition. 151

Table C.8 The IAE of pressure Results for handle disturbance to the change in reactant feed composition. 151

Table C.9 The IAE of composition Results for handle disturbance to the change in reactant feed composition. 152

Table C.10 The IAE of temperature Results for handle disturbance to the change in catalyst activity. 152

Table C.11 The IAE of pressure Results for handle disturbance to the change in catalyst activity. 152

Table C.12 The IAE of composition Results for handle disturbance to the change in catalyst activity. 153

CHAPTER 1
INTRODUCTION

1.1 Importance and reasons of research

Formally, launching a plant generally took analyzing and designing skill with only a simple control system, from single loops to single processing unit which includes multiple loops. Nevertheless, most industrial applications involve larger problems with multiple processing units and complex interactions. To solve those problems, there is a subject of plantwide control dealing with unit-to-unit interactions through the choice of measured and manipulated variables in each unit and a selection of the control strategy.

Because of a consequence of unit-to-unit interactions, designing the control system for any new plant can be quite difficult. Thus, the important ways to achieve effective plant operations; are an understanding of the potential source of these interactions and finding ways in which they can be significantly diminished. To minimize capital cost and optimize utilization of material and energy are goals of process design. However, most continuous process plants contain numerous units. In order to achieve a goal causes interactions among individual units is even more complex and rather harder to control. Fortunately, the typical plantwide control system consists of many single-loop controllers as well as multivariable controllers can counteract disturbances before they propagate from their sources to other units. That makes the plantwide control system is more efficient too than former simple control systems because it covers consideration - not only on individual units, but among other process factors as a whole.

In the chemical industry, iC4 (i-C_4) is a more valuable raw material than nC4 (n-C_4). It is used in to process of high octane number-gasoline blending components, propylene oxide and tertiary butyl alcohol. The isomerization process is converting the n-C_4 to the more valuable i-C_4. Foremost, in order to obtain the maximum benefit from the process efficiency to produce desired component and to be able to control the overall process to maintain its optimum conditions without being affected by disturbances. Similarly, former executions which have shown a great result; Thus, desired iC4 from the butane isomerization process is investigated applying plantwide control has brought into consideration.

1.2 Research Objectives

To design new plantwide control structures of the butane isomerization process using Wongsri's procedure.

1.3 Scopes of research

1. Steady-state and dynamic simulation of the butane isomerization process is performed by using a commercial process simulator, Hysys.
2. Description and data of the butane isomerization process is given by Luyben (1999).
3. Design procedure is given by Wongsri (2015).

1.4 Expected Contributions

1. Steady state and dynamic models of the butane isomerization process is obtained by using a commercial process simulator.
2. Two new plantwide control structures based on Wongsri (2015).

1.5 Research Procedures

The procedures of this research are as follows:

1. Study of the general theory on plantwide control structure, butane isomerization process and other concerned information.
2. Simulate of the butane isomerization process at steady state condition and dynamic condition.

3. Study about the procedure design plantwide control structures of Wongsri (2015).

4. Design new control structures of the butane isomerization process are as follow Wongsri’s procedures.

5. Simulate a dynamic of the new design plantwide control structures to develop a butane isomerization process.

6. Assess the dynamic performance of the design plantwide control structures.

7. Analyze the simulation results.

8. The research provides conclusions.

1.6 Research Framework

The thesis matter is classified six chapters as follow;

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

Chapter II: presents literature review related to plantwide control structures design procedures, review of previous works on butane isomerization process design, and control structure design.

Chapter III: covers some background information of plantwide and theory concerning with plantwide control fundamentals, new plantwide control structure design procedures. .

Chapter IV: describes process description and the design control structure for the butane isomerization process.

Chapter V: describes the design of plantwide control structures and dynamic simulation results and comparison with control structures of Luyben (1999).

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

CHAPTER 2
LITERATURE REVIEW

2.1 Review of plantwide control structure design

Luyben (1994) presented a numerical case study of a recycle system that exhibited the "snowball" effect. The example process consisted of a continuous stirred-tank reactor followed by two distillation columns. In the simple binary first-order case of A → B, an analytical solution can be found for a recycle flow rate as a function of a fresh feed flow rate and a fresh feed composition. Two other kinetic systems are studied numerically: consecutive first-order reactions A → B → C occurred in the reactor, with component B as desired product. The volatilities of the three components were assumed to be $\alpha_A > \alpha_B > \alpha_C$. Unreacted component A was recycled back to the reactor from the top of the first column. Luyben proposed the generic rule that one flow rate in a liquid recycles loop should be flow controlled. The results of the case study confirm that snowball effects problems can be prevented by using a control structure that fixes the flow rate of one stream somewhere in a liquid recycle loop. In processes with one recycle, the flow rate of the reactor effluent can be set. In processes with two or more recycle streams, a flow rate of each recycle can be fixed.

Luyben (1996) presented one of the central problems in developing a steady-state process flowsheet. It is finding a number of variables that must be specified to completely define a process. This number is called the design degrees of freedom (DOF). The number of design optimization variables can be calculated by subtracting all variables that are set by specifications, on production rate, product qualities, safety constraints, and environmental limitations. For complex industrial processes, there are hundreds of variables and equations, and it is difficult to make sure that the correct variables and equations have been defined. The design of a control structure requires that the control degrees must of freedom be known. It is very easy to calculate a number of variables that can be controlled because it is equal to several manipulated variables. These variables are different than those of design optimization. Several progressively more complex recycles process case studies are used to show these results. The practical significance is that a model is not needed and avoidance of the error procedure of accounting for all variables and equations can be reached.

Luyben, Tyreus & Luyben (1997) proposed the plantwide control structure design procedure by heuristics. The nine-step of this procedure centers around the fundamental principles of plantwide control: energy management, production rate, product quality, operational, environmental and safety constraints, liquid level and gas-pressure inventories, component balances and economic or process optimization. They illustrated this procedure with three industrial examples: the vinyl acetate monomer process, Eastman process and hydrodealkylation (HDA) process. The procedure produced a workable plantwide control structure.

Larsson, Govastmark, Skogestad, and Yu (2002) presented control structure selection for a simple plant with a liquid-phase reactor, a distillation column, and recycle of unreacted reactants. A definition of the operational objectives, constraints, and degrees of freedom is clearly specified. Active constraints should be controlled to optimize the economic performance. The selection of the suitable controlled variable for the remaining unconstrained degree of freedom is the main focus. The concept of self-optimizing control is used which is to search for a constant setpoint strategy with an acceptable economic loss. Both in the case of given feed rate where the energy costs should be minimized and in the case where the production rate should be maximized. A good feed variable is a reflux ratio L/F. This applied to single-loop control as well as multivariable model predictive control.

Skogestad (2004) presented a systematic procedure for control structure design for complete chemical plants. It started with defining the operational and economic objectives, and the degree of freedom available to fulfill them. Other issues include inventory and production rate control, decentralized versus multivariable control, loss in performance by bottom-up design, and a definition of a the “complexity number” for the control system.

Konda, Reangaiah and Krishnaswamy (2005) presented the novel plantwide control (PWC) methodologies are becoming increasingly important as chemical processes are becoming more and more integrated with recycles for reason of safety, environmental considerations, and economics consideration. They proposed an integrated framework of simulation and heuristics. By adopting this framework, simulators can be more efficiently utilized and they also offer invaluable support to the decisions taken by heuristics. The proposed framework is then successfully applied to an industrially relevant case study: the hydrodealkylation of toluene (HDA) process. An analysis of results has shown that the proposed framework built synergies between the powers of both the simulation and the heuristics, thereby resulting in a practical plantwide control structure methodology that leads to a viable control system.

Konda et al. (2006) used simulation-based heuristic approach for designing plantwide control structure. This work is proposed for obtaining both of economic and operational benefits. The approach consists of two stages. The first stage, Alternatives are systematically originated and ranked based on economics. Then, A few top-ranked alternatives are forwarded to the second stage for analysis on their dynamics to define the best process that is economical as well as easy to operate.

Konda et al. (2006) presented the important steps for designing plantwide control through control degrees of freedom. The control degrees of freedom are measured in cases of highly integrated processes. The purpose is for giving a maximum number of flows that be able to manipulate simultaneously in a process, and some (or all) of them can be utilized based on process requirement. The procedure is clearly simpler than the conventional approach.

There are servals thesis works based on Wongsri’s procedure, latest work since 2011 are shown in Table 2.1. Two of them are presented below.

Nuntanoy (2014) presented the economically optimum design of a methanol reactor/ distillation column system with three gas recycle streams to produce high purity methanol form synthesis gas. The new control structure gives better responses in significant plant condition upon changes in synthesis gas feed and CH_4 composition in feed than Luyben’s structure, the plant condition is held much closer to their specifications.

Sikumwong (2014) present a new plantwide control structure design procedure of a styrene monomer process by Wongsri (2015). It’s based on heuristics and process control principles to compare to the designed structure to an Integrated Framework of Simulation, Heuristics and Optimization (IFSHO) proposed by Suraj Vasudevan and G. P. Rangaiah (2011) and SOC procedure proposed by Skogestad (2004). The new procedure has been proven to yield a competitive control structure that performs significantly better in terms of dynamic measures, and still competitive in economic measures.

Table 2.1 Previous thesis using Wongsri's procedure

Year	Title	Authors
2011	Plantwide control structure for methanol process	Kanika Phetyodsri
2011	Plantwide control structure design of styrene process	Keadsuda Machuay
2011	Plantwide control structure design for an auto-refrigerated alkylation process	Safiya Benchavichien
2011	Plantwide control structure design for acetone process via dehydrogenation of 2-propano	Saowarat Thongkam
2012	Plantwide control structure design for ammonia production process	Boonserm Sophonudomsub
2012	Plantwide control structure design of monoisopropylamine process	Chanisa Chumna
2012	Plantwide control structure design of methyl acetate process	Chokesakunt Arrayasinlapathorn
2012	Plantwide control structure design of the methoxy-methyl-heptane process	Kantarakorn Katawetitathum
2012	Plantwide control structure design for butyl acetate process	Oracha Rattananipon
2012	Plantwide control structure design of a methyl amines process	Pantita Laklert
2012	Plantwide control structure design for modified ethyl benzene process	Pira Kanchanawong
2012	Plantwide control structure design of tert-amyl methyl ether (tame) process	Thitima Tapaneeyapong

2013	Plantwide control structure design of isopropyl alcohol dehydration process	Siwapat Tochan
2014	Plantwide control structure design of styrene monomer plant	Boontum Sikumwong
2014	Plantwide control structure design of methanol process	Korakhot Nuntanoy
2014	Plantwide control structure design for cumene process	Saiyawit Korprasert
2014	Plantwide control structure design of hydrodealkylation process	Teerapong Juncharoenwongsa

2.2 Review plantwide control structure design of butane isomerization process

Jagtap, Goenka and Kaistha (2011) developed the economic optimum steady-state design to optimize for a given C_4 fresh feed processing rate (Mode I) and maximum production (Mode II). An effective plantwide control structure is also developed while the design and control of an interesting multiunit process has been explored. The reactor, furnace, heat exchanger, condenser, recycle, two columns process presents challenging problems. The butane isomerization process displayed several unusual features. From the set of active constraints, regulatory plantwide control structures, CS1 and CS2, which minimize the back-off from the economically dominant active constraints, are synthesized along with a simple supervisory optimizing scheme to drive the process operation as close as possible to the active constraints. Comparison with a conventional plantwide control structure, CS3, where the fresh feed is flow controlled, shows that the maximum achievable throughput (profit) for CS2 is higher by ~2%

CHAPTER 3

Plantwide Control Design Principle

Plantwide process control is very important to control a chemical plant with many interconnected unit operations. A typical chemical plant flowsheet has a mixture of multiple units connected that consist of reaction section, separation section and heat exchanger network. So plantwide process control is a complex process control. We must devise the necessary logic and strategies to operate the plant safely and to achieve its design objectives. Our purpose of this chapter is to present plantwide control fundamentals.

3.1 The plantwide process control

A chemical plant may have thousands of measurements and control loops. By the term plantwide control it does not meant the tuning and behavior of each of these loops, but rather it meant the control philosophy of the overall plant with emphasis on the structural decisions. The structural decisions include the selection/placement of manipulators and measurements as well as the decomposition of the overall problem into smaller sub problems. In conclusion, plantwide control that meant control structure design for complete chemical plant.

In an industrial environment, a plant’s control strategy should be simple enough, at least conceptually, so that everyone from the operator to the plant manager can understand how it works. Our governing philosophy is it is always best to utilize the simplest control system that will achieve the desired objectives. The more complex the process, the more desirable it is to have a simple control strategy. This view differs radically from much of the current academic thinking about process control, which suggests that a complex process demands complex control. Our viewpoint is a result of many years of working on practical plant control problems, where it is important to be able to identify whether an operating problem has its source in the process or in the control system.

The design of a plantwide control strategy becomes a more important part of the overall process control design problem. The interrelation of the plantwide control strategy with the process chemistry and economics requires both control theory and also process knowledge. It has become apparent that the design of plant-wide control strategies involves not only the development and application of process control theory but also, in a more fundamental sense, the development of a methodology one uses to approach the plantwide control problem.

The need for a plantwide control perspective arises from three important features of integrated processes: the effects of material recycle, of chemical component inventories and of energy integration. However, we did not describe how we arrived at these strategies, and many of our choices may seem mysterious at this point. Why, for instance, did we choose to use fresh liquid reactant feed streams in the control of liquid inventories? What prompted us to have a reactor composition analyzer? Why were we concerned with a single direct handle to set a production rate?

The subject of plantwide control deals with unit-to-unit interactions through the choice of measured and manipulated variables in each unit and the selection of a control strategy-namely, how to pair controlled and manipulated variables in individual loops, where to use multiple loop controllers, where to use multivariable controllers.

The goals for an effective plantwide process control system include:

1. Safe and smooth process operation

2. Tight control of product quality in the face off disturbance

3. Avoidance of unsafe process conditions

4. A control system runs in automatic, not manual, requiring minimal operator attention

5. Rapid rate and product quality transitions

6. Zero unexpected environmental releases

3.2 The main function of control system

Most industrial processes, the installed control system have main function.

- To reject disturbance

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

- To maintain stability

Every control system must guarantee first the stability. The control system is set to improve the process stability for ensure on product quality and safety to equipment of process and plant.

- To keep the high performance of process

Besides the rejection of disturbance and maintaining stability, the control system can achieve greater profit because it losses less energy and raw materials during the operation. Furthermore the product will meet the desired specification and have production rate.

3.3 Integrated Process

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

- The effect of material recycles.
- The effect of energy integrations.
- The need to account for chemical component inventories.

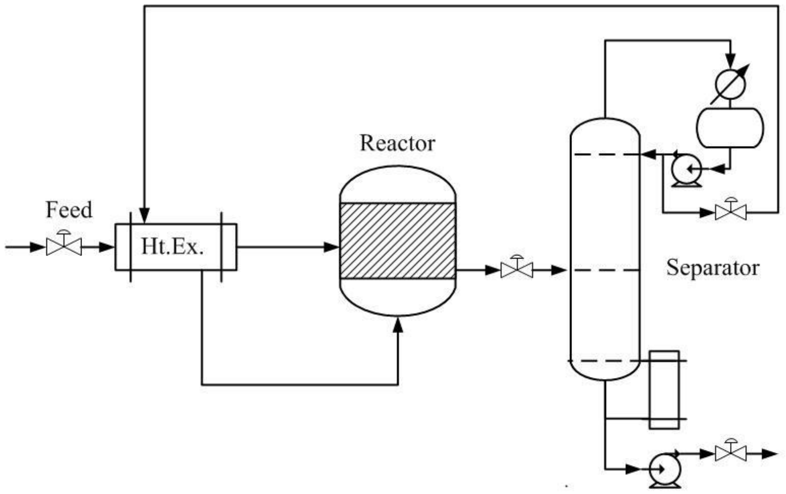


Figure 3.1 Integrated Processes Flowsheet.

3.3.1 Material Recycle

There are six basic and important reasons for material recycled:

1) *Increase Conversion*: If chemical processes involve reversible reactions, conversion of reactants to products is limited by thermodynamic equilibrium constraints. Therefore the reactor effluent must contain both reactants and products. Separation and recycle of reactants are necessary to make the process be economically viable.

2) *Improve Economics*: Generally systems, it is simply cheaper to use a reactor with incomplete conversion and recycle reactants than it is to use of one reactor or multiple reactors in series reaches the desired conversion. A reactor followed by a stripping column with recycled is cheaper than one large reactor or three reactors in series.

3) *Improve Yields*: In reaction systems such as A → B → C, where B is the desired product. Therefore the concentration of B should be kept as low as possible in the reactor and a large recycle of A is required for minimizing the per-pass conversion of A to avoid the formation of the undesirable product C.

4) *Provide Thermal Sink*: In adiabatic reactors with highly exothermic reaction, it is often necessary to feed excess a reactant or one product to the reactor to prevent large amounts of temperature increase. High temperature can create several desired events: it can lead to thermal runaways, catalyst deactivation, undesirable side

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

6) *Control properties*: In many polymerization reactors, the conversion of monomer is limited to achieve the desired polymer properties such as average molecular weight, molecular weight distribution, degree of branching, particle size, viscosity of polymer solutions, 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.

3.3.2 Energy Integration

The energy integration is used to improve the thermodynamic efficiency of the process. In the purpose of reducing the utility cost. Especially for energy-intensive processes, the savings can be quite significant.

3.3.3 Chemical Component Inventories

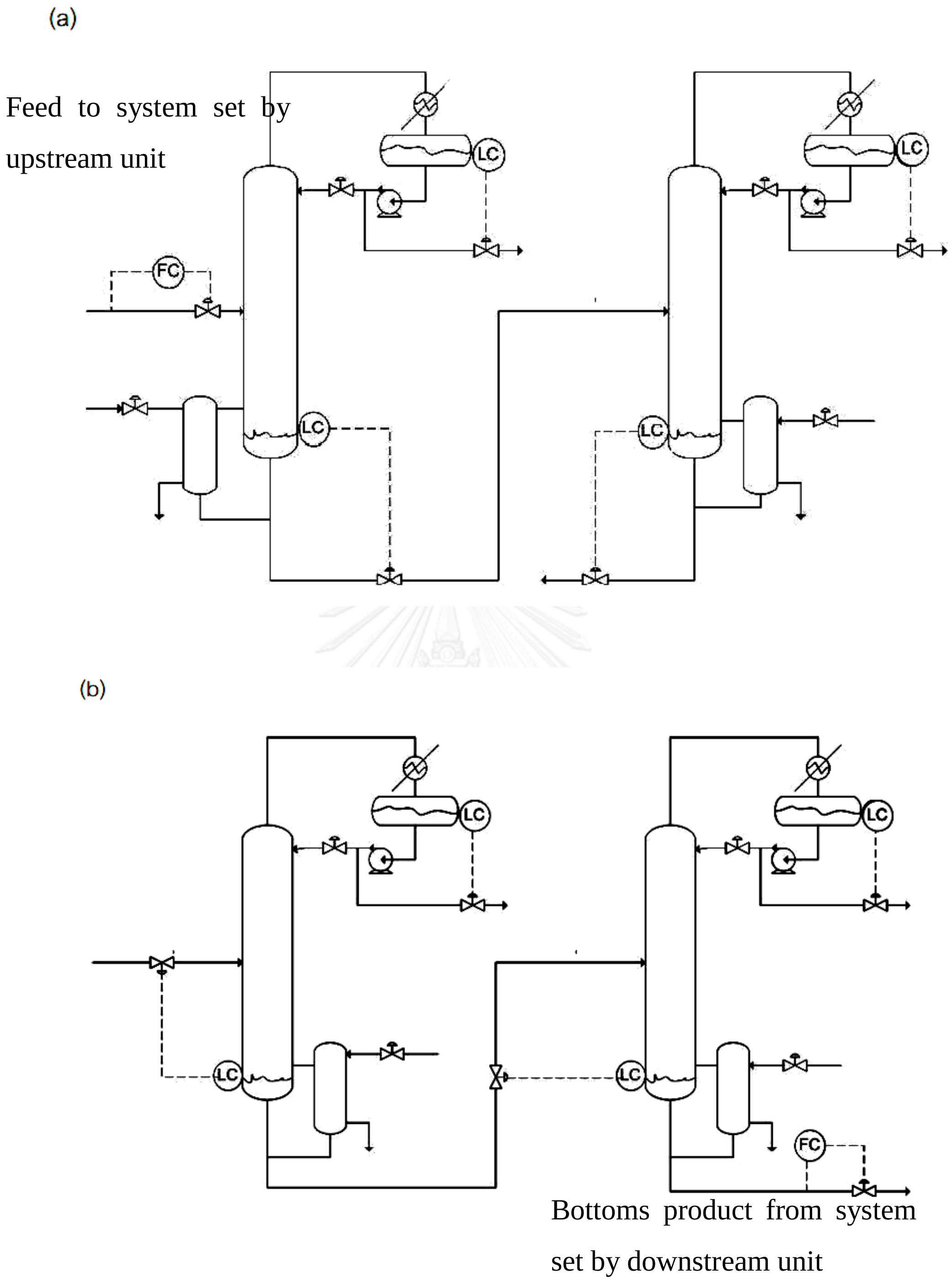
In the chemical process, the chemical component can be characterized into three types: reactants, products, and inerts. A material balance for all components must be satisfied. However, the problems usually found when we consider reactants (because of recycle) accounting for their inventories within the entire process. In order to minimize the quantity of reactants leaving from the process as impurity or purge, we must ensure that every mole of reactant fed to the process is consumed by the reactions, this represents a yield penalty. From the viewpoint of individual units, chemical component balancing is not a problem because exit streams from the units automatically adjust their flows and compositions. However, when we connect units together with recycled streams, the entire system behaves almost like a pure integrator in terms of the reactants. If additional reactant is fed into the system without changing reactor conditions to consume the reactant, this component will build up gradually within the plant because it has no place to leave the system.

3.4 Plantwide Control Problems

3.4.1 Units in Series Problem

If units in entire process are arranged in a series configuration and there is no material and energy recycles, the plantwide control problem is greatly simplified and the issues discussed in the previous section do not have to pay a lot of attention. We can simply configure the control scheme on each individual unit to handle load disturbances.

If production rate is set at the front end of the process, the load disturbances to each unit will come from its upstream neighbor only. If the production rate depends on demand and is set at the terminus, changes in throughput will propagate back through the process. So any individual unit will see load disturbances coming from both its upstream and its downstream neighbors. Figure 3.2 shows these two possible configurations, the production rate is set at the front end and terminus of the process.



Feed to system set by upstream unit

Bottoms product from system set by downstream unit

Figure 3.2 Units in series (a) Level control in direction of flow
(b) Level control in direction opposite flow

3.4.2 Effects of Recycle

We can identify two basic effects of recycle as follows:

(1) Time constants in recycling systems, recycle has an impact on the dynamics of the process. The overall time constant can be much different than the sum of the time constants of individual units.

(2) Recycle leads to the “snowball effect”. This has manifestations, one steady state and one dynamic. A small change in throughput or feed composition can lead to a large change in steady-state recycle stream flow rates. These disturbances can lead to even larger dynamic changes in flow, which propagate around the recycle loop.

Snowball effect is high sensitivity of the recycle flow rates to small disturbances. When feed conditions are not very different, recycle flow rates increase drastically, usually over a considerable period of time. Often the equipment cannot handle such a large load. It is a steady-state phenomenon, but it does have dynamic implications for disturbance propagation and for inventory control.

The large swings in recycling flow rates are undesirable in plants because they can overload the capacity of separation section or move the separation section into a flow region below its minimum turn-down. Therefore, it is important to select a plantwide control structure that avoids this effect.

3.5 Reaction/Separation Section Interaction

Different control structures produce different behavior in handling disturbances. To understand the interaction between reaction and separation section, the two control structures of the same process are considered as shown in figures 3.3 and 3.4.

In the first shown in figure 3.2, only the separation section must absorb almost all of the changes. To increase the production rate of product which is in the bottom stream of distillation column by 20 percent, the overall reaction rate must increase by 20 percent. Because of both reactor temperature and reactor holdup fixed constant, only the composition in the reactor must increase 20 percent. This is a significant change in the composition of the feed stream to the distillation column. The load on the separation section changes significantly producing large variations in recycling flow rates.

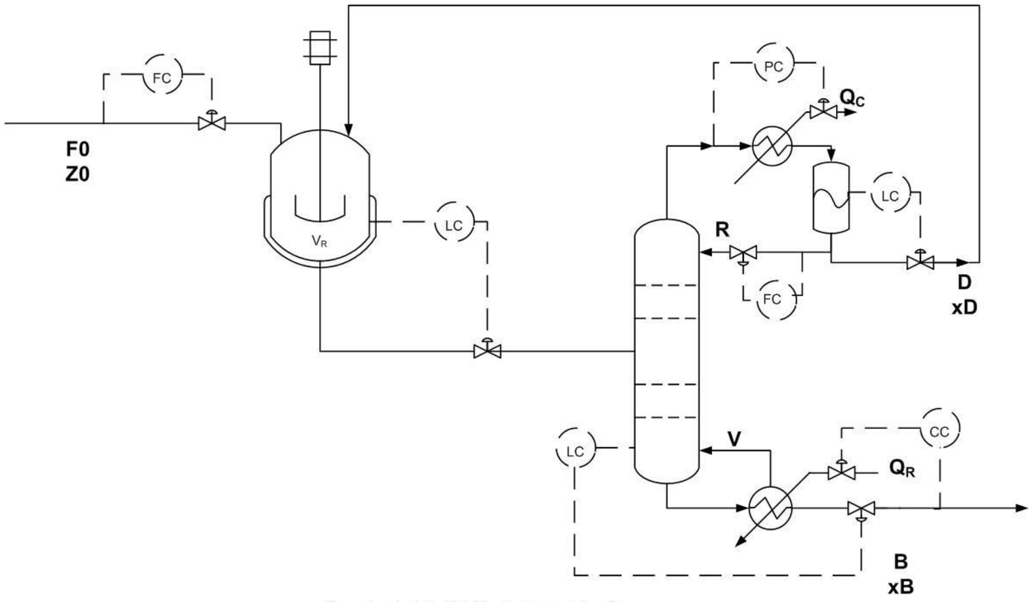


Figure 3.3 Conventional Control Structure with Fixed Reactor Holdup

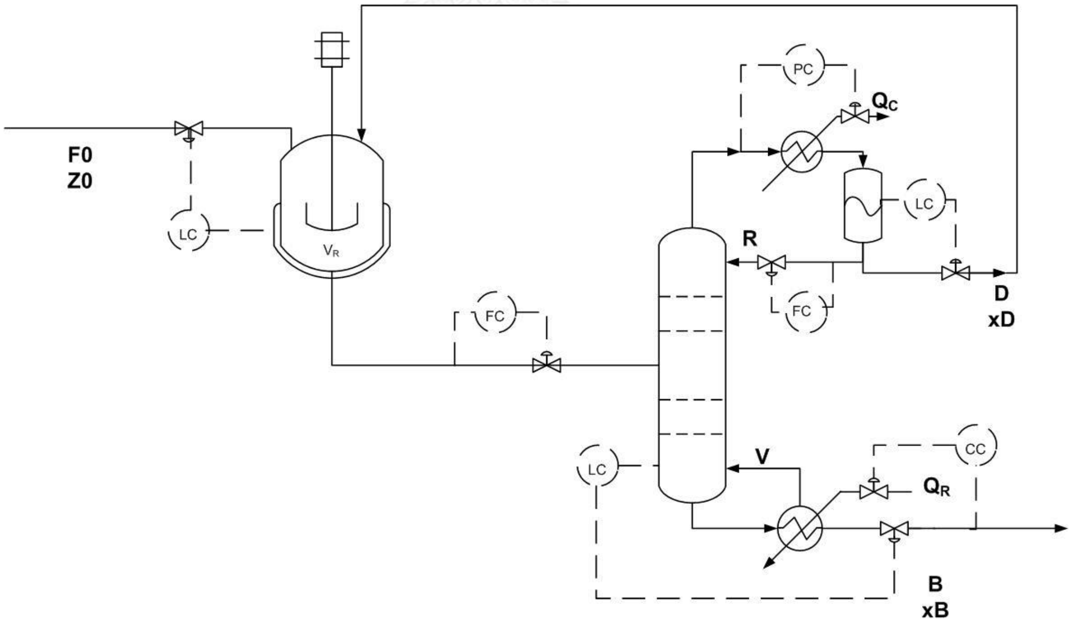


Figure 3.4 Control Structure with Variable Reactor Holdup

In the second structure shown in figure 3.4, both reactor holdup and reactor composition can change, so the separation section sees a smaller load disturbance. The effects of the disturbance can be distributed between the reaction and separation sections which resulting a smaller change in recycling flow.

3.6 Basic Concepts of Plantwide Control

3.6.1 Buckley Basics

Buckley (1964) was the first person suggesting the idea of separating the plantwide control problem into two parts. Those are material balance control and product quality control. The first part focuses on the flow of material throughout the system. This part will choose the proper vessels used to regulate low-frequency disturbances by looking at the flow rates of liquid and gas streams. Then, proper variables will be chosen for a product quality control part. These variables have to be able to efficiently deal with high-frequency disturbances in order that overall stream flows will become smoother. However, this procedure provides little guidance concerning three important aspects of a plantwide control strategy.

1. Energy management is not explained obviously.

2. It does not address the specific issues of recycled systems.

3. It does not deal with component balances in the context of inventory control.

3.6.2 Stephanopoulos

Stephanopoulos (2000) clearly summarized the creativity involved in the control-strategy synthesis for chemical plants. However, there is still little guidance in the literature on designing workable plantwide control systems. We present here such a systematic design procedure. Our framework in tackling a problem of this complexity is based upon heuristics that account for the unique features of plantwide control. The scope embraces continuous processes with reaction and separation sections. The procedure is based upon research by us and others, and our governing philosophy is it is always best to utilize the simplest control system that will achieve the desired objectives. Three industrial processes illustrate its application in this article. In addition, the procedure has been used successfully over several years in a number of commercial processes at DuPont.

3.6.3 Downs drill

Jim Downs (2011) have insightfully pointed out the importance of looking as the chemical component balances around the entire plant and checking to see that the control structure handles these component balances effectively. We must ensure that all components (reactants, products, and inerts) have a way to leave or be consumed within the process. Most of the reactants fed into the process must be chewed up in the reaction because we want to minimize raw material costs and maintaining high purity products. And the stoichiometry must be satisfied down to the last molecule. Chemical plants often act as pure integrator's in terms of reactants due to the fact that we prevent reactants from leaving the process through composition controls.

3.6.4 Luyben laws

Luyben (1997) has proposed a result of a number of case studies of many types of system, Bill Luyben suggests the following:

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

2. For systems with reaction types such as $A + B \longrightarrow$ Products, a fresh reactant feed stream cannot be flow-controlled unless there is essentially complete one-pass conversion of one of the reactants.

3. If the final product comes from a distillation column, the column feed should be:

- Liquid, if the product is in the distillate flow.

- Vapour, if the product is in the bottom flow.

By this arrangement, disturbances in the feed flow rate or the feed composition have a smaller dynamic effect on the product than otherwise.

3.7 Nine steps of Plantwide Process Control Design Procedure of Luyben

The design procedure is carried out in nine steps, which contented the two fundamental of the overall conservation of energy and mass. Additionally, the procedure accounts for nonconserved entities within a plant such as chemical components (produced and consumed) and entropy (produced). Five of the nine steps deal with plantwide control issues that would not be addressed by simply combining the control systems from all of the individual unit operations. Steps 1 establish the control objectives. Step 2 determines the available degrees of freedom. Step 3 establishes energy management within the process and ensures the production of heat (entropy) is properly dissipated. In Steps 4 and 5 the business objectives are concerned production rate, product quality, and safety. Step 6 involves total mass balance control, whereas in Step 7 check component balances, ensures chemical components are accounted because of nonconserved. In Step 8 individual unit operations complete for the control systems. Finally, Step 9 uses the remaining degrees of freedom for optimization and improved dynamic controllability. This heuristic procedure will generate a workable plantwide control strategy, which is not necessarily the best solution. Because the design problem is open-ended, the procedure will not produce a unique solution.

Step 1: Establish Control Objectives.

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

This is probably the most important aspect of the problem because different criteria lead to different control structures. These objectives include reactor and separation yields, product quality specifications, product grades and demand determination, environmental restrictions, and the range of safe operating conditions.

Step 2: Determine Control Degrees of Freedom.

Count the number of control valves available.

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

Step 3: Establish Energy Management System.

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

We use the term “energy management” to describe two functions:

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

2. If heat integration does title for economic and operational reasons. Hence we should select manipulated variables such that the dynamic relation occurs between process streams, then the second function of energy management is to provide a control system that prevents the propagation of thermal disturbances and ensures that the exothermic reactor heat is dissipated and not recycled. Process-to-process heat exchangers and heat-integrated unit operations must be analyzed to determine that there are sufficient degrees of freedom to control.

Step 4: Set Production Rate.

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

Determine what valve will be used to set throughput. Often design constraints require that production be set at a certain point. An upstream process may establish the feed flow sent to the plant. A downstream process may require on-demand production, which fixes the product flow rate from the plant. If no constraint applies, then we select the valve that provides smooth and stable production rate transitions and rejects disturbances.

Step 5: Control Product Quality and Handle Safety, Operational, and Environmental Constraints.

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

We want tight control of these important quantities for economic and operational reasons. Hence we should select manipulated variables such that the dynamic relationships between the controlled and manipulated variables feature small time constants and dead times a large steady-state gains. The former gives small closed-loop time constants, and the latter prevents problems with the range ability of the manipulated variable (control-valve saturation). The magnitudes of various flow rates also come into consideration.

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

Fix a flow in every recycle loop and then select the best manipulated variables to control inventories.

Determine the valve to control each inventory variable. These variables include all liquid levels (except for surge volume in certain liquids recycle streams) and gas pressures. In most processes a flow controller should be present in all liquid recycle loops. This is a simple and effective way to prevent potentially large changes in recycling flows that can occur if all flows in the recycle loop are controlled by levels. Two benefits result from this flow control strategy. First, the plant's separation section is not subject to large load disturbances. Second, consideration must be given to alternative fresh reactant makeup control strategies rather than flow control. In a dynamic sense, level controlling all flows in a recycle loop is a case of recycling disturbances and should be avoided.

Step 7: Check Component Balances

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

Ensure that the overall component balances for each chemical species can be satisfied either through reaction or exit streams by accounting for the component's composition or inventory at some point in the process. Component balances can often be quite subtle. They depend upon the specific kinetics and reaction paths in the system. They often affect what variable can be used to set production rate or rate in the reactor.

Step 8: Control Individual Unit Operations.

Establish the control loops necessary to operate each of the individual unit operations. Many effective control schemes have been established over the years for chemical units (Shinskey, 1988). For example, Liquid solvent feed flow to an absorber is controlled as some ratio to the gas feed.

Step 9: Optimize Economics or Improve Dynamic Controllability.

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

After satisfying all of the basic regulatory requirements, we usually have additional degrees of freedom involving control valves that have not been used and set points in some controllers that can be adjusted. These can be used either to optimize steady-state economic process performance (e.g., minimize energy, maximize selectivity) or to improve dynamic response.

3.8 Wongsri's Plantwide Control Design Procedures

Luyben et al. (1997) proposed a nine step heuristic procedure to design the plantwide control structure. This procedure is widely studied and used the plantwide process control community. Although the heuristic methodologies are easier to understand and implement, but they just need the basic understanding of the process along with some experience. Thereby the structural design procedure is given explicitly. The designer's process engineering insight is necessity to pair CVs and MVs. Skogestad (2004) presented the new design procedure based on mathematical analysis. First, the operational objectives and the dynamic and steady state degree of freedom are identified. Then the set of primary controlled variables is selected. They basically are the active constraints and variables that must be maintained to achieve minimal economic loss when disturbances or implement error occur. Then the production rate is set based on the optimization resulted in the previous step. The secondary controlled variables are selected and the pairing of the choose sets of MVs and CVs is done hierarchically: regulatory control, supervisory control and optimization layers. In these steps, several analysis tools are utilized, e.g. pole vector analysis, RGA, CLDG analysis, linear and nonlinear optimization. This procedure does not discuss about the precedence of control variables and how to handle the disturbances.

3.7.1 Fixture Point Theorem

Hagglund (1995) presents the real-time oscillation detection by calculateing the integrated absolute deviation (IAE) between successive zero crossing of controller error signal. Its motivation is an automatic monitoring of control-loop performance. The concept of material and energy disturbance propagation controls lead to fixture point theorem.

The plantwide control structure design procedures are developed by Wongsri via fixture point theory. The heuristic is used to select pairs of controlled and manipulated variables and uses the dynamic simulation to obtain the main control structure.

Fixture point theorem analysis:

1. The process is considered at dynamic mode (we run the process until the process responses are at the steady state).

2. Controlled variable can be arranged to follow the most sensibility of the process variable by step changes of the manipulated variable in an open loop control (change only one MV, the other should be fixed then alternate to another until complete).

3. Study the magnitude of integral absolute error (IAE) of all process variables that deviates from steady state.

4. Select controlled variable (CV) by considering CV that gave the most deviation from steady state (high value score).

3.7.2 New plantwide control structure design procedures of Wongsri (2015)

Wongsri (2015) presented the new plantwide control design procedure carried out in five stages with eight steps. The major steps deal with plant level design, i.e. establishing a fixture plant, meaning structuring the essential control loops to coherently straighten component flows in entire plant and rectify ongoing disturbances. The straightening of component flows is accounted by regulating the amount of the component at their quantifiers. The component quantifiers can be volumes or flows where such components are concentrated or having maximum concentration. The regulation is made using their proper handlers. The disturbances entering or occurring in the process must be directed to the immediate exits using the material and energy disturbance management to avoid the disturbance propagation throughout the plant. Each step is as follows:

Table 3.1 Wongsri’s plantwide control design procedure

Stage 1. Plant Information and Analysis.	
Step 1	Gather relevant plant information and control objectives, including constraints for control.
Step 2	Perform plant analysis.
	2.1 Count control degree of freedom (CDOF). 2.2 Find heat pathways. 2.3 Identify material pathways. 2.4 Locate material quantifiers. 2.5 Assess reaction section. 2.6 Inspect separation section. 2.7 Schematize production rate manipulation and mode of operation.

Stage 2. Plant Level Design: Fixture Plant Design and Disturbance Management.	
Step 3	Establish fixture plant.
	3.1 Keep the materials entered and/or reentered fixed. 3.2 Adjust the flow of exit material streams (products, byproducts, and inert) according to their accumulations. 3.3 Handle the material that's not leaving the process. 3.4 Control the amount of the rest of the component at their quantifiers. 3.5 Maintain the production rate.
Step 4	Manage disturbance for quality control.
	4.1 Manage heat disturbance. 4.2 Manage material disturbances.
Stage 3. Unit Level Design.	
Step 5	Design the rest of the control loop.
Stage 4. Enhanced control design.	
Step 6	Add enhanced controls, e.g. cascade, feed forward controls.
Stage 5. Energy Management and Optimization.	
Step 7	Save energy via heat exchanger networks.
Step 8	Optimize for economics and/or improving control performance.

Stage 1. Plant Information and Analysis.

Step 1: Gather 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: Perform plant analysis.

Several tasks to assist design decision in Step 3 are:

2.1 Count control degree of freedom (CDOF).

The number of controls the degree of freedom is counted physically by the following simple proposed rule:

"*Each single independent stream, physical or virtual, material or energy, must have a handle or one control degree of freedom.*"

2.2 Find 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 Identify material pathways.

A material pathway is a flow path of a component or group of components with a significant amount of its entry or generated points to its exit or end points. Along the pathway, a material quantifier of the representing pathway can be located. Such quantifiers are used to establish the fixed plant.

2.4 Locate material quantifiers.

A material quantifier is a place indicating the significant or large amount of a chemical component (or a group of components) in the plant. Usually, it is a place where such a component as the highest concentration. The material quantifier can be handled quite readily by adjusting their handlers. A handler or a manipulated variable must be chosen according to cause and effect rule considering the responding time and avoiding undesirable effects. In some case, a quantifier may be a flow. It is, but not necessarily, the place with highest gain of component flow to the total flow.

2.5 Assess 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 separators. One can examine which variables affect reactor productivity by simply looking for these variables at kinetic expression. To see the extent of these variables on the productivity, we can perturb a reactor by changing reactor inlet conditions via a series of steady state simulation.

2.6 Inspect separation section.

The proper directions of plus and minus 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 outlet paths of D+ and D- in the separation section are analyzed and then designed in order to shift plus or minus disturbance loads to the desired exits to achieve our plantwide control objectives, i.e. keeping the main process stream conditions fixed avoiding disturbance propagation and recycling, the fixture plant concept. It is also to maintain product quality. The paths of D+ and D- in the separation section must be shifted to the proper exits. In the case that shifting, plus or minus disturbance to the proper exit is not possible due to process stream conditions or separator specification, the product quality is compromised or disturbance recycling is inevitable.

Next, tray location for composition/temperature control is determined. The tray with the largest changes in the temperature from its initial steady state upon the changes in feed conditions, e.g. composition, total flow, temperature, and component flow, is determined. A set of steady state simulation is performed for these perturbations by keeping the reboiler heat duty and reflux flow or reflux ratio or reflux fraction or boil up ratio constants fixed to reflect what are controlled.

2.7 Schematize production rate manipulation and mode of operation.

Minor throughput changes can be achieved by altering reactor condition. The changes are limited by the bounds of reactor temperature, reactant concentration, reactor holdup. For production or throughput rate change by increasing/decreasing feed rate, recycle flow are adjusted proportionally.

The mode of operation is either dictated by a business objective or process structure. The modes of operation are on-supply (fixed feed rate), on-demand (fixed product rate), and on-internal (fixed internal flow rate). The quantifier (inventory) control along the major flow path follows cause and effect law (see 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 (on-internal) or the separation section (on-supply). The on-internal is recommended according to our fixture plant principle (see below). The material flow to the reactor must be fixed to preserve the principle, i.e. keep the materials entered and/or reentered fixed.

Stage 2. Plant Level Design: Fixture Plant Design and Disturbance Management.

This stage is a major design stage. A 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 shortcomings, Konda (2005). There are two operational objectives: (i) the nominal plant material balance must be maintained and (ii) the heat and material disturbances must be rejected to the nearest exits or directed to less significant streams.

Step 3: 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. This step is to design the control loops to regulate the nominal material flows.

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 insignificantly, each recycle stream may be flow-controlled. Otherwise, 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 that the composition cannot be measured, the nominal composition of the combined stream is used to maintain the combined reactant flow.

In the case of changing throughput via the combined stream of make-up feed and recycle, 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 flow of the product is controlled (mode of operation is on-demand), the quantifier of the product is used to regulate the incoming flow, e.g. level of reflux drum is used to control distillation feed flow and so on.

3.3 Handle the material that’s 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.

In handler selection for quantifier (liquid inventory) control, in addition to fast dynamic response and a causal relationship exists between the two, we recommend choosing a handler that provides quantifier control with minimum offset for a step response.

Step 4: Manage disturbance for quality control.

The nominal conditions of process streams are maintained by specifying the disturbance shifting directions. This step is to design the control loops for disturbance rejection. The principles of disturbance management are followings:

4.1 Manage heat disturbance.

The heat disturbance is divided into two categories. Heat disturbance of category 1 (HDC1) is the heat disturbance that does not instantly effect 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 Manage material disturbances (MMD).

The configurations of the control loops are decided based 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 which distillation control structure yields desired material disturbances (D+ and D-) shifting as established in Step 2.6. The determination is performed using a dynamics process simulator to test 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 feed disturbances, such as temperature, flow rate, composition, and component flow rate are generated in the disturbance shifting test.

The dual end control structure, also, can be testified. The norms are DB, LB, LV, DV, etc. In addition, the best candidate of single end and dual end controls can be mixed and matched. For example, if it is found out that RR structure of single end control and LV structure for dual end control are winning competitors, these two structures are combined resulted as a special structure, RRV structure. Instead of using top temperature or composition to adjust the reflux flow, the new breed uses the top information to adjust the reflux ratio.

4.2.1 Direct the material disturbances of byproducts, and inert to the environment via the nearest exit points.

4.2.2 Direct the deficit material disturbances of the desired products to the environment via the nearest exit points as in Step 4.2.1. Direct the surplus material disturbances of the desired products to their normal exits.

4.2.3 MMD rule for a recycle stream: Direct surplus disturbance of an unreacted raw material to a recycle stream and direct deficit disturbance to the environment in order to maintain a make-up at its nominal flowrate.

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 Design.

The judgment of control loop design at this stage is solely based on individual unit operation.

Step 5: Design the rest of the control loop.

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

Stage 4. Enhanced control design.

Additional control degrees of freedom are identified to add an enhanced control loop to fulfill control objective or improve control performance.

Step 6: Add enhanced controls, e.g. cascade, feed forward controls.

Stage 5. Energy Management and Optimization.

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

Step 7: Save energy via heat exchanger networks.

In the case that the exothermic heat of reaction is large enough to be able 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 (Wongsri, 1990). The design of a control system that prevents the propagation of the heat disturbance of Wongsri and Hermawan (2005) is recommended.

Step 8: Optimize for economics and/or improving control performance.

The design and control issue remains an open research area regarding the plantwide control design, so the opportunity to alter the process design is possible.

Design Validation. The validation of the design control structures using rigorous nonlinear simulation is inevitable; whatever may be the design procedure. 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 4

BUTANE ISOMERIZATION PROCESS

4.1 Introduction

Iso-butane (iC_4) is a hydrocarbon. It is an important chemical substance. Most of them are used in order to make high octane gasoline blending components. iC_4 can be produced from nC_4. The iC_4process consists of several unit operations, furnace, reactor, heat exchanger, condenser, two distillation columns, and liquid recycle stream. Fresh feed C_4 and recycled iC_4 rich condensed (68.6 mol%) are fed into the first column. Thus, this process provides a good example of multi-unit complex process, which is a typical of many chemical plants in the industry.

4.2 Reaction Kinetics

In this process, the reaction kinetics using simple power-law models is the following:

$$R = Ae^{E/RT} V_R C_{nC_4}$$

Table 4.1 shows the reactions kinetic, which including activation energy and intrinsic rate constant. Where R is the reaction rate in kmol/m^3-h (consistent with commercial process simulation requirements), k is the intrinsic rate constant, E is reaction activation energy in kJ/kmol, V_R is the volume of the reactor, temperature units are Kelvin and concentration units are molarity (kmol/m^3).

Table 4.1 Butane isomerization Reaction Kinetics

A	4×10^8
E	69780

4.3 Iso-butane process description

Component physical property data for the butane isomerization process are described from Luyben (1999) in a fair amount of detail.

Column C1: The de-isobutanizer (DIB) column has 50 trays. Reflux ratio is 7.8. The column operates at 6.8 atm. The diameter of the column is 4.877 m. The reflux drum temperature is 323 K with the heat duty of the condenser used is 10.72 MW. Heat duty of the reboiler is 11.11 MW and the base temperature is 336 K. The design specification is to keep iC_4 from dropping out the distillate. Since the specified product (iC_4) purity is 93.5 mol%. The reactant (nC_4) leaves from the bottom with, some iC_4 (light key) and iC_5 from fresh feed. Figure 4.1 and 4.2 gives the temperature and composition profiles in DIB Column, respectively.

Fresh feed C_4 is 263.1 kmol/h (305 K) fed on stage 30 below recycle iC_4 rich condensed stream fed on stage 20 (394 kmol/h) fed to the DIB for separate iC_4 249.5 kmol/h as the distillate with C_3 impurity. The bottom is nC_4 component 407.6 kmol/h and iC_5 from the fresh feed. This bottom stream is fed to purge column to eliminate iC_5 (13.61 kmol/h) with nC_4 rich distillate (394 kmol/h).

Column C2: The bottom of the first column is fed to a second distillation column that produces nC_4 distillate and C_5 bottoms. The purge column has 20 trays and is fed on stage 11. Reflux ratio is 0.8. The column operates at 4.4 atm. The diameter of the column is 1.829 m. The reflux drum temperature is 316 K which heat duty of the condenser used is 3.88 MW. The heat duty of the reboiler is 3.56 MW and the base temperature is 352 K. The separation required of this column is the removal of C_5 from nC_4. The distillate of column purity is 76.2 mol%. The bottom of the column is purged of the process. Figure 4.3 and 4.4 gives the temperature and composition profiles in purge Column, respectively.

Reaction section: The distillate from purge column is preheated by the hot stream from reactor in a feed effluent heat exchanger (FEHE). The cold stream outlet temperature is 405 K to feed in the furnace. The hot nC_4 stream enters an adiabatic packed bed reactor where nC_4 isomerizes irreversibly to iC_4. A reactor size contains

19.6 m^3 (2.134 m. in diameter, 5.486 m. in length). The inlet temperature is set at 472 K by furnace and high pressure 45.25 atm. The hot reactor outlet after losing heat in FEHE is cooled and condensed in a flood condenser. The *i*C_4 rich condensed stream is recycle fed to the DIB above the fresh feed for recovering the *i*C_4.

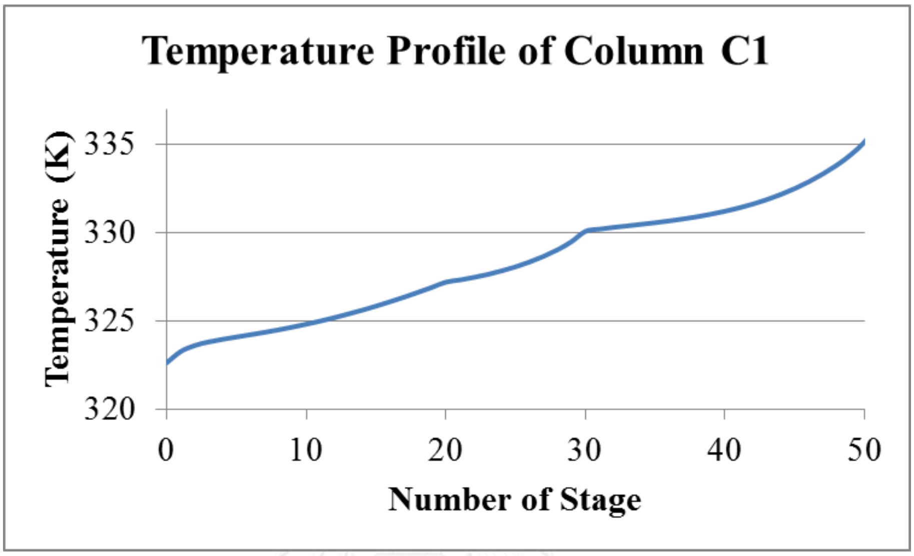


Figure 4.1 Temperature profile of Column C1

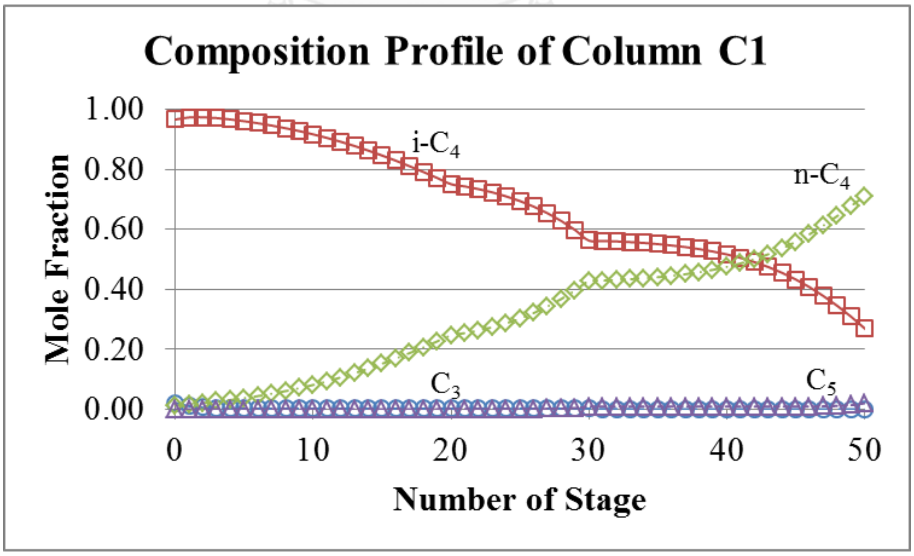


Figure 4.2 Composition profiles of Column C1

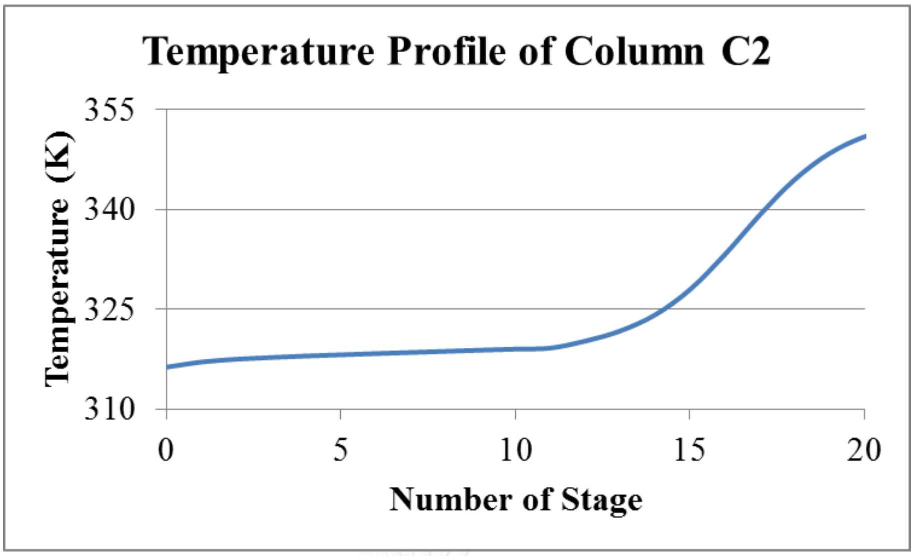


Figure 4.3 Temperature profile of Column C2

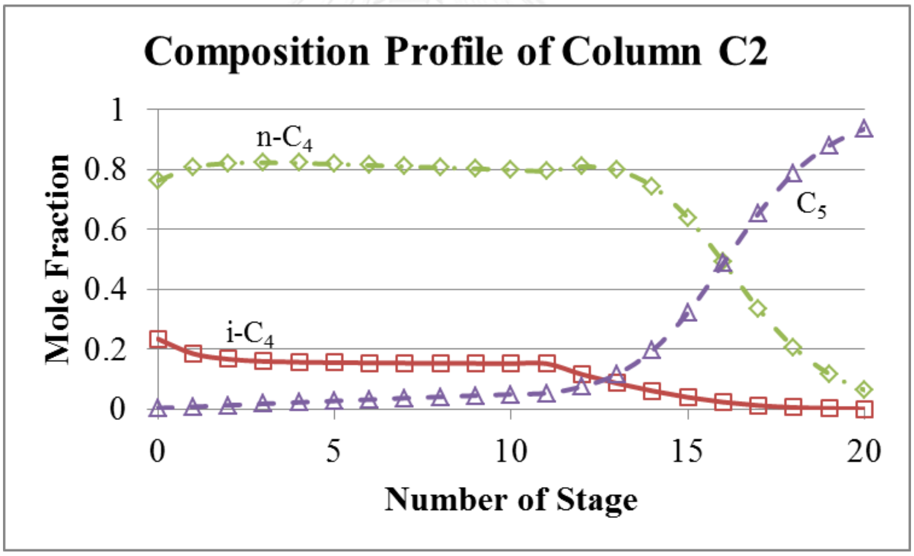


Figure 4.4 Composition profiles of Column C2

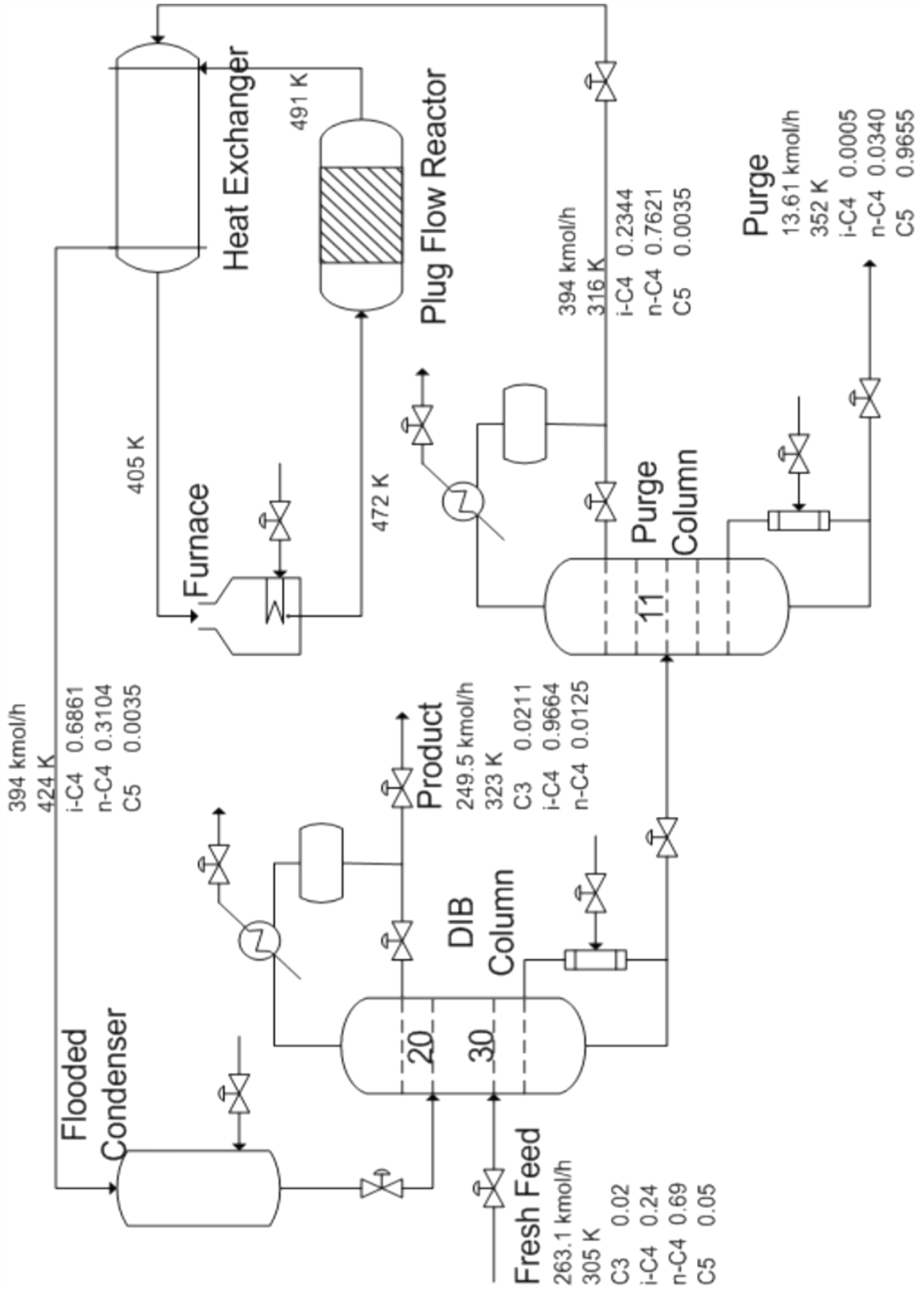


Figure 4.5 Butane Isomerization process flowsheet.

CHAPTER 5

CONTROL STRUCTURES DESIGN

5.1. Introduction

The issues of plantwide control structure design for the Butane Isomerization process are described based on the new design procedure given by Wongsri (2015). The eight major steps in plantwide control structure design are discussed in detail each step. The plantwide control design procedure presented here was developed after many years of work and research in the field of process control and process design. Plantwide control structure design procedure of Wongsri (2015) satisfies the two fundamental chemical engineering principles of the overall conservation of energy and mass. The lists below are the fundamental concepts and techniques of Wongsri (2015) procedure.

5.2. Control Structure Design Procedures Applied to the Butane Isomerization Process.

In this section, control structure design procedure is applied to the Butane Isomerization process. Plantwide control system is developed and tested for the innovated flowsheet. Wongsri's procedure is used for design, which illustrated in the previous section. The design of a plantwide control system consists of eight major steps as follows:

Stage 1: Plant information and analysis.

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

In this step, two categories of plant information must be provided are

1. Plantwide control objectives.

2. Operating Condition.

The plantwide objective includes product quality and production rates. Normally, the plantwide control objectives depend on customer requirements, expected sales and operating costs. In part of the process constraints, stabilizations of

The process is considered. Identify process constraints that must be satisfied, which include safety requirements, environmental constraints and operating conditions. For the butane isomerization process, the objectives are the same as those listed in Table 5.1.

Plantwide objectives for the Butane Isomerization process as follows:

1. Product quality: The product should contain approximately 93.5 %mol of iC_4 product.

2. Production capacity: production rate about 97,200 tons/year of iC_4.

Table 5.1 Condition for the butane isomerization process

Condition	**Temperature (C)**	**Pressure (atm)**
Reactor	198.9	44.4
Column C1	50	6.4
Column C2	42.8	4.35

From lists of all process constraints, feed stream of natural gas separation process is a mixture of nC_4 and excess iC_4 which is too high content for immediate fed into the process. This results as a flowsheet to start with a feeding stream into distillation column C1, in order to separate that excess product and increase reactor efficiency with the concentrated reactant stream. The cause of moderate production quality of 93.5% is due to feed stream impurities; small amount of propane which is not economically worth to separate. Consequently, product stream is shown with small amount of propane presence as well as not over 0.4 %mol of nC_4 of reactant left.

Step 2: Perform Plant Analysis.

2.1 Count control degree of freedom (CDOF).

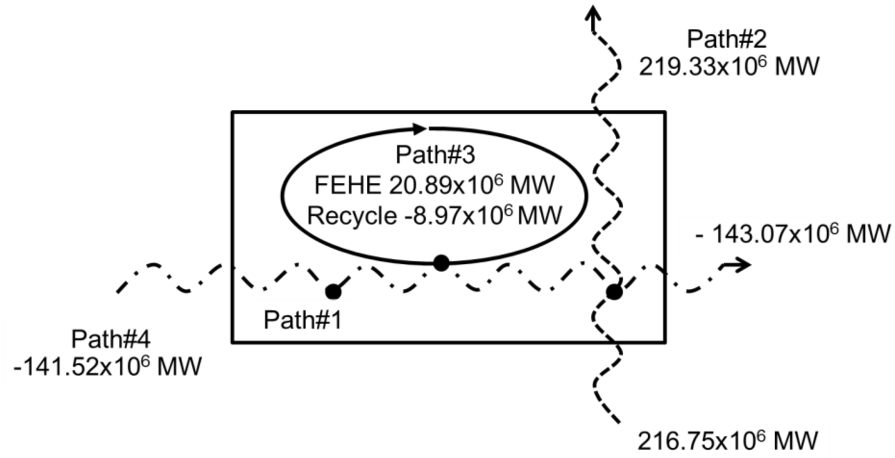
The CDOF can be obtained using the guideline given in Table 5.2. The number of control degrees of freedom equal to the number of manipulating variables. Recall that a controlled degree of freedom is allocated each time a manipulated variable is utilized in a control loop, except in cascade control or in other applications where a setpoint is manipulated instead of a control valve. The steady-state and dynamic CDOF need to be identified. In the butane isomerization process, as mentioned earlier, the CDOF is determined to be 14.

Table 5.2 Steady-State Degrees of Freedom for the Butane Isomerization Process

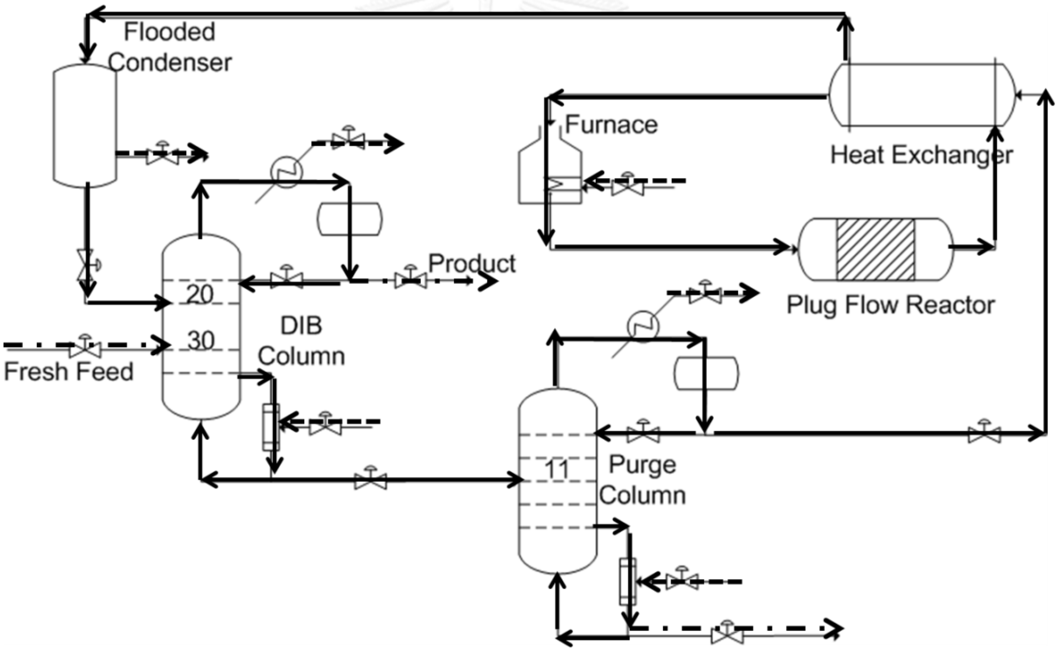
Unit	**Manipulated variable**	**Quantity**	**DOF**
Independent streams	Flow rate	1	1
Furnace	Heat flow	1	1
Condenser	Cooling flow and Flow rate	1	2
Distillation columns	Distillate flow rate, Bottom flow rate, Reflux flow rate Reboiler heat input Condenser heat removal	2	10
Total degrees of freedom			14

2.2 Find heat pathways.

From the plantwide overall image, four different "heat pathways" are introduced in the process. The first pathway is concentrated on the heat generated by exothermic reaction dissipates to the environment. However, the reactor is adiabatic, so the first pathway is not existed in this process. The second pathway is focused on heat from utilities exchanged into the process (216.75×10^6 MW) and to the environment (219.33×10^6 MW). The third pathway is internal heat generated within the process (FEHE 20.89×10^6 MW and Recycle -8.97×10^6 MW). Heat would circulate within the process and between unit operations such as recycle stream. The desired magnitude of energy depends upon the heating and cooling and the amount of heat integration implemented. Whenever the process requires heat, the heat has to be supplied from the utilities. The same amount of heat must be rejected to the environment, nearest exit point in the process or the utilities. The fourth pathway is the enthalpies entered (-141.52×10^6 MW) and left the plant (-143.52×10^6 MW) via process streams.



(a)



(b)

Figure 5.1 Heat pathways of Butane Isomerization Process

2.3 Identify material pathways and 2.4 Locate material quantifiers.

nC_4 is raw material fed to the process, which affects to reaction and iC_4 product quality. Mixed butane (69 %mol nC_4) fed into the first column along with mixed butane (69 %mol iC_4) recycle stream. The first column, iC_4 in distillate has been separated from nC_4 at the bottom. The light impurity is mixed with product flow in the distillate. Since iC_4 is light key and nC_4 is heavy key component. The bottom of the first column is fed to the second column, a heavy impurity (iC_5) in bottom separated from reactant in distillate. Since nC_4 is light key and iC_5 is a heavy key component. The distillate of the second column is fed to heat exchanger to preheat by heat from the reactor and to furnace to heat for reaction. In reactor, nC_4 is converted to iC_4. The nC_4 quantifier is the distillate at the reflux drum of column C2. The heavy impurity quantifier is the bottom at the reboiler of column C2. After isomerization reaction, iC_4 is fed to heat exchanger again to the precool and flooded condenser to cool for feedback to first column. The iC_4 quantifier is the distillate at the reflux drum of column C1.

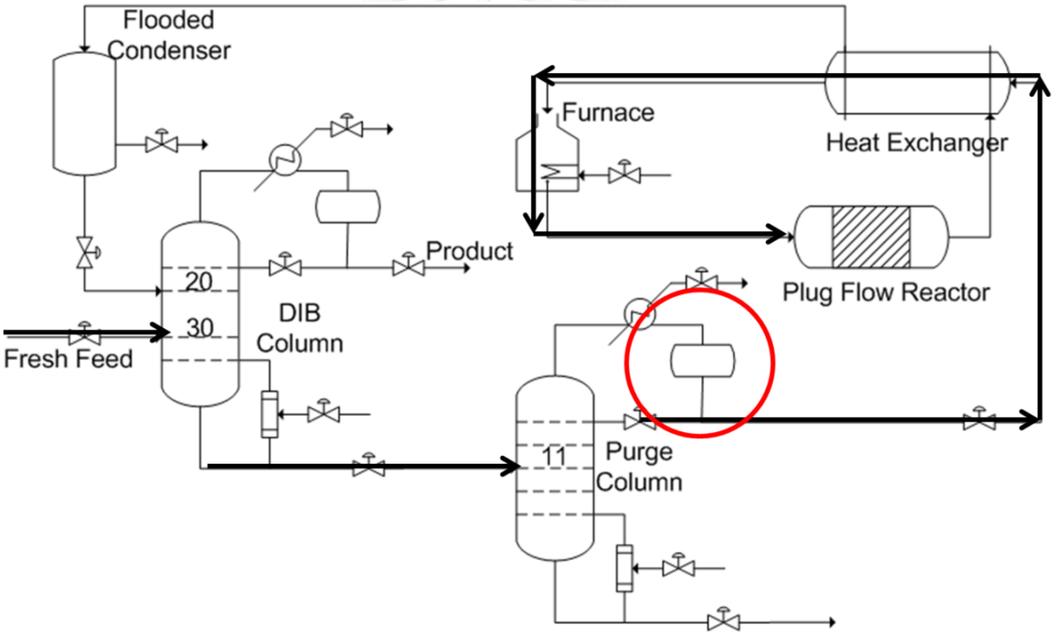


Figure 5.2 nC_4 pathway and quantifier

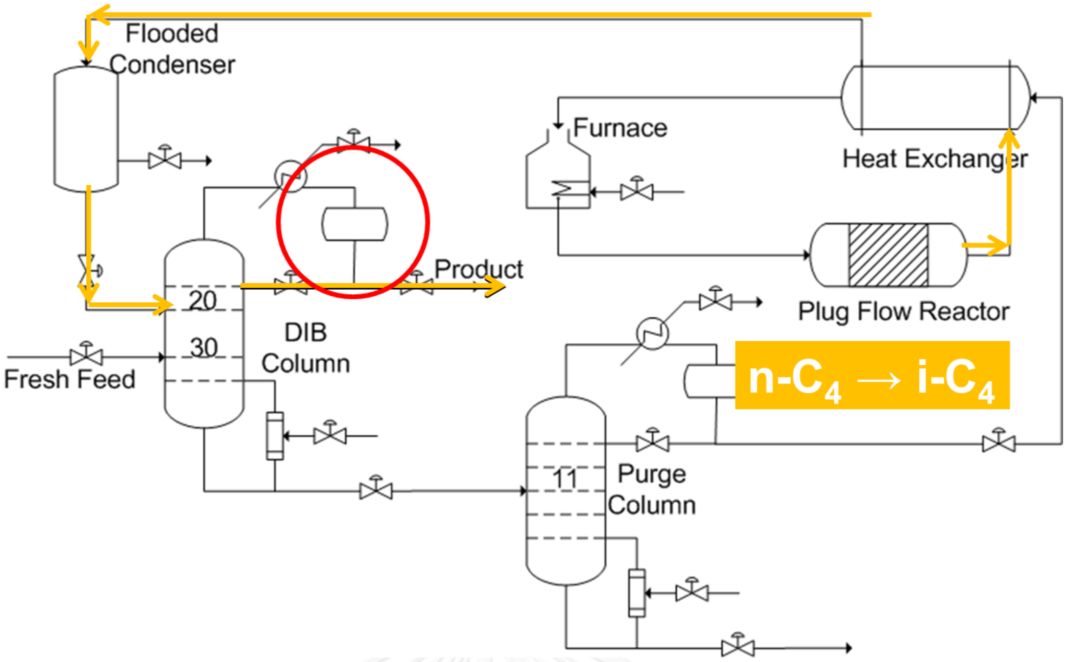


Figure 5.3 iC_4 pathway and quantifier

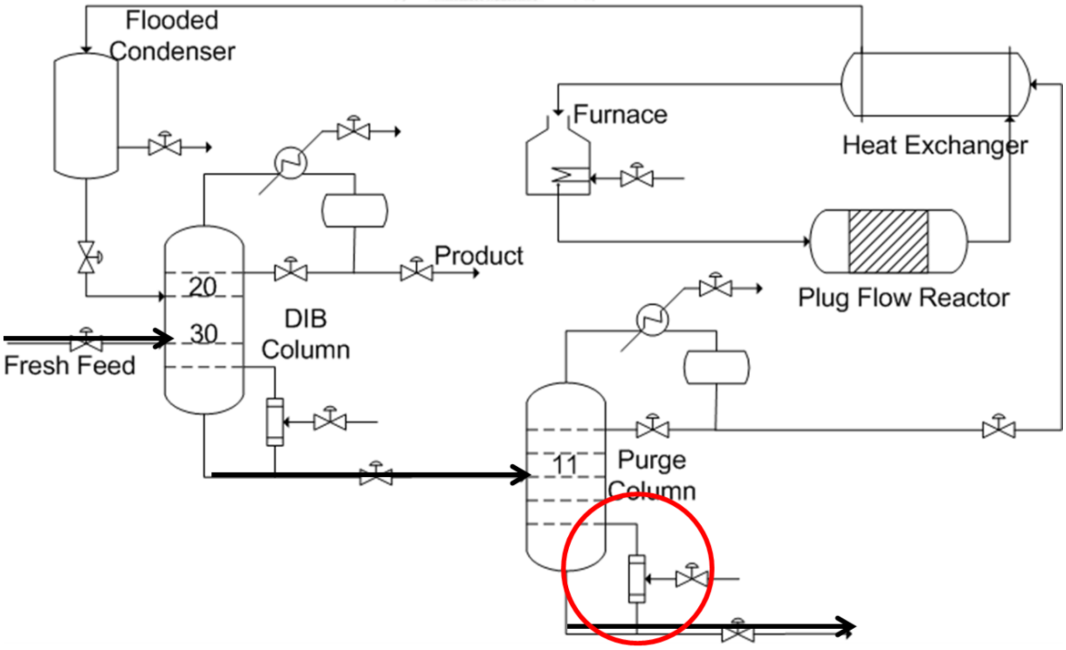


Figure 5.4 iC_5 (heavy impurity) pathway and quantifier

2.5 Assess reaction section.

- Reactor operated at adiabatic condition with reaction nC_4 is converted to iC_4

- Reaction is exothermic reaction

- Heat of reaction is 8,400 kJ/kmol

- Domain of reactor is reactant temperature. Result that from test in steady state simulation with change flow rate, temperature, composition, reactant component and pressure by ±5%, ±10% and ±20% changes. All disturbances are changing at the inlet stream of reactor as show in x-axis and composition of product at the outlet stream of reactor as show in y-axis.

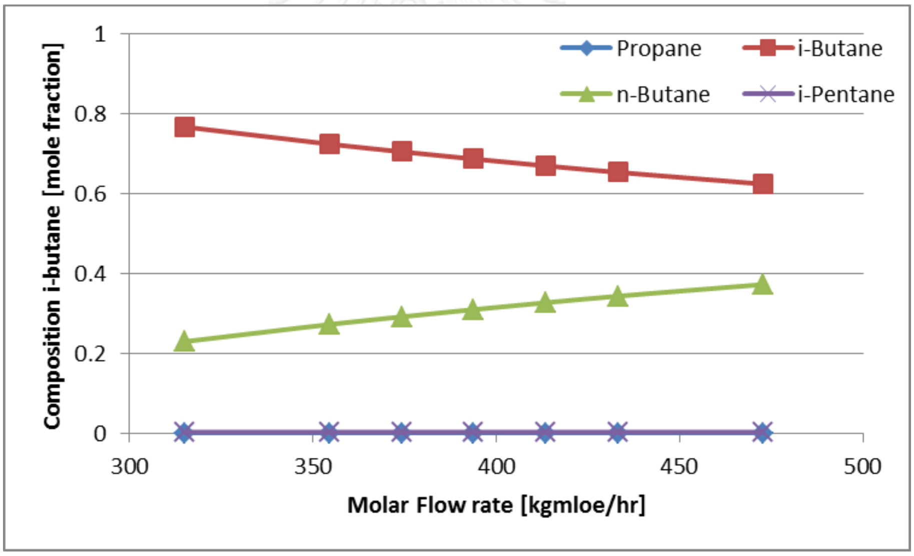


Figure 5.5 Effect of molar flow rate disturbances on composition profile

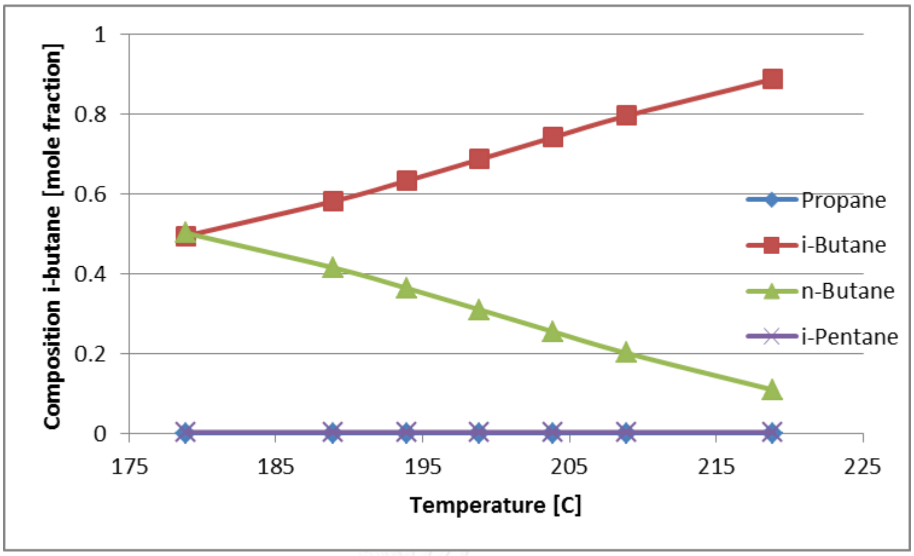


Figure 5.6 Effect of temperature disturbances on composition profile

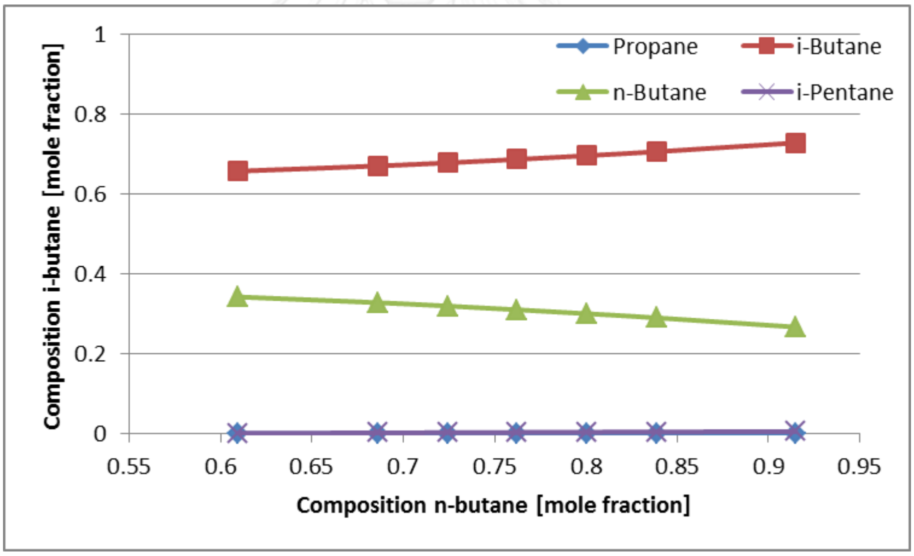


Figure 5.7 Effect of composition nC_4 disturbances on composition profile

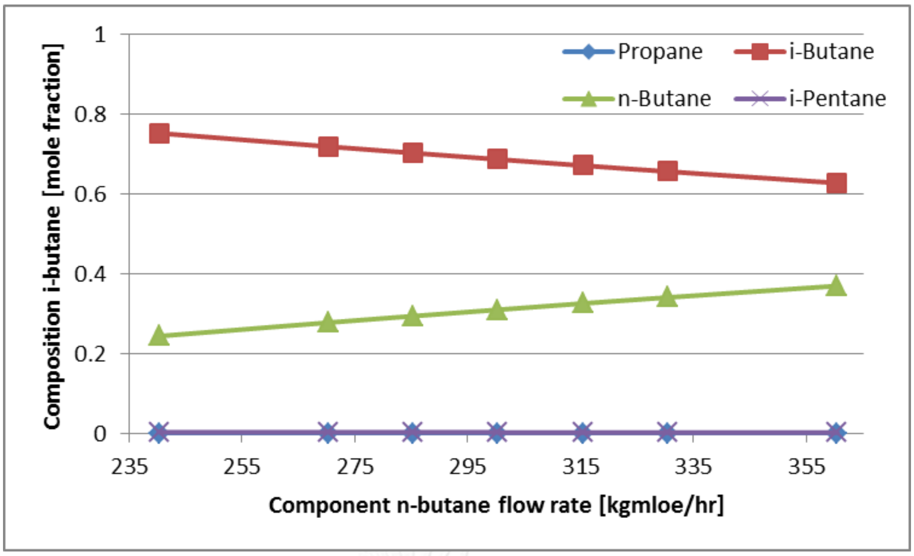


Figure 5.8 Effect of component nC_4 disturbances on composition profile

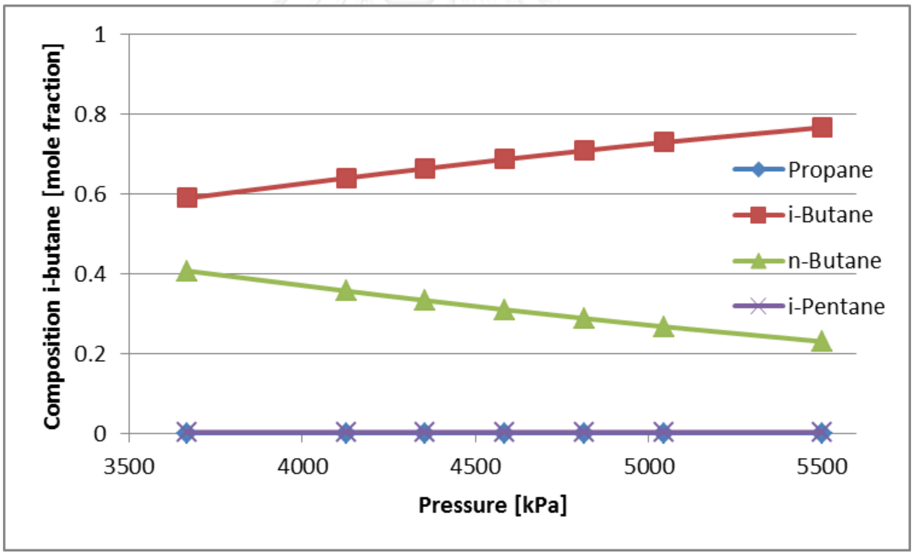


Figure 5.9 Effect of pressure disturbances on composition profile

Figure 5.5 represents the effect in which changing input flow increased the remaining reactant, but decreased the composition of product at output flow. Insignificant change on the quantity of product shows that the flow had little effect on the reaction. Regarding temperature disturbances, products increased at the time input temperature was rising as shown in Figure 5.6. Because temperature is the main factor which directly affects the reaction. Considering the change of reactant composition, It was due to the fact that the increased reactant composition caused the composition of product, which was previously mixed in the reactor substances, to decrease instantly. Referring to Figure 5.7, the increase of the reactant composition of the input flow escalated the reaction rate. Figure 5.8 shows that increasing the reactant component, on the contrary, reduced the rate of reaction as the product composition dropped slightly. Lastly, Figure 5.9 shows the result in view of the input pressure adjustment. The rate of reaction varies directly as the pressure, though the effect is not considerably strong comparing to that of temperature disturbances. It can be concluded that the dominance variable of this reaction is temperature.

2.6 Inspect separation section.

Selection for temperature control tray location

Several methods for selecting the best tray have been proposed. The methods for selecting temperature control tray location are “open-loop sensitivity analysis” and “effect of disturbances”. Sensitivity analyses are performed on the two distillation columns. The procedure of the methods for selecting which tray to temperature control is to look at the steady state temperature profile in the column. The stages in the column are considered where there are large changes of temperature from tray to tray, which is the most disturbed or sensitive points. Note that the tray number is counted from the top (condenser is 0) to bottom stage.

Column C1

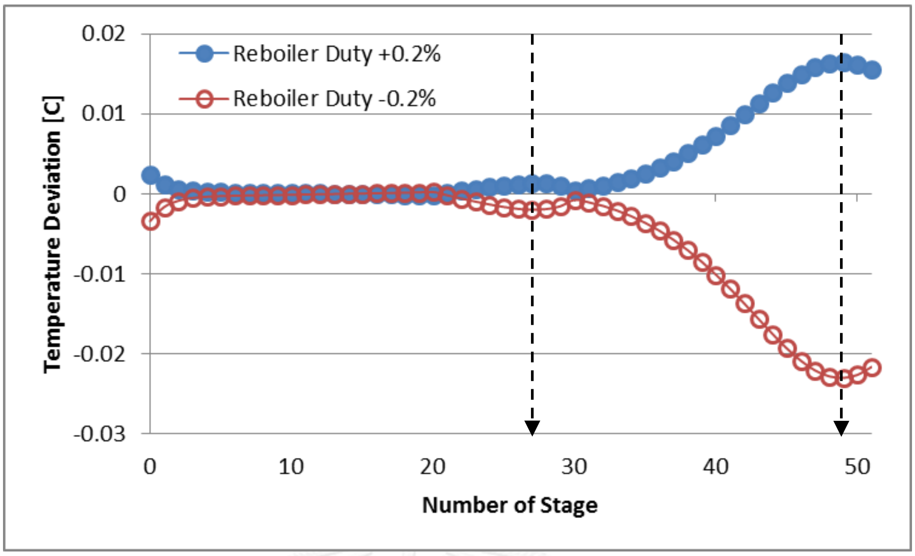


Figure 5.10 Open loop sensitivity analyses between steady stage temperature deviation and manipulate variable (reboiler duty) in the column C1

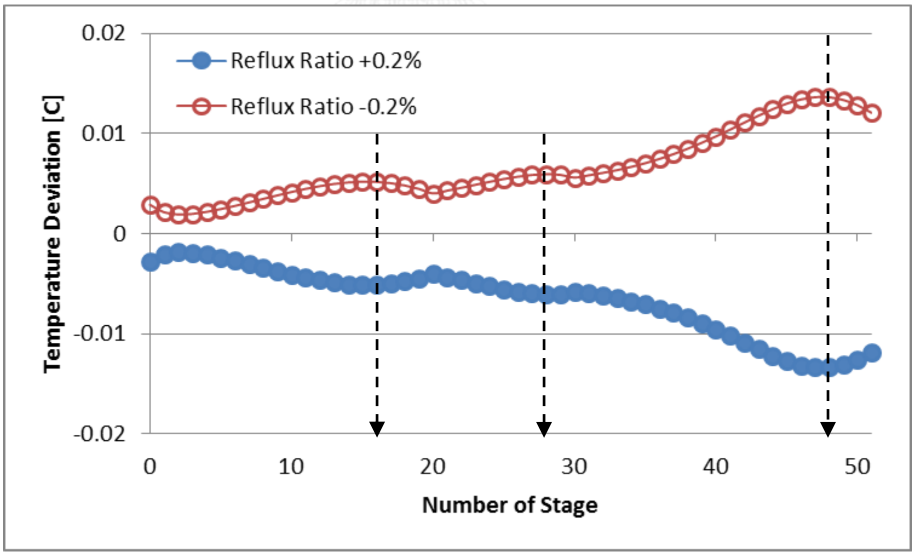


Figure 5.11 Open loop sensitivity analyses between steady stage temperature deviation and manipulate variable (reflux ratio) in the column C1

Open Loop Sensitivity is used stimulation to run at steady state to perform the calculation for outputs of an operation at the distillation column corresponding to given constant reflux ratio and reboiler heat input. Either the reflux ratio or reboiler heat input, then has to be varied for plotting graphs in order to analyses the temperature deviation between at steady state and that with the manipulated variables. A larger difference in temperature reflects the higher sensitivity of the particular tray. To select the specific one out of three different parts, it is considered with two separated feed columns of column C1. Figure 5.10 shows that stage 27 and stage 48 are sensitive stages, as the result of fixing the value of reflux ratio and adjusting the reboiler duty within a range of ±0.2%. Meanwhile, Figure 5.11 represents three sensitive stages: stage 16, stage 28, and stage 47, as the result of fixing the value of reboiler duty and adjusting the reflux ratio within a range of ±0.2%.

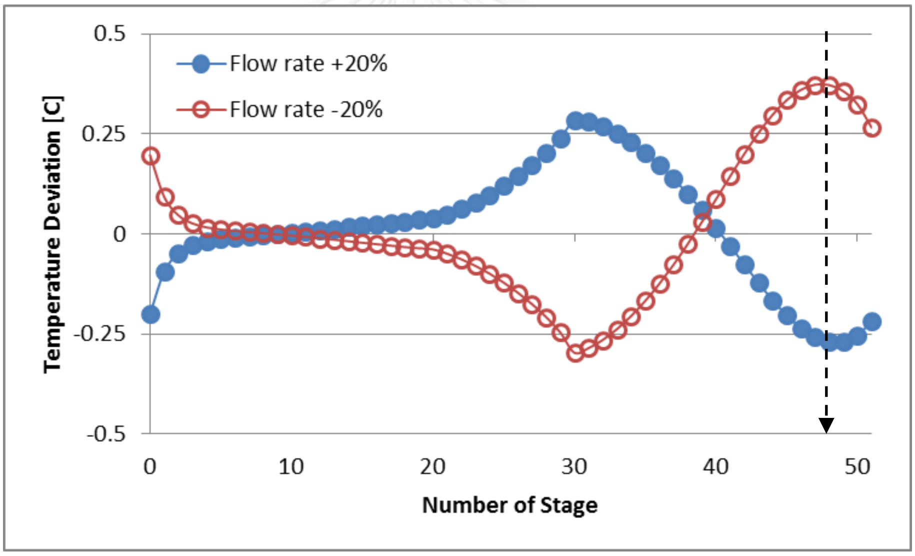


Figure 5.12 Effect of total fresh feed flow rate disturbances on the steady state temperature deviation profiles of the column C1

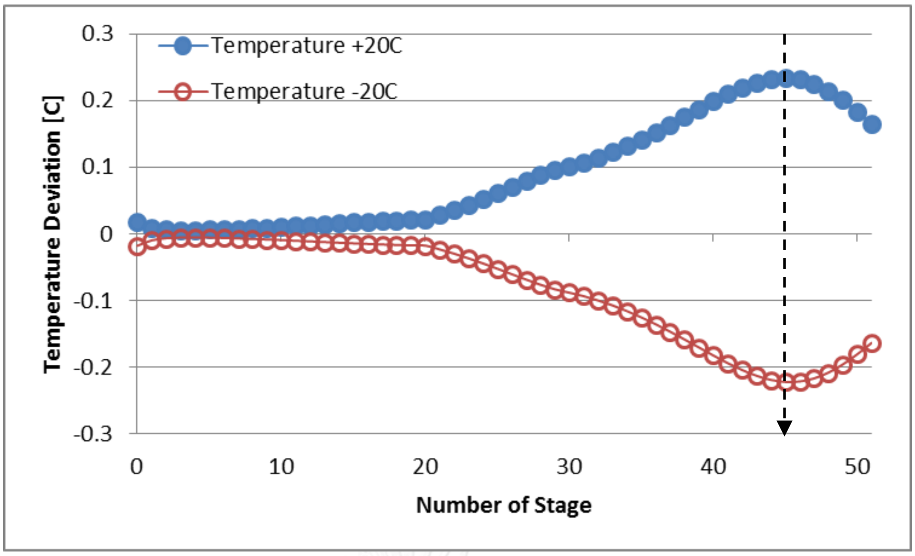


Figure 5.13 Effect of fresh feed temperature disturbances on the steady state temperature deviation profiles of the column C1

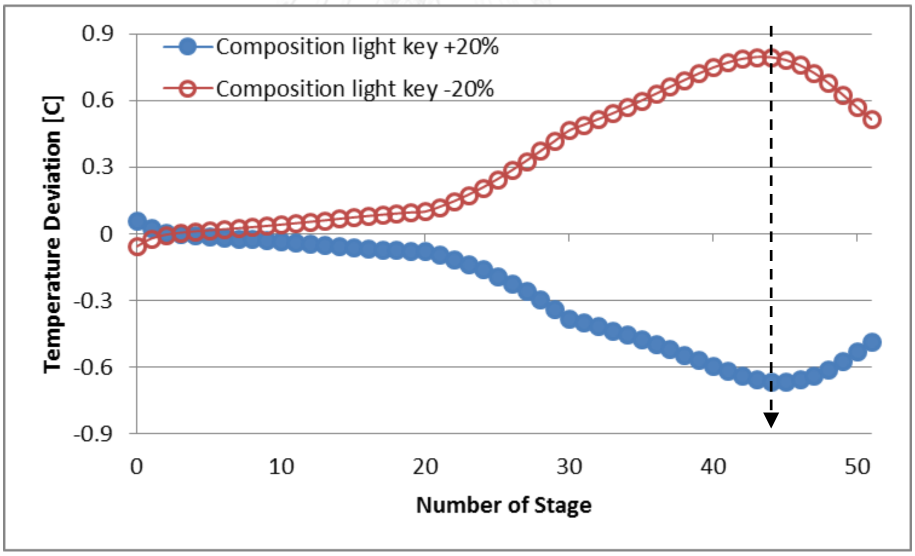


Figure 5.14 Effect of fresh feed light key composition disturbances on the steady state temperature deviation profiles of the column C1

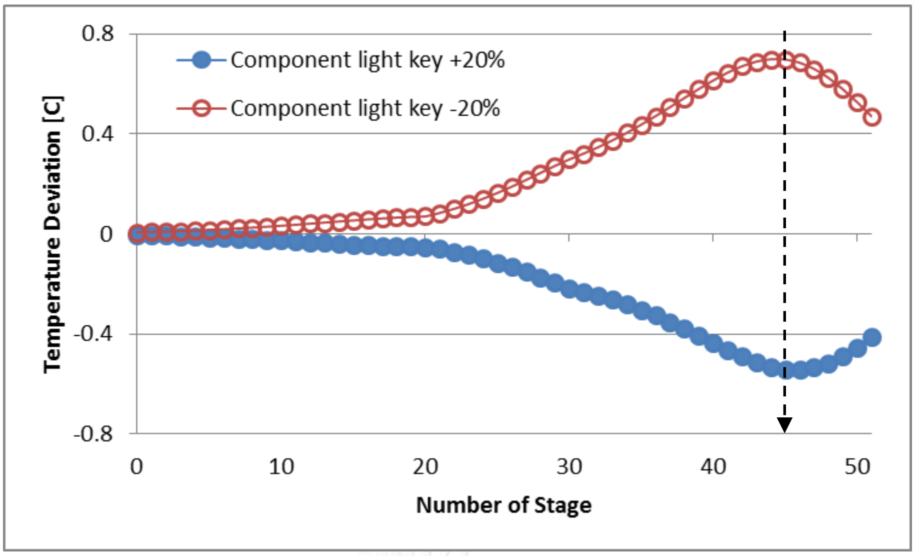


Figure 5.15 Effect of fresh feed light key component flow rate disturbances on the steady state temperature deviation profiles of the column C1

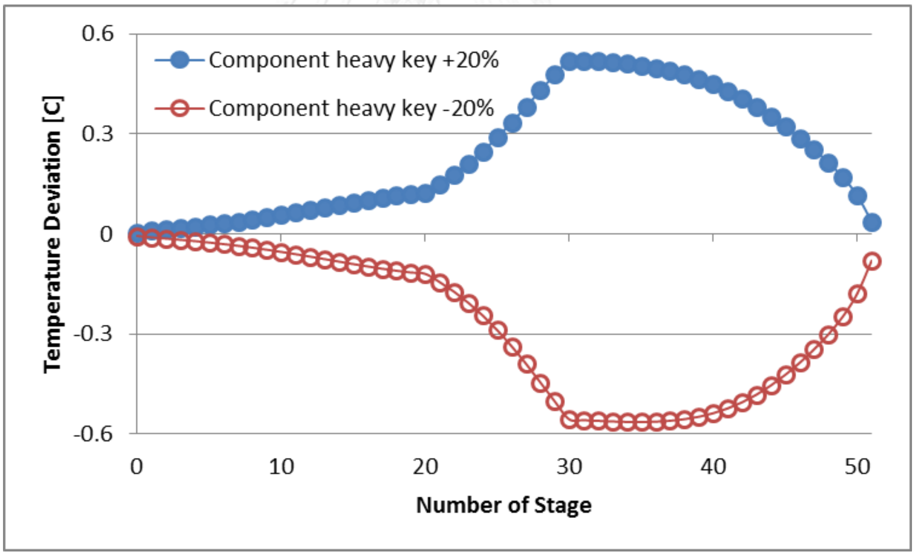


Figure 5.16 Effect of fresh feed heavy key component flow rate disturbances on the steady state temperature deviation profiles of the column C1

Figure 5.12 – 5.16 shows the different temperature deviation profiles of the column C1 for various total feed flow rate, feed temperature, light key component feed composition, light key component flow rate and heavy key component flow rate on steady state analysis. The temperature control location will be selected at a

particular stage with the most sensitive point, where the change will occur prior to those of other trays when there are disturbances at the column.

In order to consider how the effect of disturbances ensues, the fresh feed stream was disturbed by manipulating the total flow within the range of ±20%. Hence, Figure 5.12 shows that stages with substantial temperature deviation are stage 30 and stage 48. However, given that the fresh feed was fed at the stage 30, that stage is not an actual sensitive point of the column. Figure 5.13 to 5.15 respectively represents the result from variations of temperature by ±20 C, light key composition by ±20%, and light key component by ±20%. They also indicate the similar movement of their temperature profile and significant temperature deviation at stage 44. Lastly, changing the heavy key component by ±20% gives the substantial temperature deviation only at stage 30 as shown in Figure 5.16; hence this Figure does not have sensitive point.

Therefore, these two stages temperature are used as the candidates of the controlled variable in the temperature control loop. It is noticed that the temperature control locations for this column accordingly be selected state used in temperature control with two positions is one point near the bottom stage (stage 44). Reason for this selection is that disturbance occurs at a fresh feed; therefore the closing stages to the fresh feed are able to detect temperature deviation faster and one point between the top feed column and fresh feed (stage 28). So, two temperature controllers or dual temperature control can be used in column C1, which might be more effective than one point temperature control. If two point temperature controls are selected, the TC12 controller manipulates the reboiler heat input to control the stage 44 temperature. The TC11 controller manipulates the reflux flow rate to control the stage 28 temperature. This is verified by the dynamic simulation given later in this paper.

The next part address another steady state test for obtaining the control structure. In order for it to happen, the test needs to disturb a feed stream of the column with component flow rate. Lastly, the test is used for determining whether the outflow at a distillate flow or a bottom flow is the component. The column has different settings for the calculation; the settings are reflux ratio, reflux flow, reflux to feed ratio, reflux to (reflux flow + distillate) ratio, reboiler ratio and duel temperature.

In addition, light key component flow rate and heavy key component flow rate are the disturbances with both being varied within the range of ±5%. Manipulating light key by 5% will result only in the change of its volume in distillate flow. While manipulating heavy key by 5% will result only in the change of its volume in bottom flow.

Table 5.3 Result of material shifting test for C1.

	RR	**RF**	**RF/FF**	**RF/(RF+D)**	**V/B**	**Duel Temp**
Distillate	LK	LK	LK	LK	LK	LK
+LK5% (+17.16)	+18.99	+17.76	+18.04	+18.99	+17.93	+17.06
-LK5% (-16.23)	-18.24	-16.46	-17.08	-18.24	-16.68	-16.96
	RR	**RF**	**RF/FF**	**RF/(RF+D)**	**V/B**	**Duel Temp**
Bottom	HK	HK	HK	HK	HK	HK
+HK5% (+15.41)	+15.13	+15.32	+15.79	+15.13	+16.66	+16.14
-HK5% (-14.99)	-14.77	-14.87	-15.44	-14.77	-16.55	-16.10

Table 5.3 shows that there is light key component of 5% (17.16 mol), combining the amount of all column’s feed. Besides, the light key component flows through the distillate stream more than its input. An excess light key component previously flowed out at the bottom. Since the light key component was added, it also left over at the column’s top. Similar to decreasing the light key component by 5%, there is a slight amount of the component at the top because it has more outflows at

the bottom. In case of feeding more heavy key component into column's feed, it can be observed that there is less amount of heavy key component flowing out at the bottom than that was fed. The rest flows out of the column's top. Increasing heavy key yet caused some light key component at the top to transfer through the bottom instead. Decreasing would give the opposite result.

To select the structure, the first column needs to isolate light key component, called product, in order to develop sufficient amount for product quality and product capacity. Therefore, it is expected that if there is more light key (iC_4^+) in the feed column, there will be more outflow at the top. Besides, if there is less light key (iC_4^-) and either more or less heavy key, called reactant, (nC_4^+, nC_4^-) in the feed column, it is likely to be outflow at the bottom. Nevertheless, no structure gives the anticipated result, according to the test.

Column C2

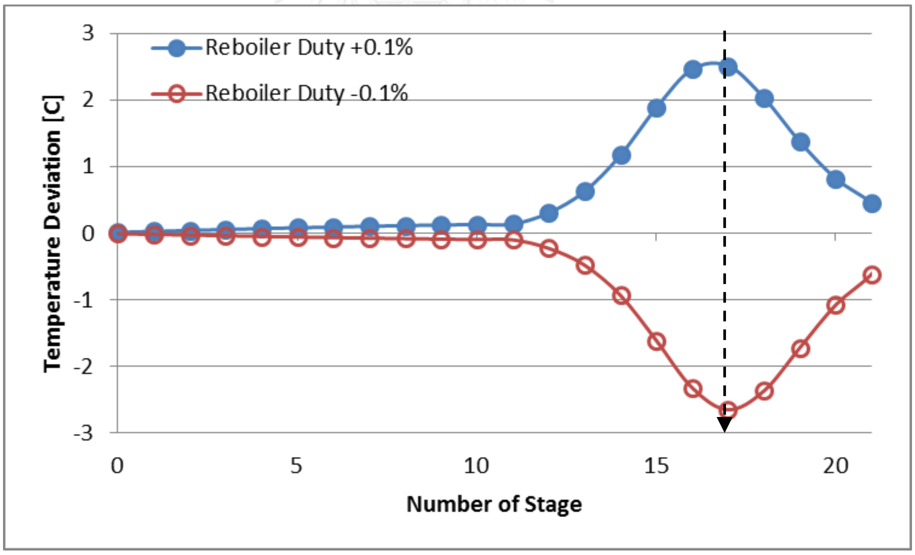


Figure 5.17 Open loop sensitivity analyses between steady stage temperature deviation and manipulate variable (reboiler duty) in the column C2

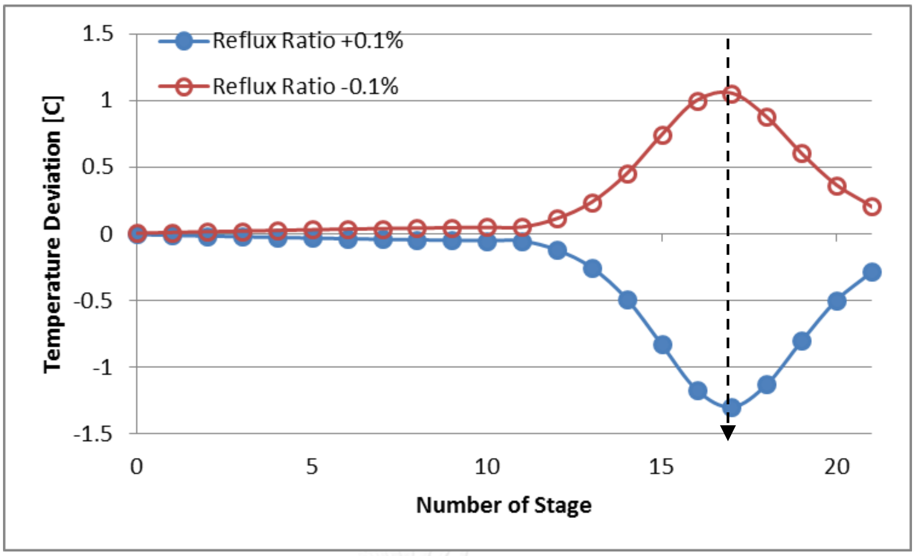


Figure 5.18 Open loop sensitivity analyses between steady stage temperature deviation and manipulate variable (reflux ratio) in the column C1

Figure 5.17 and 5.18 show that stage 17 is sensitive stages, as the result of the value of reboiler duty by ±0.1% and reflux ratio by ±0.1%, respectively.

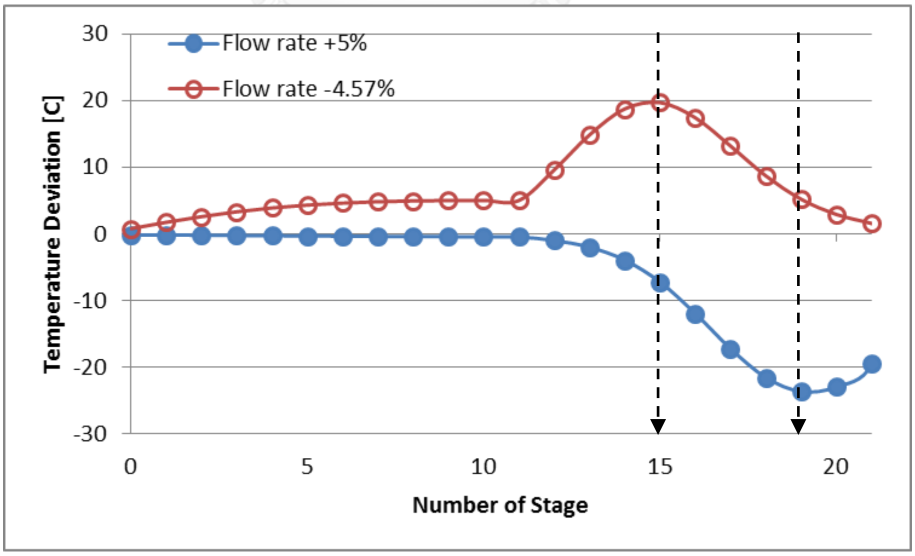


Figure 5.19 Effect of total fresh feed flow rate disturbances on the steady state temperature deviation profiles of the column C2

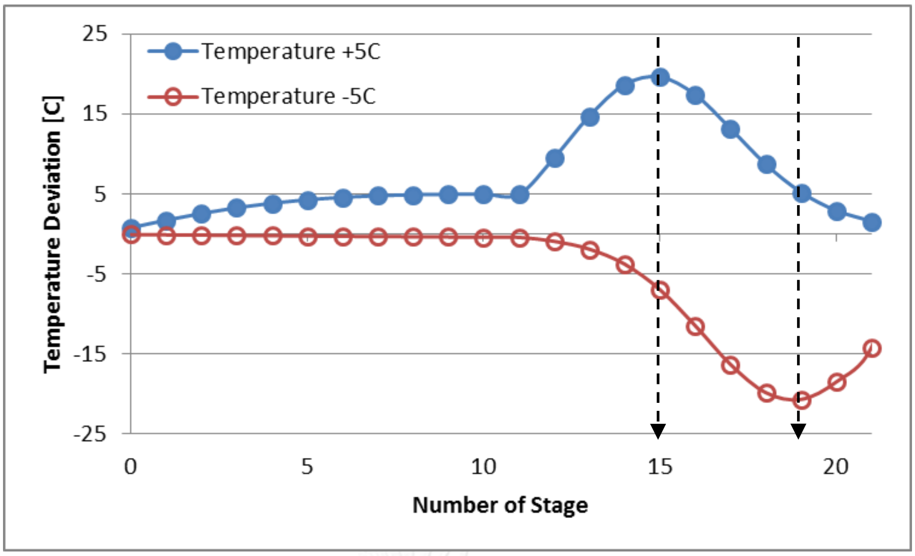


Figure 5.20 Effect of fresh feed temperature disturbances on the steady state temperature deviation profiles of the column C2

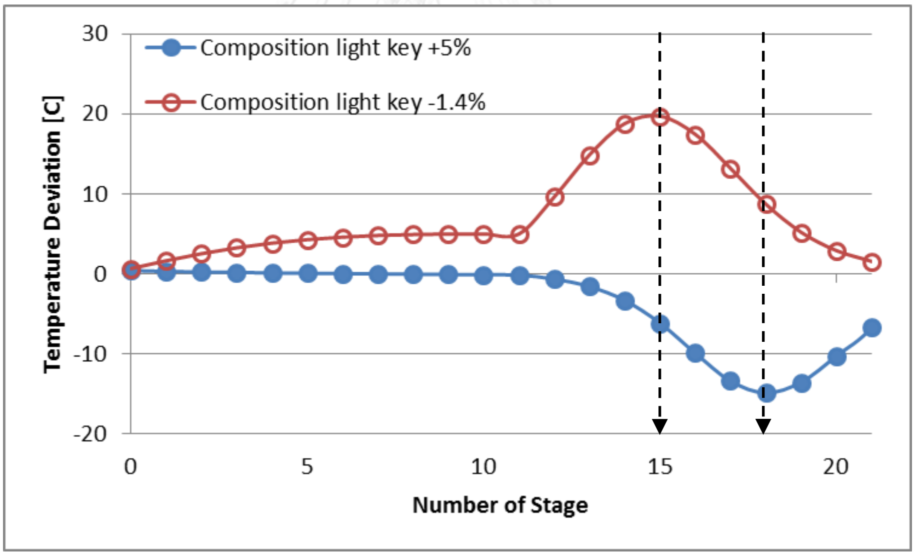


Figure 5.21 Effect of fresh feed light key composition disturbances on the steady state temperature deviation profiles of the column C2

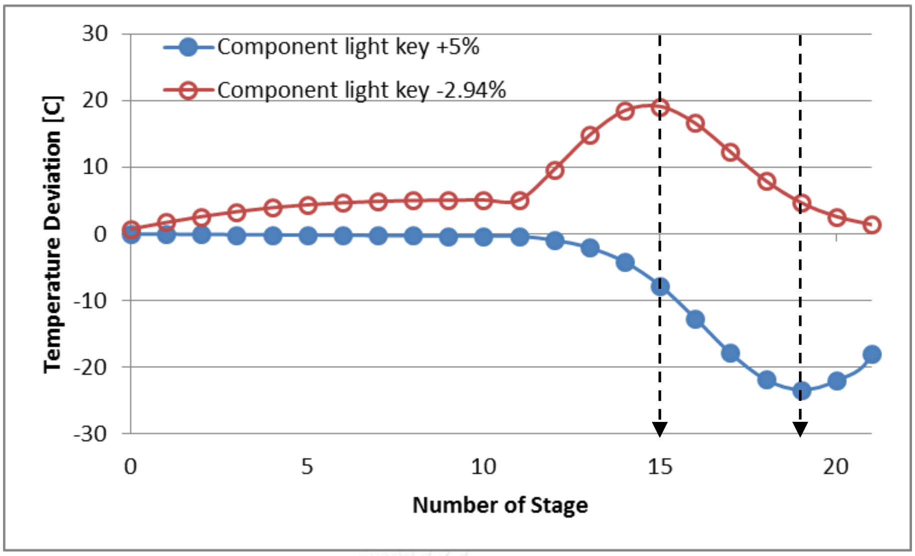
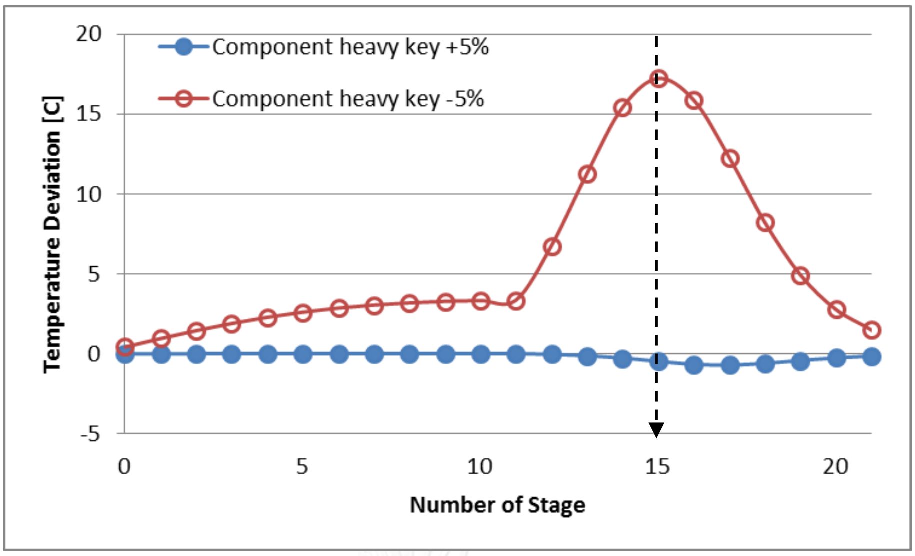
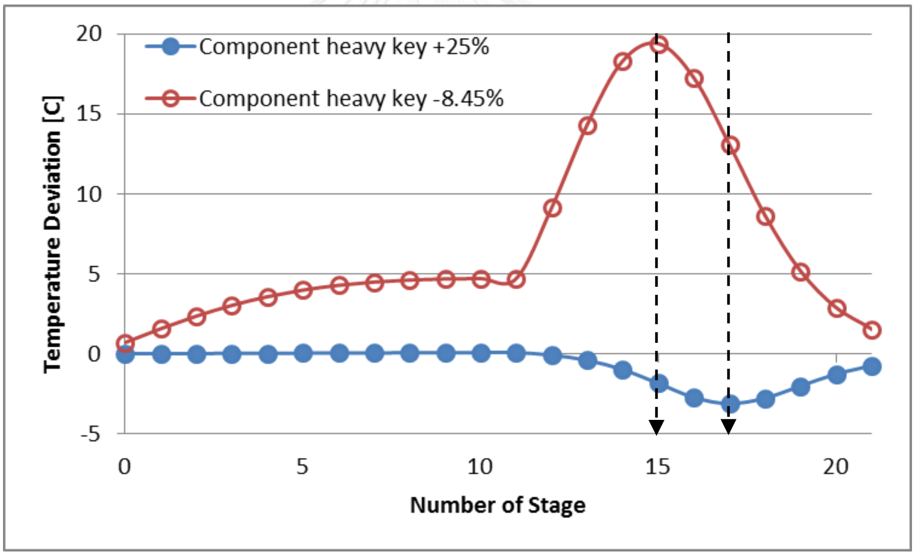


Figure 5.22 Effect of fresh feed light key component flow rate disturbances on the steady state temperature deviation profiles of the column C2



(a)



(b)

Figure 5.23 Effect of fresh feed heavy key component flow rate disturbances on the steady state temperature deviation profiles of the column C2

Figure 5.19 represents the effect of manipulating the flow with growth of 5% and a reduction of 4.57%, which is the lowest value possible. Stage 15 is the sensitive stage resulting from decreasing flow and stage 18 is the one resulting from increasing flow. Figure 5.20 is the result from the variation of temperature within ±5C. It denotes that the stages express inverse results between increasing’s and decreasing’s. Figure

5.21 describes when the composition light key grows by 5%, but it can be decreased only by 1.4%. Therefore, decreasing composition gives stage 15 and increasing composition gives stage 17. While Figure 5.22 describes when the Component light key grows by 5%, but it is possibly decreased only by 2.94%. It has the same sensitive stage as that of Figure 5.19. Figure 5.23(a) shows that manipulating value of composition heavy key by 5% only yields stage 15 as the sensitive stage resulting from decreasing side. While increasing side is not able to give any sensitive stage. Therefore, the component heavy key can be risen up to 25% and give the sensitive stage at stage 18. Though, it can be reduced to 8.40% and expresses the same sensitive stage.

Table 5.4 Result of material shifting test for C2.

	RR	**RF**	**RF/FF**	**RF/ (RF+D)**
Distillate	LK	LK	LK	LK
+LK5% (+15.04)	+15.10	+15.11	+15.10	+15.10
-LK5% (-15.04)	-14.97	-14.98	-14.98	-14.97
	RR	**RF**	**RF/FF**	**RF/ (RF+D)**
Bottom	HK	HK	HK	HK
+HK5% (+0.726)	+0.67	+0.67	+0.68	+0.67
-HK5%	-0.69	-0.69	-0.70	-0.69

(-0.726)				

Table 5.4 has the same patterns of substance separation, which address both parts of adding light key component and adding heavy key components, as shown in Table 5.3. The similarity of those patterns is caused by both particular columns are hydrocarbon compounds.

To select the structure, it can be considered that the second column needs to isolate light key component, called reactant, in order to develop sufficient amount for product quality and product capacity. It is then expected that if there is more light key (nC_4^+) in the feed column, there will be more outflow at the top. Besides, if there is less light key (nC_4) and either more or less heavy key, called heavy impurity, (iC_5^+, iC_5^-) in the feed column, there is likely to be outflow at the bottom. However, according to the previous test, no structure gives the anticipated result.

2.7 Schematize production rate manipulation and mode of operation.

Discipline to select control scheme are on-supply mode and on-demand mode.

In case of on-supply mode, fix flow at fresh feed stream, at tanks to control level to adjust another flow stream next to that tanks or opposite of fresh feed stream as shown to the Figure 5.24. That is reactant into the process focused on constant. When reactant into the process as needed to get the product in quantities it should be. In case of on-demand mode, fix flow at product stream, at tanks to control level to adjust another flow stream before that tank or opposite of product stream as shown in the Figure 5.25. That is the product out of the process focused on constant. When product out of the process as needed to get the product in quantity system will adjust to the return to the reactant stream. Both on-supply mode and on-demand mode can mix together called on-internal mode by control fix flow between fresh feed stream and product stream. This position will be the control is divided into two parts. The first part, from fresh feed stream to that position control with on-demand mode And the second part, from that position to product stream control with on-supply mode as shown in the Figure 5.26.

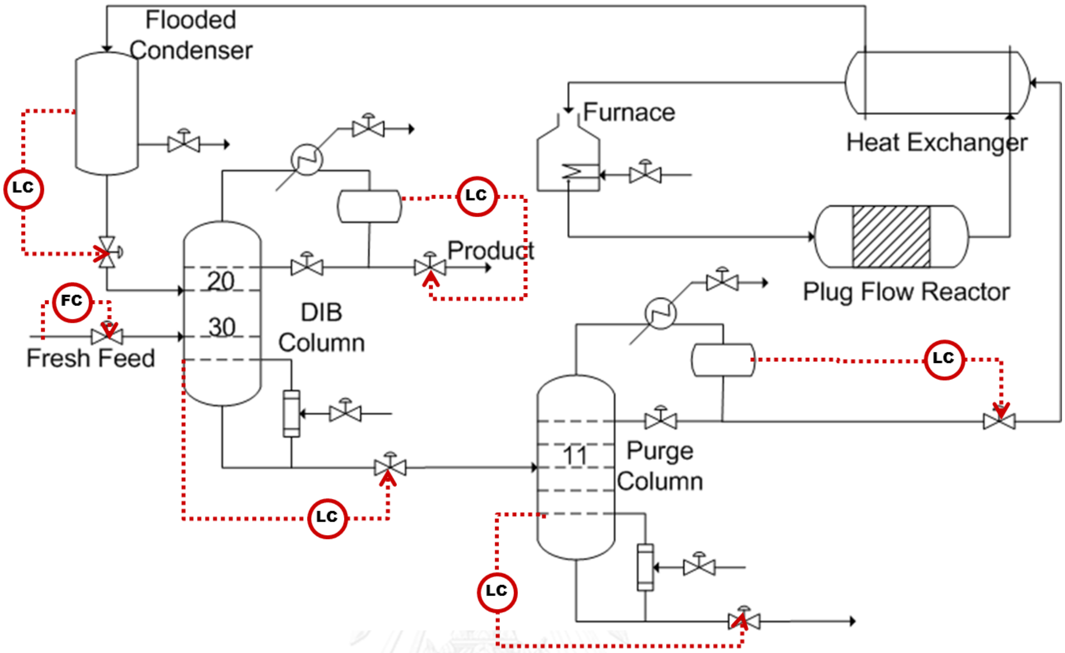


Figure 5.24 Flowsheet of butane isomerization process to use mode of operation on-supply mode

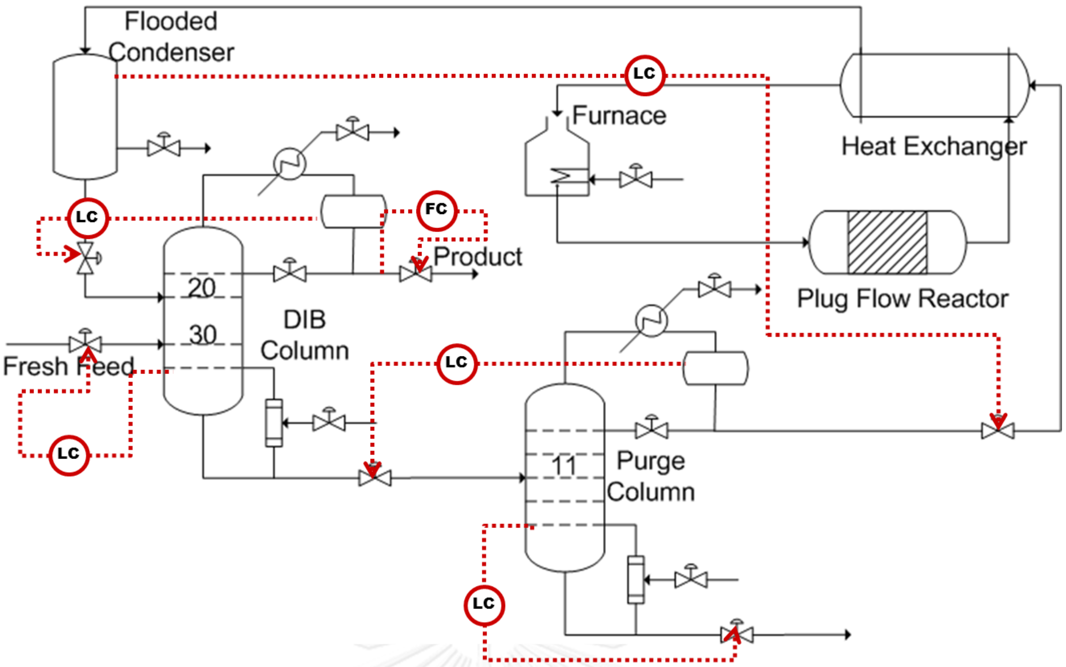


Figure 5.25 Flowsheet of butane isomerization process to use mode of operation on-demand mode

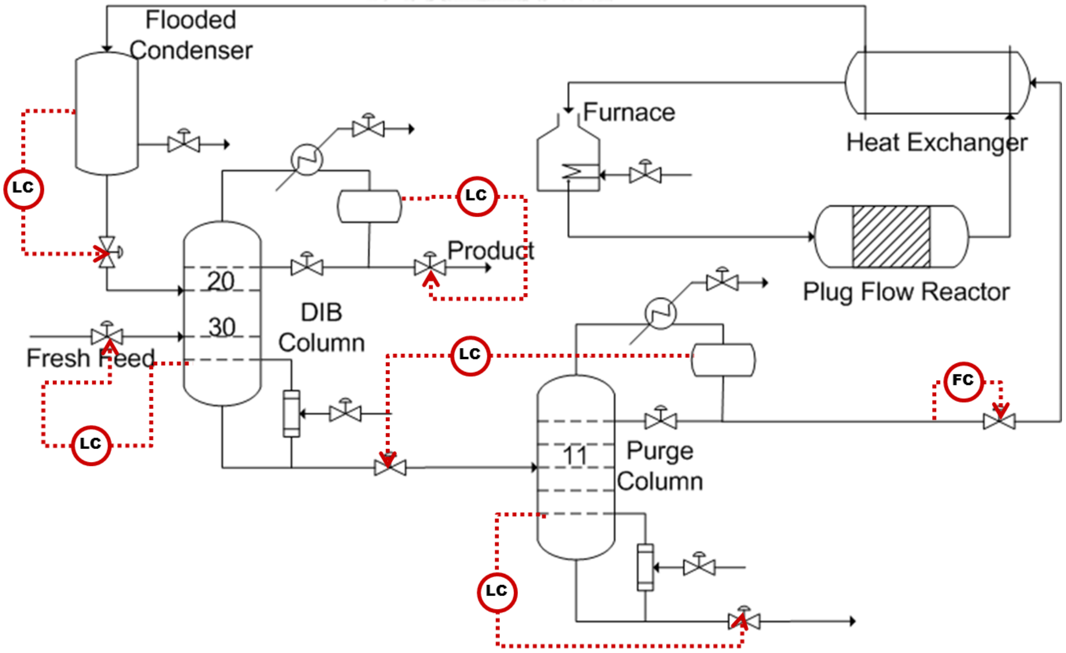


Figure 5.26 Flowsheet of butane isomerization process to use mode of operation on-internal mode

Stage 2: Plant Level Design: Fixture Plant Design and Disturbance Management.

Step 3: Establish fixture plant.

The principal idea of establishing a fixture plant is to first have entire plant fluid-filled and material-balanced. This idea is similar to creating “hydraulic” control structure proposed by Buckley (1964). This step in Wongsri’s plantwide control structure design procedure is classified into 5 topics.

3.1 Keep the materials entered and/or reentered fixed.

There is one material stream fed into the process: a mixture of a reactant (nC_4), a product (iC_4), and impurities (C_3 and iC_5). The stream will then go into a separate process. The product is separated from the reactant. The reactant will be sent into the reaction process and later turned into the product. It will then be fed into the recycle stream which is the distillate of the purge column and go back to the separation process. This makes the recycle stream different from the normal one as it is filled with the product instead of the reactant. One might consider the stream leaving the reactor and feeding the DIB column is the recycle stream too. The reactant fed to the plant plus reentered via the recycle stream should be kept constant according to this Wongsri’s instruction. The appropriate entry point is the point that is rich in nC_4 (the most) and has the ability to control the reflux drum effectively. C2 distillate stream, therefore serves as the best entry point in this case. Besides, the quantifier and the controlling position should not be alienated because in a normal case, a sequence of unit operations in a chemical plant starts with a reactor and then a separator with the feed entering the reactor. The C2 distillate flow (nC_4 entered and reentered) into the reactor must be fixed.

Next, how do we regulate nC_4 in the plant? We must look at nC_4 quantifier which is the reflux drum level of C2 indicating nC_4 level in the plant. This also indicates whether nC_4 should be fed in more or less according to the reflux drum level. You should not forget that C2 distillate is fixed. Therefore, C2 reflux drum level is used to adjust the fresh feed by the following order: C2 reflux drum level adjusts C2 feed, C1 base level adjusts the fresh feed.

It is seen that by following the loop design of Wongsri's procedure, i.e. regulating nC_4 at its quantifier (C2 reflux drum level) by manipulating its handler which is nC_4 in the main feed and recycle stream, the mode of operation is dictated. The nC_4 quantifier (inventory) control is prior to the others and it must be controlled as 'on-supply production'. The above control structure is on-internal mode. (Figure 5.27)

Alternatively, the combined nC_4 in the flooded condenser outlet stream and in the feed is used to regulate the flow of the feed. Therefore, the nC_4 leaving the DIB column, the purge column, and entering the reactor is maintained. The amount of nC_4 in the plant is represented by the reflux drum level in the purge column, see Figure 5.28.

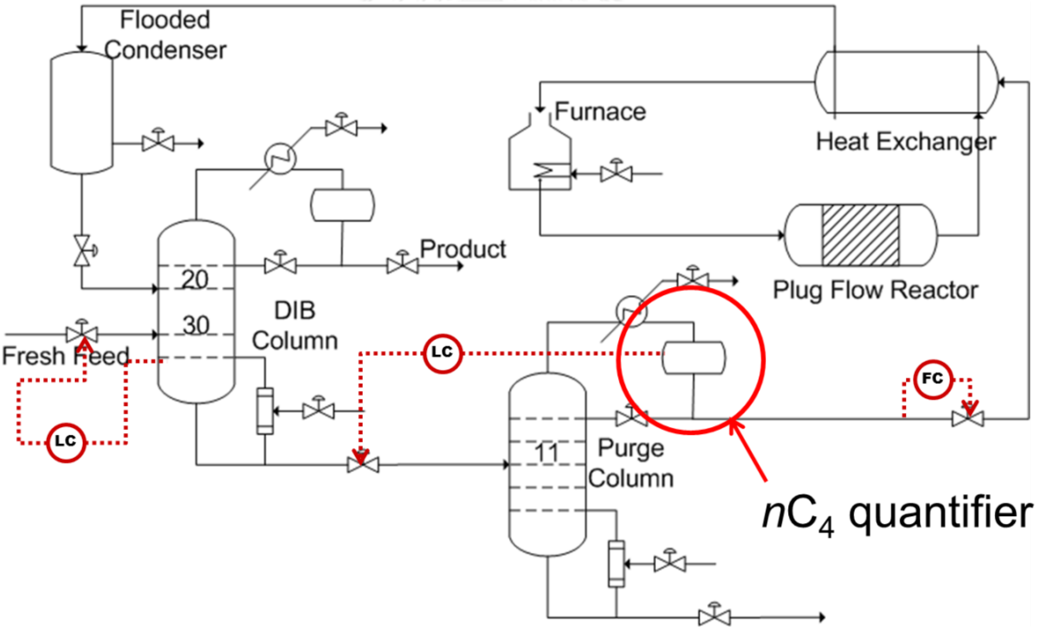


Figure 5.27 Flowsheet of butane isomerization process to keep the raw materials entered and reentered fixed

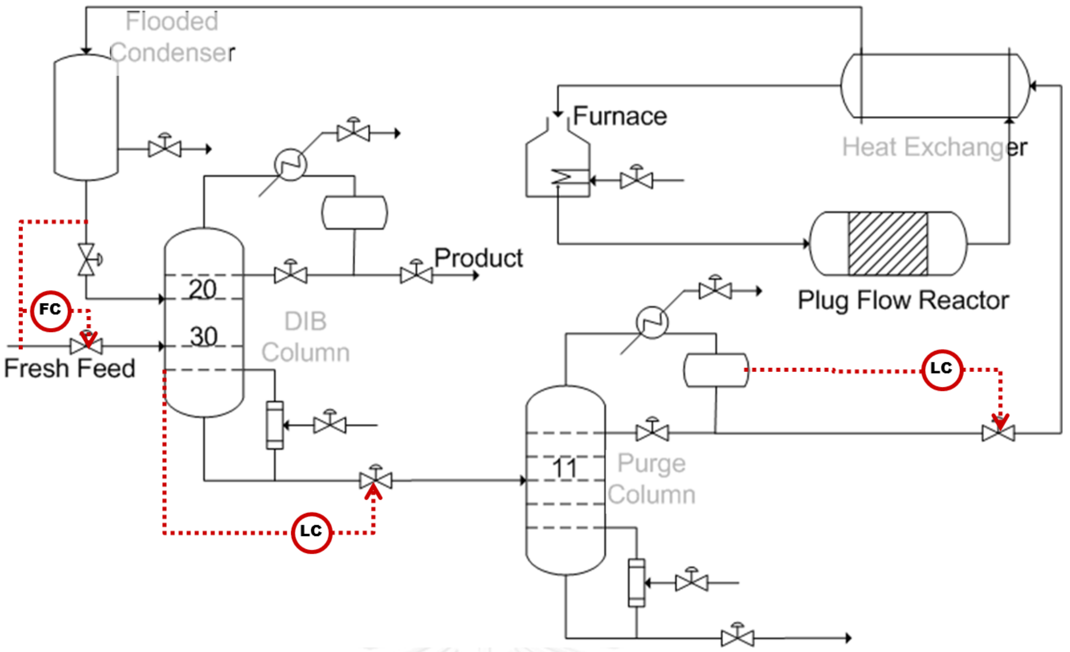


Figure 5.28 Flowsheet of butane isomerization process to keep combine feed

3.2 Adjust the flow of exit material streams (products, byproducts, and inert) according to their accumulations.

The chemical processes are classified into three types: products, by-products and inert impurity. In the butane isomerization process, the exit material stream is iC_4 product, C_3 impurity and iC_5 impurity. Product component conducts out of the process though the distillate of C1 column with 93.5 %mol. It's adjusted according to its accumulation via the reflux drum level control of Column C1. C_3 which is the light impurity feed to the column C1 and leaves from the top of column C1 with the product. iC_5 impurity separate from reactant at column C2

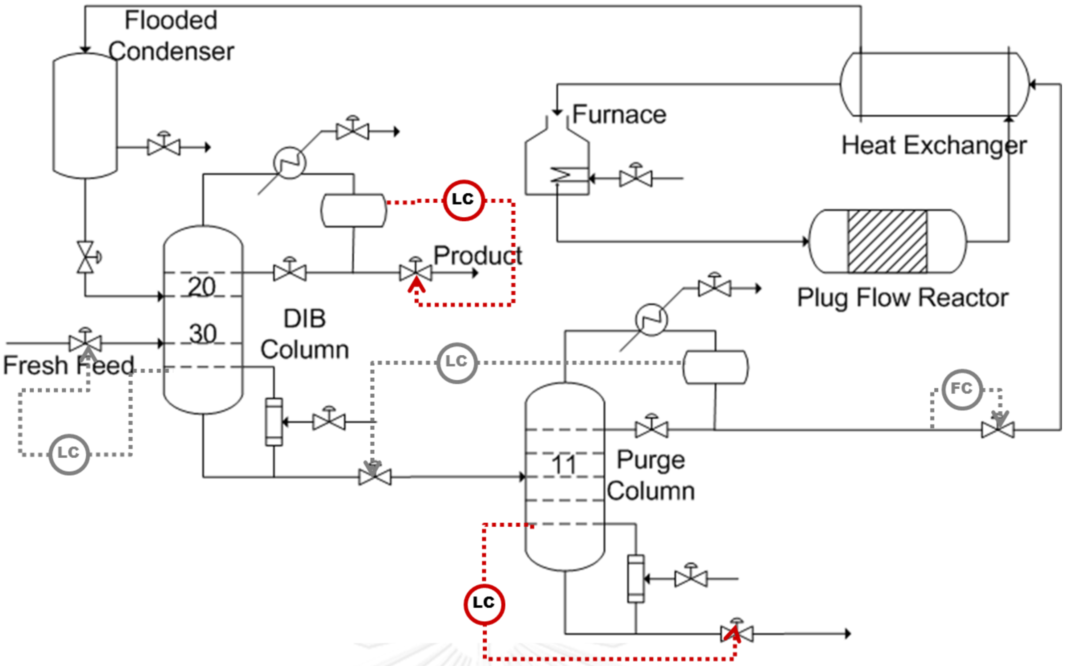


Figure 5.29 Flowsheet of butane isomerization process for adjusts the flow of exit material streams according to their accumulation.

To purify product at the top of the column is to control the level of reflux drum and adjust at the stream both reflux flow and distillate flow. Theoretically, to control levels at reflux drum should adjust reflux flow in case of reflux ratio more than 5. If the reflux ratio is less than 1, Distillate flow is better control. If the reflux ratio is between 1 and 5, distillate flow is good to control to level. The distillate flow was adjusted to purify at column C1 although it constrain is reflux ratio more than 5 because exit material should be adjust first and then adjust reflow flow for reflux ratio more than 5 in next step. In the bottom of column should control bottom flow.

3.3 Handle the material that’s not leaving the process.

Only nC_4 not leaves the process because it enters, reacts and recycles. Therefore, this step is not controlled.

3.4 Control the amount of the rest of the component at their quantifiers.

The quantifier of components can be implied as a quantity or amount of the specified component such as level is liquid quantifier, etc. In the butane isomerization process has 4 components: nC_4, iC_4, C_3 and iC_5. Quantifiers of each component are different location as shows in Table 5.5. iC_4 product and C_3 impurity are identified in Step 3.2. The rest of the component is nC_4, which the quantifier is decided to be the reflux drum level of C2. The control loop is placed to regulate its inventory via the level controller (LC) which it obtains in Step 3.1.

Table 5.5 Locate the quantifiers of the components

Components	**Quantifiers**
nC4	Reflux drum of column C2
iC4	Reflux drum of column C1
iC_5	Base of column C2

3.5 Maintain the production rate.

3.5.1 Consume the limiting reactant.

This procedure is to designate a chemical substance as limiting reactant and another follow to stoichiometry. The reactant of butane isomerization is only nC_4, so there is no need for this procedure

3.5.2 Regulate the production rate.

This procedure is to control chemical activity enough to product. Refer to Step 2.5, process variable which affects to productivity is temperature. So inlet temperature of the reactor was selected to study by adjusting temperature via heat input at the furnace. Reactor feed temperature is suggested to be adjusted by iC_4 composition of the reactor outlet to cascade its.

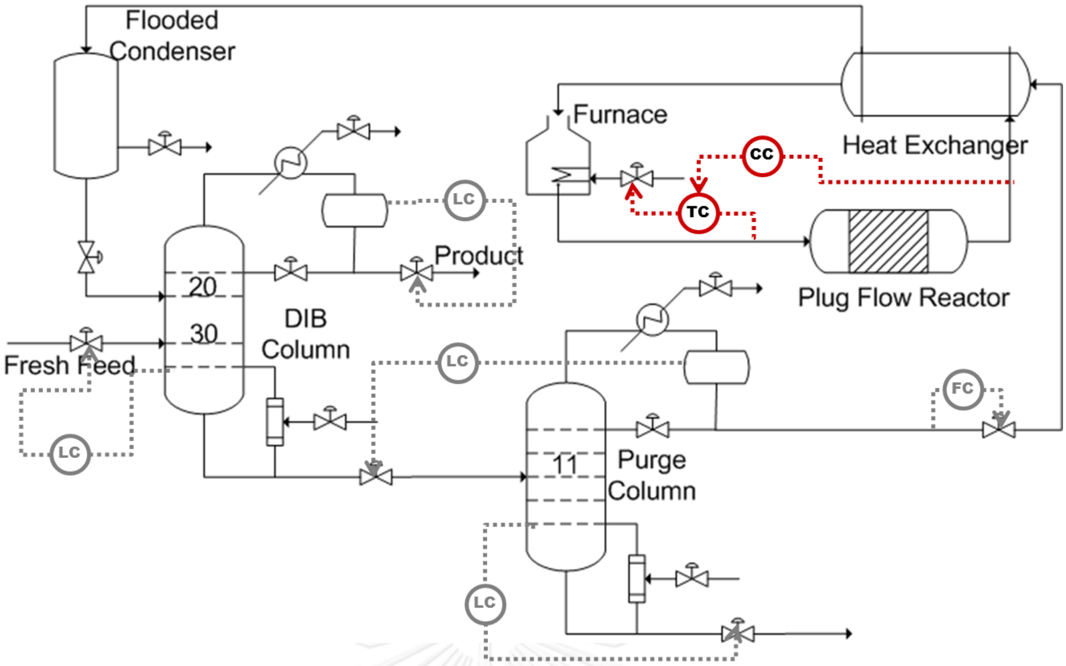


Figure 5.30 Flowsheet of butane isomerization process for regulate the production rate.

Step 4: Handling the disturbances

In this step, we configure the quality control loops by employing the notion of all disturbance management. The type and magnitude of disturbances affects to a reaction and separation section. The design of the control loops to manage with both thermal (Step 4.1) and material (Step 4.2) disturbances, which the heat disturbances are classified into 2 types. An analysis of disturbances encountered in the butane isomerization process follows:

4.1 Manage heat disturbance.

4.1.1 Direct the heat disturbances that are not directly related to quality. Since there is no have thermal manage a point.

4.1.2 Manage the heat disturbance that related to quality in order to maintain the product constraints.

To complete controller for temperature control is to remove heat disturbance affecting to the process as soon as possible

For butane isomerization process, Heat disturbance can remove at 3 points.

- The first, second positions are C1 and C2 where heat disturbance removes at the heat input of reboiler and affect to product quality (4.1.2). When the heat input of reboiler was adjusted, light key and heavy key of the outlet at the top of column change.
- The last position at the furnace, when heat disturbance incurs before arriving or after arriving in there, it will be eliminated by not directly related to product quality (4.1.1).

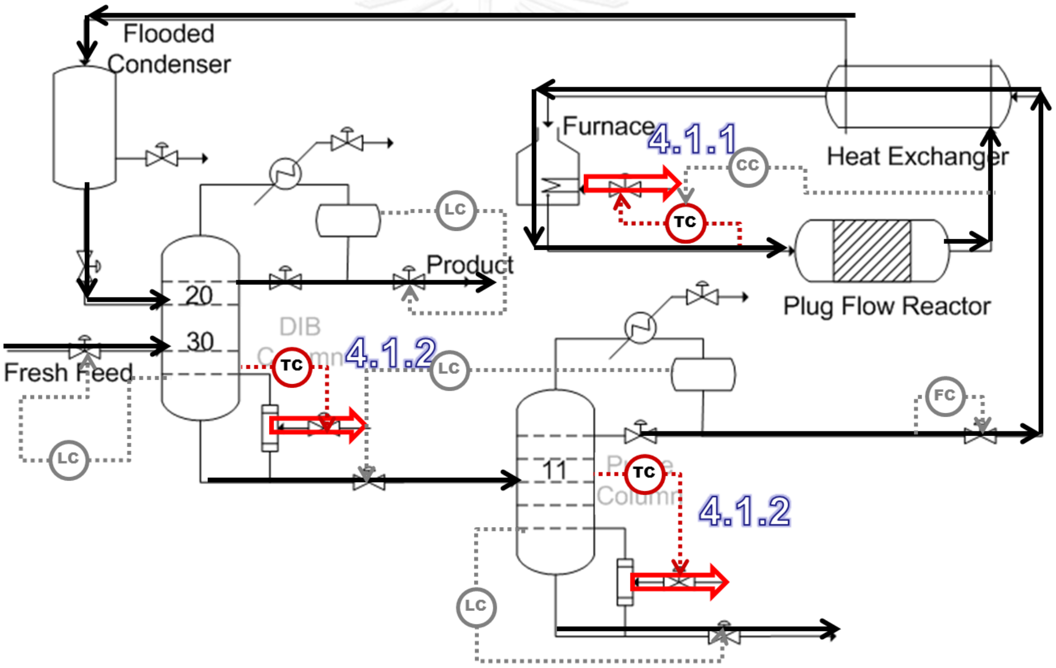
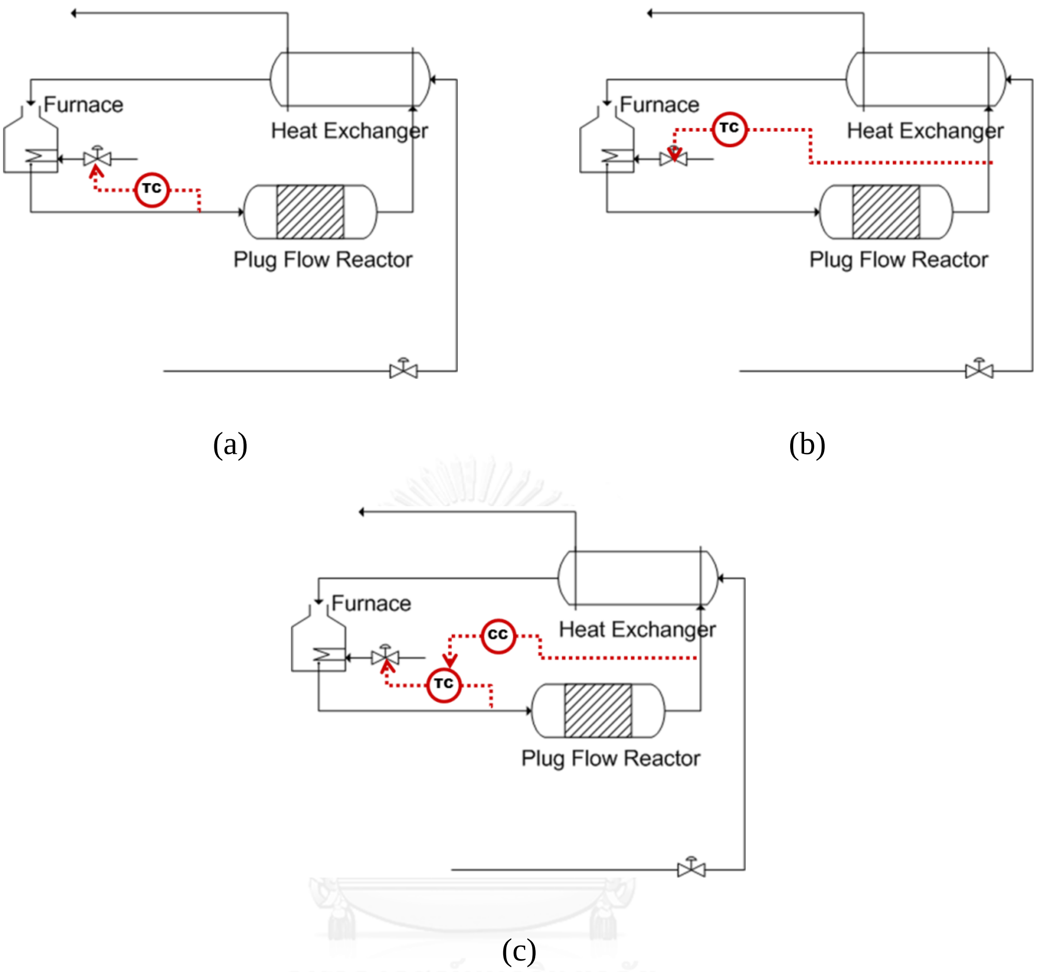


Figure 5.31 Flowsheet of butane isomerization process for handle the heat disturbances.

4.2 Manage material disturbances.

This process has two important sections; reaction section and separation section. Control structure in each section has been specified by analyzing the material disturbance tests on changes of flow rate, temperature, composition and component. The material path ways demonstrated to manage each material according to their pathways. The material pathways of any material presented in Figure 5.2-5.4 or Table 5.5. This step is to identify and manage the material pathways of each component. Two components need to be analyzed by including iC_4 and nC_4. An exclusion of C_3 is caused by entering C_3 at fresh feed always flows out at the top of C1 supplemented with products. The propane is not associated with any other process. This is also on the grounds that C_3 is going through the bottom of C2, which causes the least possibility of contaminating reactors.

Control structure of reactor



(a) (b)

(c)

Figure 5.32 The control structure for handle disturbances in reactor section

(a) Fixed inlet temperature (IT) control structure
(b) Fixed outlet temperature by cascade to inlet temperature (TCT) control structure
(c) Fixed outlet composition by cascade to inlet temperature (CCT) control structure

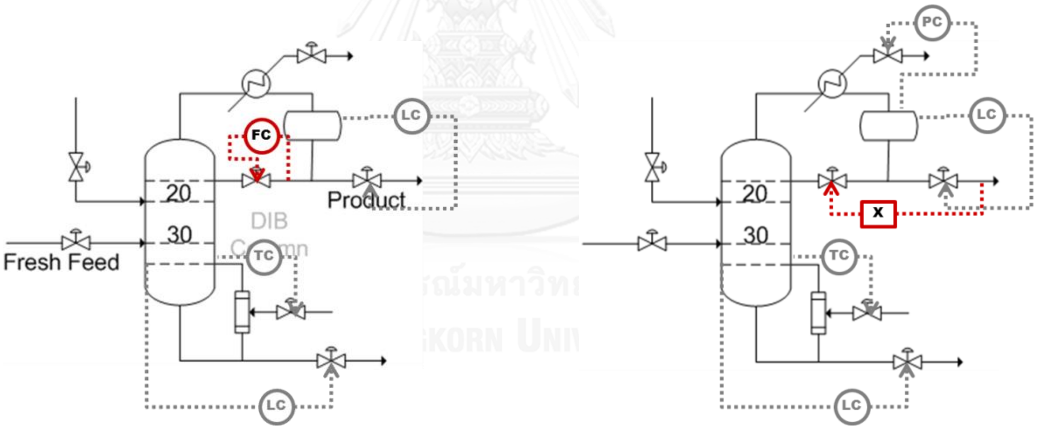
There were 3 controlling structures of reaction section that were chosen for testing and analyzing.

- Figure 5.32 (a) which was the result from earlier step was shown fixed inlet temperature control structure by inlet temperature measurement in order to manipulate furnace heat input.

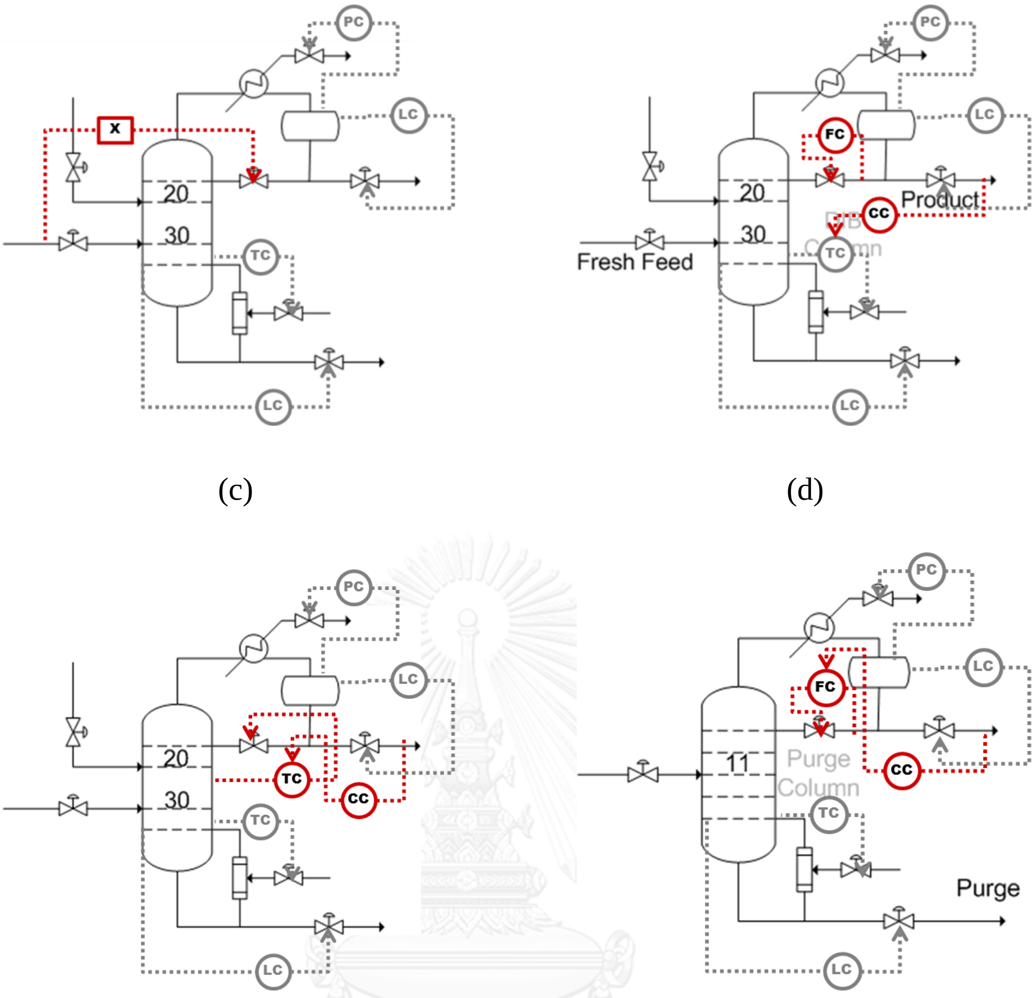
- Figure 5.32 (b) was shown the fixed outlet temperature control structure. Similar to 5.32 (a), but outlet temperature measurement instant.
- Figure 5.32 (c) was shown the fixed outlet composition by cascade inlet temperature control structure. Similar to 5.32(a), in order to control the stable output quantity, adding the measure composition of reactor outlet to manipulating the setpoint of temperature loop.

For reactor testing, after the various disturbances, the output responses were compared to find the best result control structure demonstrated by graphing furnace inlet temperature, furnace heat input, reactor outlet flow rate, reactor inlet temperature, reactor outlet temperature and reactor outlet composition of the product.

Control structure of column



(a) (b)



(c) (d)

(e) (f)

Figure 5.33 The control structure for handle disturbances in temperature section

(a) Fixed reflux flow rate (R) control structure

(b) Fixed reflux ratio (R/D) control structure

(c) Fixed reflux to feed ratio (R/F) control structure

(d) Fixed reflux flow rate with composition cascade to temperature loop (R-VCC) control structure

(e) Dual temperature with composition cascade (RCC-V) control structure

(f) Fixed reflux flow rate with composition cascade to flow loop (RCC) control structure

The separation section is comparably important as the reaction sections. It also had 4 chosen control structure for testing and analyzing.

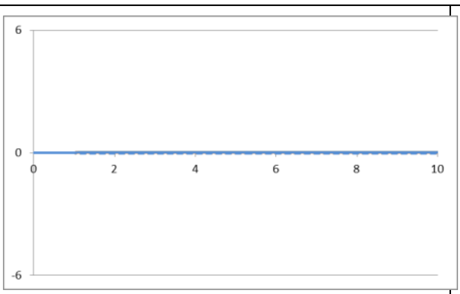
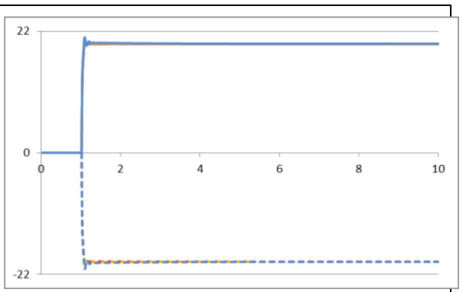
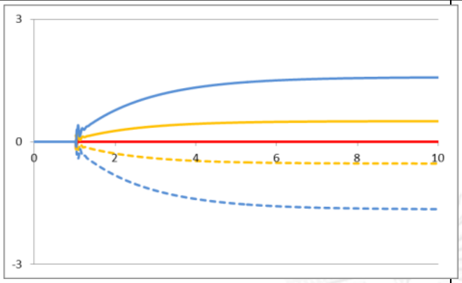
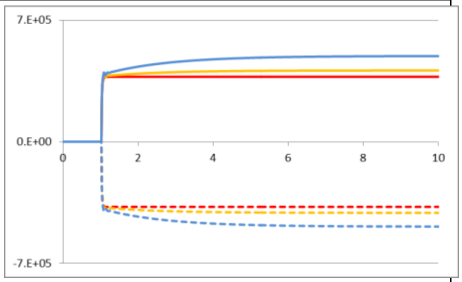
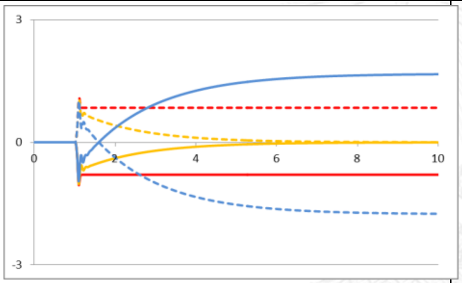
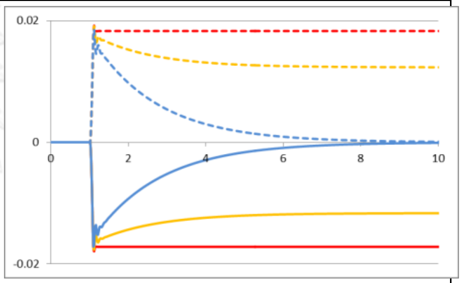
- Figure 5.33 (a) was shown fixed reflux flow control structure which was the fixing for stable reflux flow.
- Figure 5.33 (b) was shown a fixed reflux ratio (ratio of reflux flow to distillate flow) control structure which was the manipulating reflux flow rate depends on distillate rate but in a constant ratio.
- Figure 5.33 (b) was shown fixed reflux to feed ratio control structure which was the manipulating reflux flow rate depends on column feed flow rate but in a constant ratio.
- Figure 5.33 (d) was shown the fixed outlet composition by cascade the temperature control structure. Similar to 5.33 (a), in order to control the stable output quantity, adding the measure composition of distillate to manipulating the setpoint of temperature loop.
- Figure 5.33 (e) was shown the fixed outlet composition by cascade the dual-end temperature control structure by using sensitive tray resulted from step 2.6. If the using stage is reflux drum nearer, the measured temperature in that tray will be used to manipulate the reflux flow rate. In the other hand, if the using stage is the reboiler nearer, the measured temperature in that tray will be used to manipulate reboiler heat input. In order to control the stable output quantity, adding the measure composition of distillate to manipulating the setpoint of temperature loop.
- Figure 5.33 (f) was shown the fixed outlet composition by cascade the fixed reflux flow control structure. Similar to 5.33 (a), in order to control the stable output quantity, adding the measure composition of distillate to manipulating the setpoint of flow loop.

For the column testing, after the fed were changed, the output responses were compared to find the best result structure demonstrated by graphing distillate flow rate, distillate composition of the light key component, bottom flow rate, bottom composition of the heavy key component, condenser heat removal and reboiler heat input.

4.2.1 Reactor

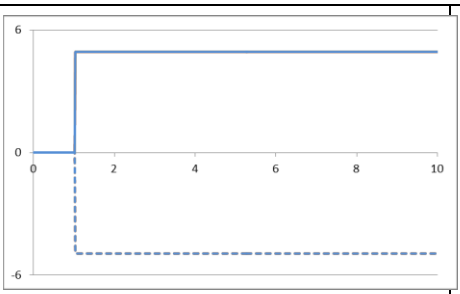
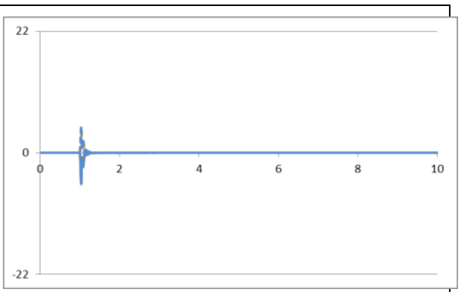
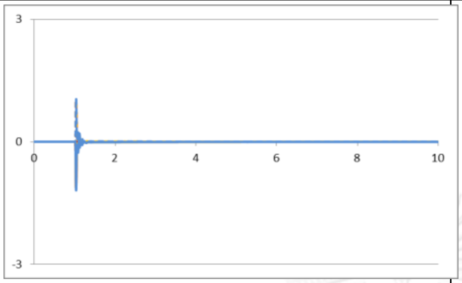
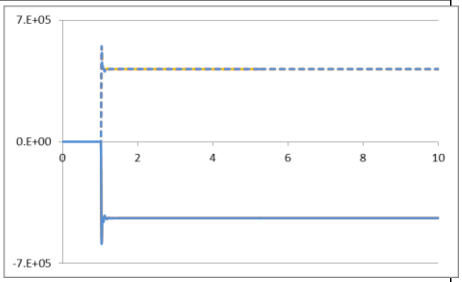
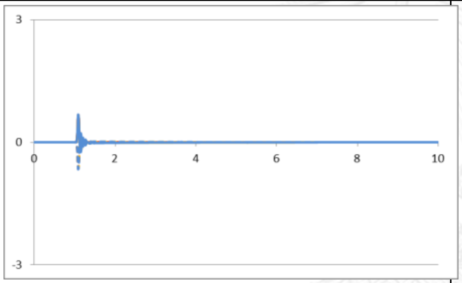
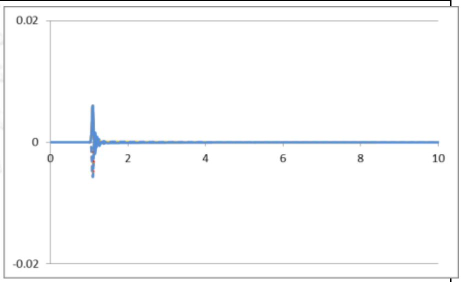
This part consists the test of reactor, which is emphasizing the amount of product. The method of stimulating at steady state concentrates on a large amount of product, based on its possibility. On the contrary, dynamic stimulation focuses on yielding product consistently. It can be adjusted to the setpoint right when there is any fluctuation. The disturbances test part includes total feed flow rate, feed temperature, reactant component feed composition and reactant component feed flow rates. All control structures are illustrated in Figure 5.32 and demonstrating as shown in Table 5.6-5.9.

Table 5.6 The dynamic responses for total feed flow rate disturbances of reactor.

Furnace feed temperature (K)		Reactor outlet flow (kmol/h)	
Reactor feed temperature (K)		Furnace duty (kJ/h)	
Reactor outlet temperature (K)		Reactor outlet composition (*i*C_4)	
	Time (h)		Time (h)

— Feed +5% IT	- - - Feed -5% IT
— Feed +5% TCT	- - - Feed -5% TCT
— Feed +5% CCT	- - - Feed -5% CCT

Table 5.7 The dynamic responses for temperature disturbances of reactor.

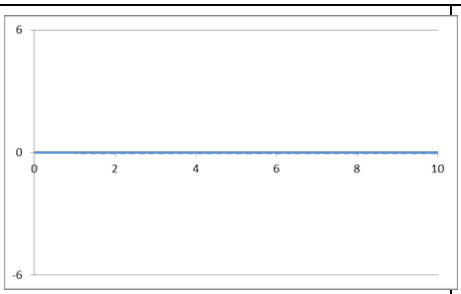
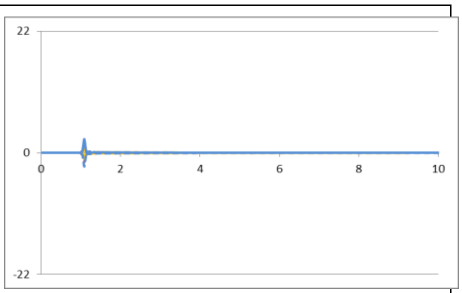
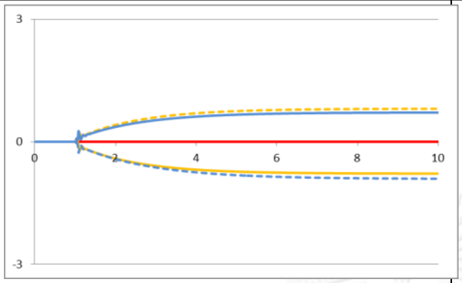
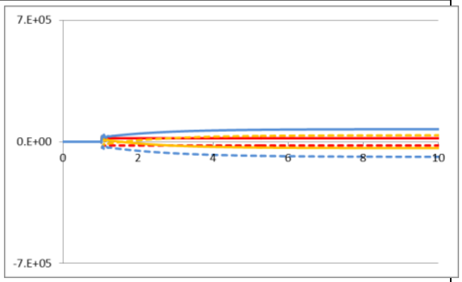
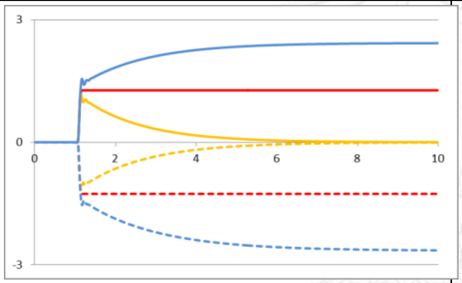
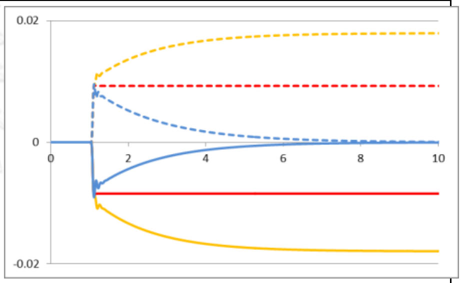
Furnace feed temperature (K)		Reactor outlet flow (kmol/h)	
Reactor feed temperature (K)		Furnace duty (kJ/h)	
Reactor outlet temperature (K)		Reactor outlet composition (iC_4)	
	Time (h)		Time (h)

━━ Temp +5% IT　　- - - Temp -5% IT

━━ Temp +5% TCT　　- - - Temp -5% TCT

━━ Temp +5% CCT　　- - - Temp -5% CCT

Table 5.8 The dynamic responses for composition disturbances of reactor.

Furnace feed temperature (K)		Reactor outlet flow (kmol/h)	
Reactor feed temperature (K)		Furnace duty (kJ/h)	
Reactor outlet temperature (K)		Reactor outlet composition (*i*C_4)	
	Time (h)		Time (h)

— Reactant mole faction +5% IT

- - - Reactant mole faction -5% IT

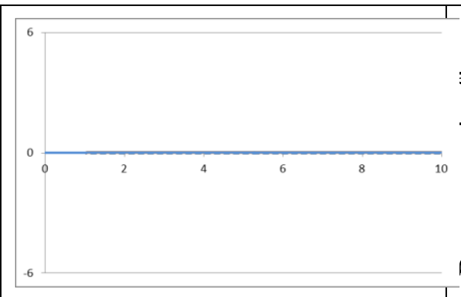
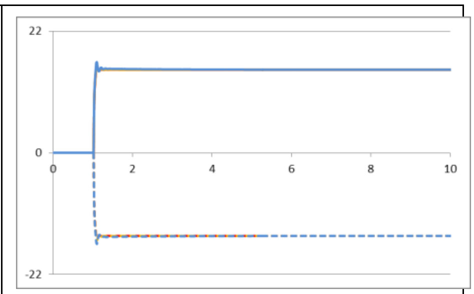
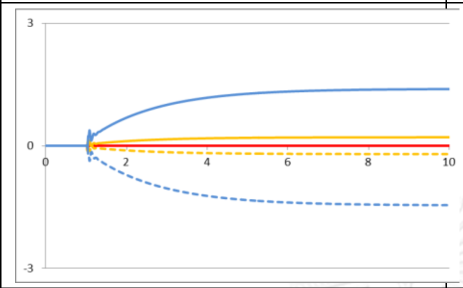
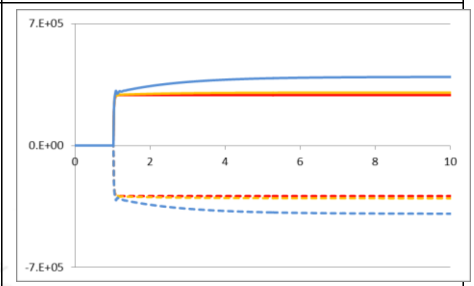
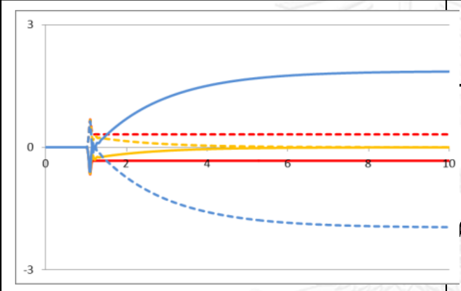
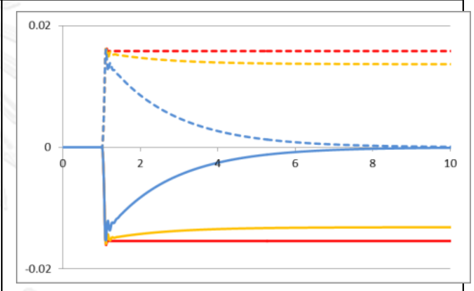
— Reactant mole faction +5% TCT

- - - Reactant mole faction -5% TCT

— Reactant mole faction +5% CCT

- - - Reactant mole faction -5% CCT

Table 5.9 The dynamic responses for component flow rate disturbances of reactor.

Furnace feed temperature (K)		Reactor outlet flow (kmol/h)	
Reactor feed temperature (K)		Furnace duty (kJ/h)	
Reactor outlet temperature (K)		Reactor outlet composition (iC_4)	
	Time (h)		Time (h)

━━ Reactant component flow +5% IT ‒ ‒ ‒ Reactant component flow -5% IT

━━ Reactant component flow +5% TCT ‒ ‒ ‒ Reactant component flow -5% TCT

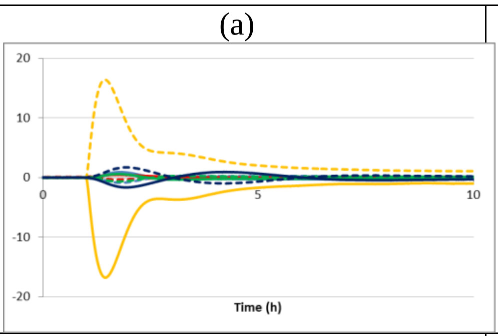
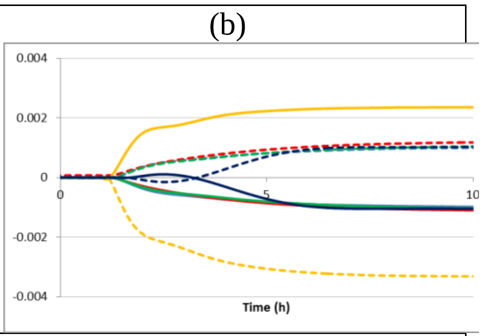
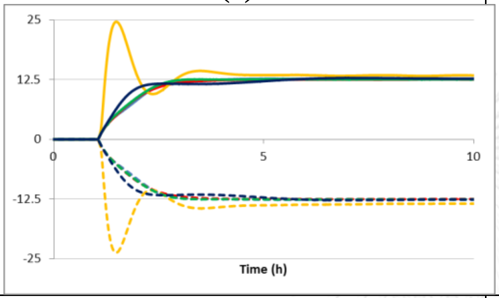
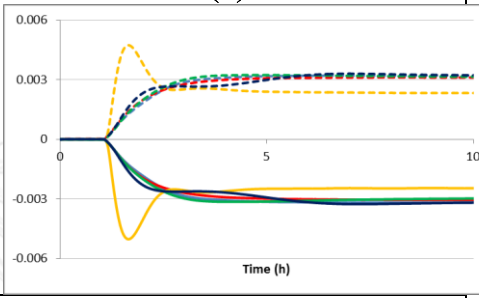
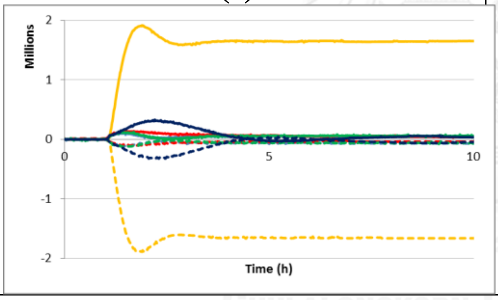
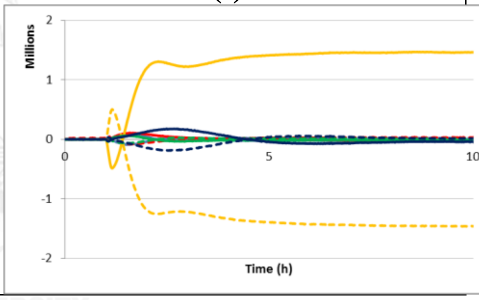
━━ Reactant component flow +5% CCT ‒ ‒ ‒ Reactant component flow -5% CCT

Temperature testing is not able to separate control structures, as shown in Table 5.7. It was mentioned that the control structure can be designated the quantity of yields products. Considering the flow rate along with outlet composition, the graphs describe how the flow rate is constant in every structure. The finest control structure that stabilizes a value of the outlet composition is called “fixed outlet composition by cascade to inlet temperature”, shown in Figure 5.32(c). There is an outlet temperature variation at the chosen structure’s reactor; however, the stable temperature can be varied over an acceptable range (within ±3°C). Besides, energy consumptions of each particular structure are insignificantly different.

4.2.2 Column C1.

The test of column C1 points out the volume of products, which consistently flow out at distillate stream over time. It is purposely managed that the particular setpoint can be retrieved as soon as possible when any disturbance occurs. The light key component is iC_4, and the heavy key component is nC_4. The disturbances test part includes fresh feed flow rate, feed temperature, light key component feed composition, light component feed flow rate and heavy key component feed flow rate. All control structures are illustrated in Figure 5.33, which shown the improvement in handling disturbances achieved when single-end (fixed reflux flow rate, fixed reflux ratio and fixed reflux to feed ratio) and dual-end control (simple dual temperature and dual temperature with composition cascade) structures is used in column C1 and demonstrated by distillate flow rate (D1), distillate composition of light key component (xD1), bottom flow rate (B1), bottom composition of heavy key component (yB1), condenser heat removal (Qc) and reboiler heat input (Qr) as shown in Table 5.10-5.14.

Table 5.10 The dynamic responses for fresh feed flow rate changes, plus 259.5 and minus 234.7 kmol/h of column C1.

Distillate flow	(a)  	iC_4 (LK) in Distillate	(b) 
Bottom flow	(c)  	nC_4 (HK) in bottom	(d) 
Qc (kJ/h)	(e)  	Qr (kJ/h)	(f) 

━ Feed +5% R	╍ Feed -5% R
━ Feed +5% R/D	╍ Feed -5% R/D
━ Feed +5% R/F	╍ Feed -5% R/F
━ Feed +5% R-VCC	╍ Feed -5% R-VCC
━ Feed +5% RCC-V	╍ Feed -5% RCC-V

Table 5.11 The dynamic responses for temperature changes, plus 37.21°C and minus 27.21°C of column C1.

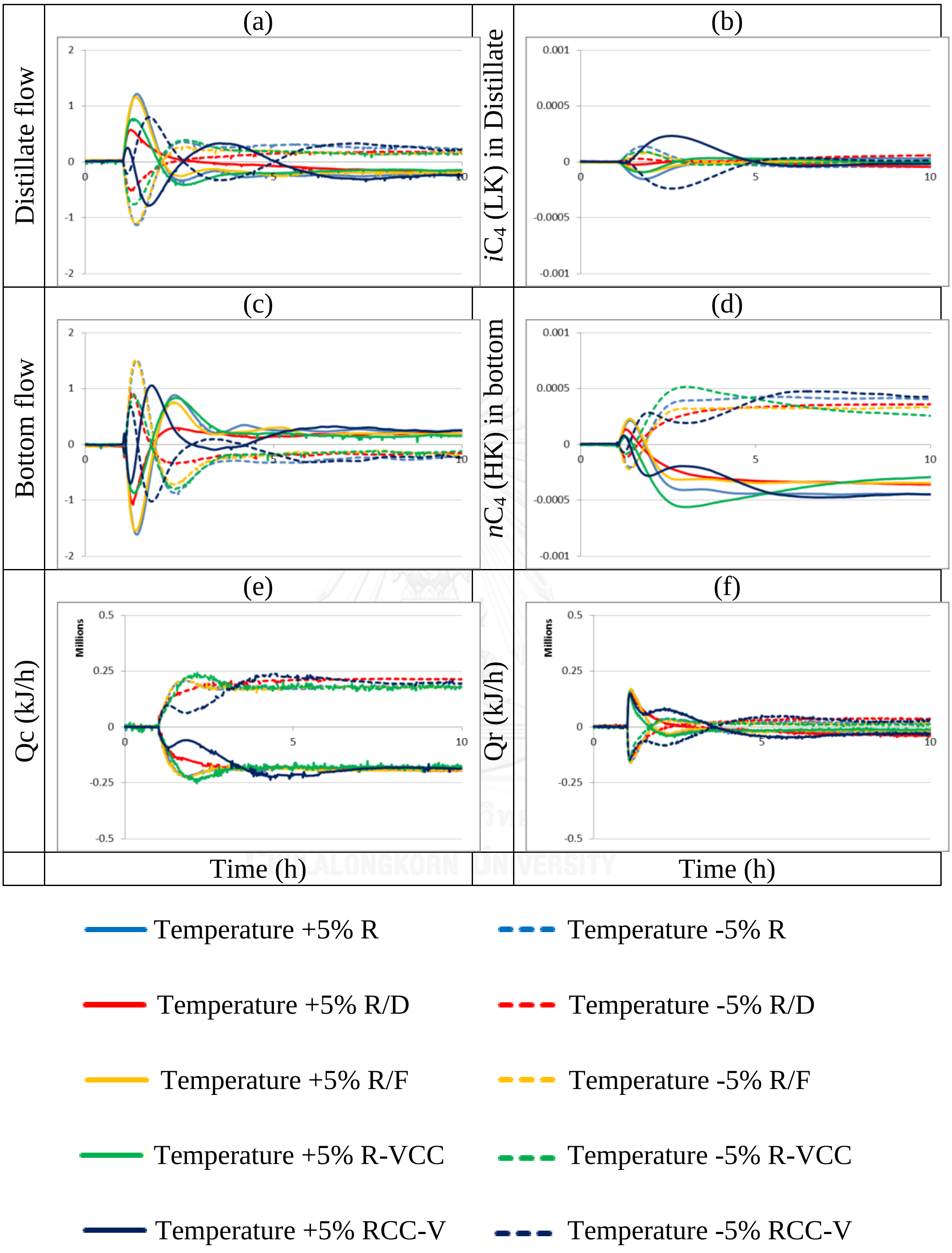
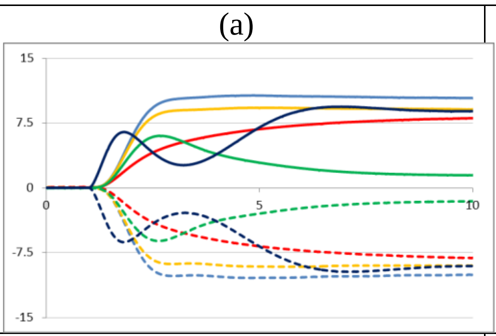
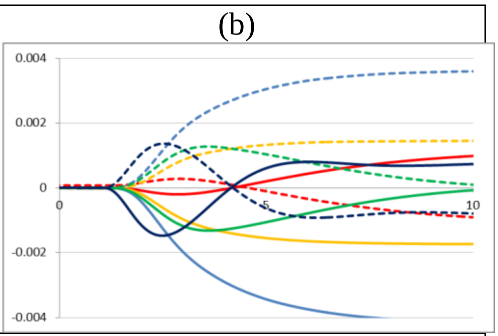
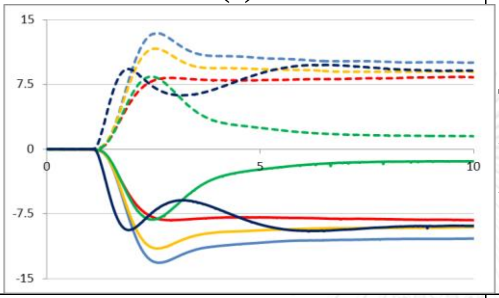
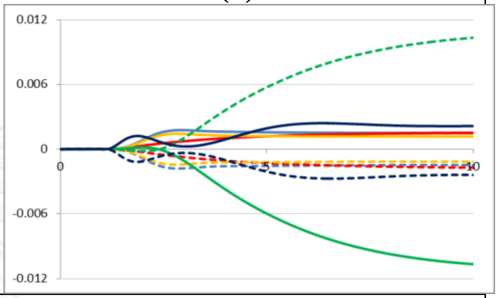
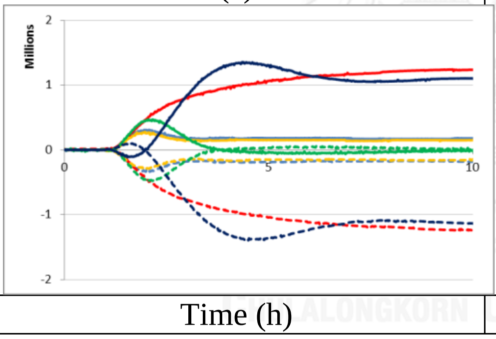
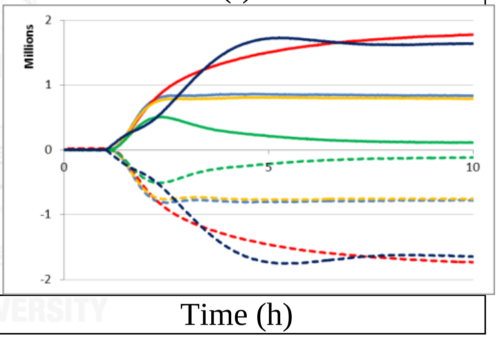
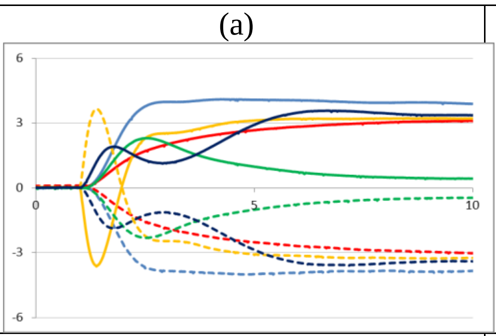
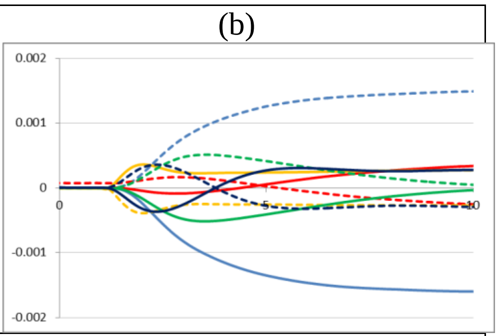
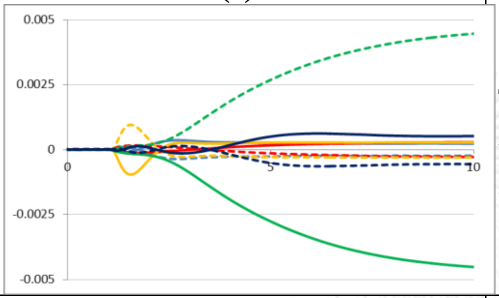
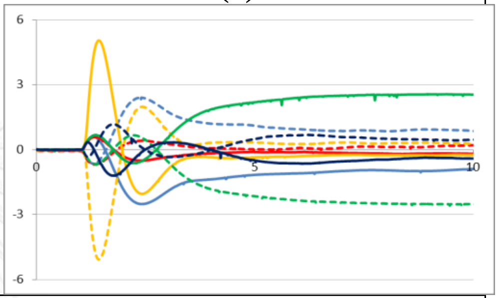
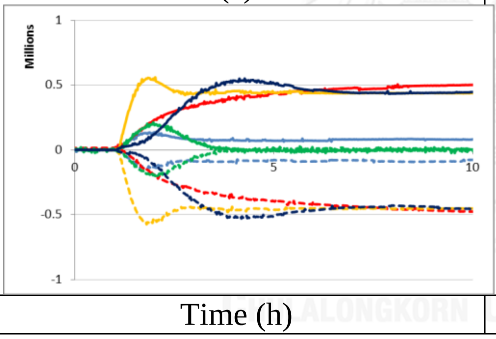
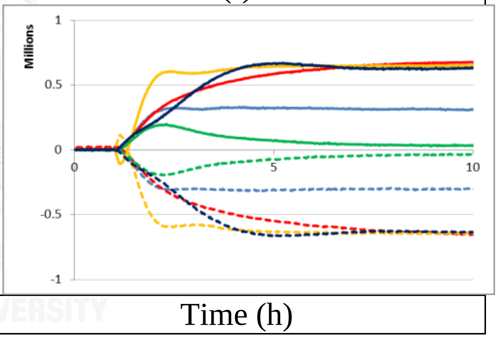


Table 5.12 The dynamic responses for LK composition changes, plus iC_4 0.2635 and minus iC_4 0.2165 of column C1.

Distillate flow	(a) 	iC_4 (LK) in Distillate	(b) 
Bottom flow	(c) 	nC_4 (HK) in bottom	(d) 
Qc (kJ/h)	(e) 	Qr (kJ/h)	(f) 
	Time (h)		Time (h)

━ LK mole faction +5% R	┅ LK mole faction -5% R
━ LK mole faction +5% R/D	┅ LK mole faction -5% R/D
━ LK mole faction +5% R/F	┅ LK mole faction -5% R/F
━ LK mole faction +5% R-VCC	┅ LK mole faction -5% R-VCC
━ LK mole faction +5% RCC-V	┅ LK mole faction -5% RCC-V

Table 5.13 The dynamic responses for LK component flow rate changes, plus iC$_4$ 62.265 and minus iC$_4$ 56.340 kmol/h of column C1.

Distillate flow	(a)  	iC$_4$ (LK) in Distillate	(b) 
Bottom flow	(c)  	nC$_4$ (HK) in bottom	(d) 
Qc (kJ/h)	(e)  	Qr (kJ/h)	(f) 
	Time (h)		Time (h)

━ LK component flow +5% R ╍ LK component flow -5% R

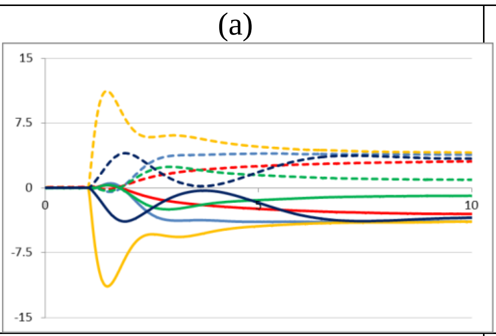
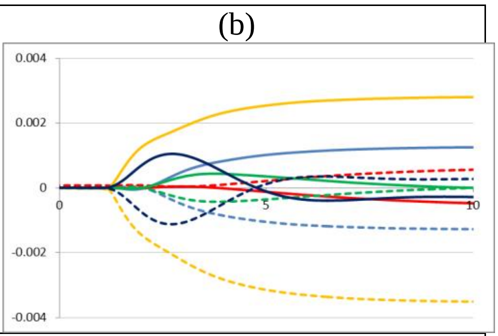
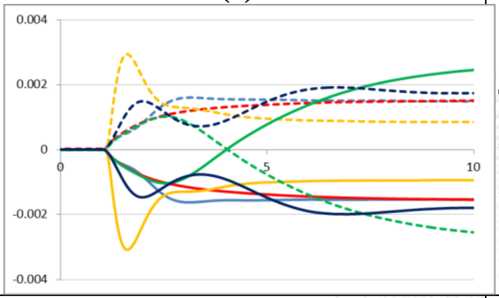
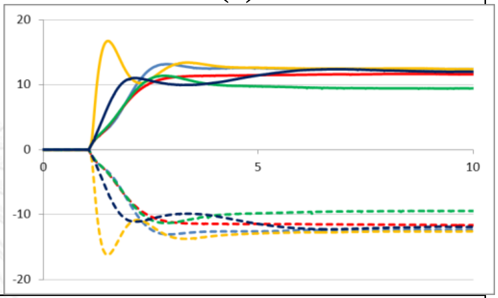
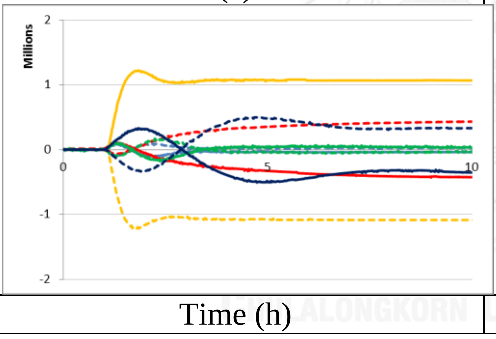
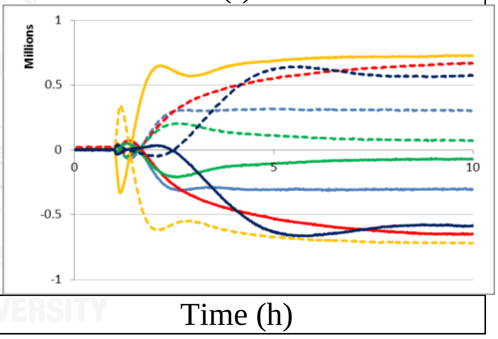
━ LK component flow +5% R/D ╍ LK component flow -5% R/D

━ LK component flow +5% R/F ╍ LK component flow -5% R/F

━ LK component flow +5% R-VCC ╍ LK component flow -5% R-VCC

━ LK component flow +5% RCC-V ╍ LK component flow -5% RCC-V

Table 5.14 The dynamic responses for HK component flow rate changes, plus nC_4 179 and minus nC_4 161.97 kmol/h of column C1 for CS1.

Distillate flow	(a)  	iC_4 (LK) in Distillate	(b) 
Bottom flow	(c)  	nC_4 (HK) in bottom	(d) 
Qc (kJ/h)	(e)  	Qr (kJ/h)	(f) 
	Time (h)		Time (h)

HK component flow +5% R

HK component flow -5% R

HK component flow +5% R/D

HK component flow -5% R/D

HK component flow +5% RF

HK component flow -5% RF

HK component flow +5% R-VCC

HK component flow -5% R-VCC

HK component flow +5% RCC-V

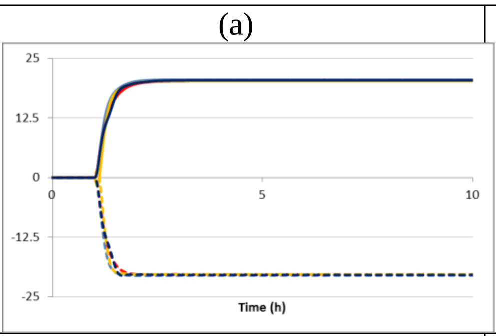
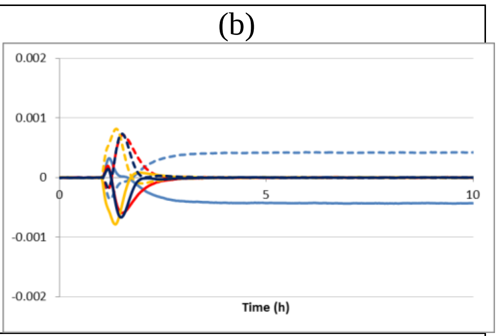
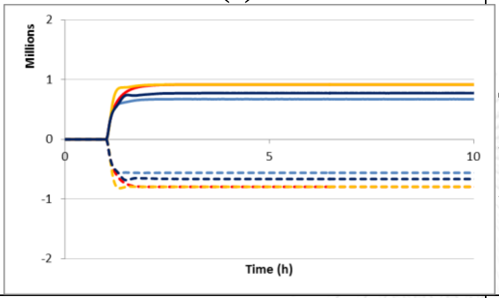
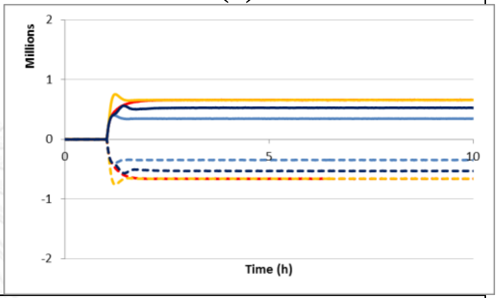
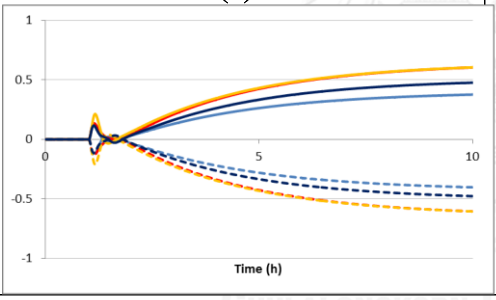
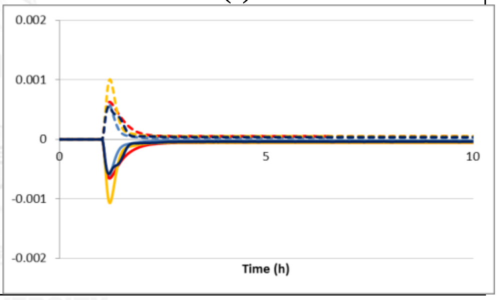
HK component flow -5% RCC-V

Due to the change in fresh feed flow, R, R/D, R-VCC and RCC-V structure will become choices for the flow test. R-VCC is the best one because increasing R/F production needs more energy. The other structures use nearly the same amount of energy as before disturbance appear. From the change in temperature, it is seen that iC_4 composition in distillate of RCC-V structure change in the highest rate compared to those in the other structure which affects the product quality but all of control structure can be nearly setpoint. From the change in LK composition and LK component flow, LK will be emitted from the top of the distillate only in R/D and RCC-V structure (in R/F structure, LK is emitted only in the LK component test) but RC can be closer to set point more than other structure. Owing to the changes in HK component, HK will be emitted from the bottom flow only in R-VCC structure. From the above reasons, the proper structure decided from the result of all the five tests is R-VCC for C1.

4.2.3 Column C2.

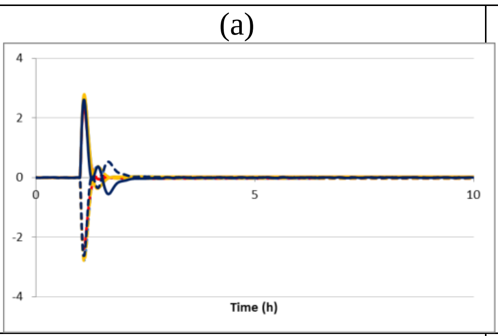
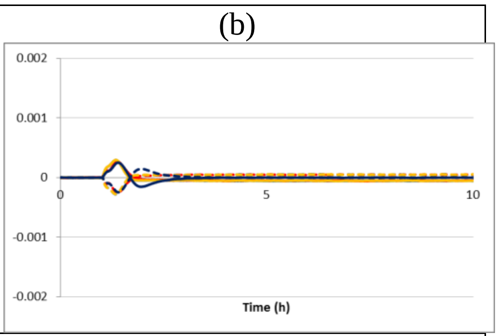
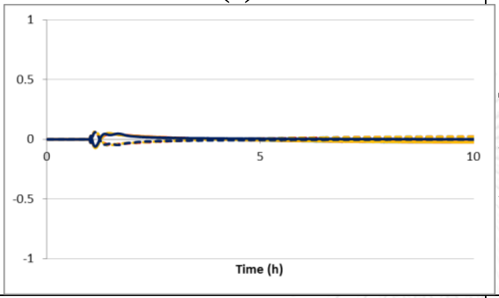
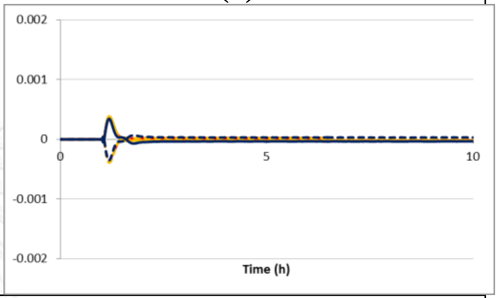
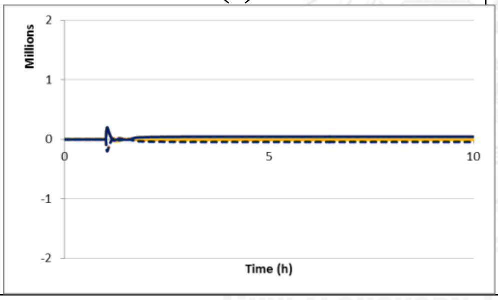
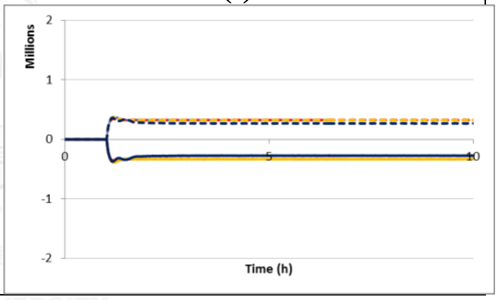
The test of column C2 points out the volume of reactant, which consistently flow out at distillate stream over time. It is purposely managed that the particular setpoint can be retrieved as soon as possible when any disturbance occurs. The light key component is nC_4, and the heavy key component is C_3 impurity. The disturbances test part as same as column C1. All control structures are given in Figure 5.33, which shown the improvement in handling disturbances achieved. Single-end (fixed reflux flow rate, fixed reflux ratio, fixed reflux to feed ratio and fixed reflux flow rate with composition cascade) control structures only are used in column C2 and demonstrated by distillate flow rate (D2), distillate composition of light key component (xD2), bottom flow rate (B2), bottom composition of heavy key component (yB2), condenser heat removal (Qc) and reboiler heat input (Qr) as shown in table 5.15-5.19.

Table 5.15 The dynamic responses for total feed flow rate changes, plus 438.0 and minus 396.2 kmol/h of column C2.

Distillate flow	(a)  	nC_4 (LK) in Distillate	(b) 
Bottom flow	(c)  	iC_5 (HK) in bottom	(d) 
Qc (kJ/h)	(e)  	Qr (kJ/h)	(f) 

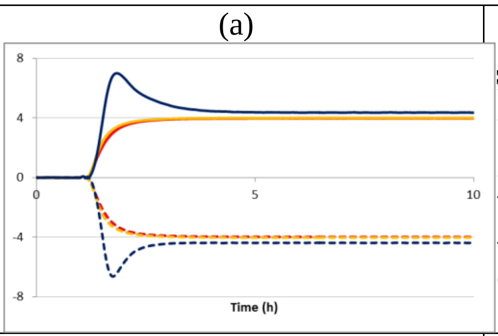
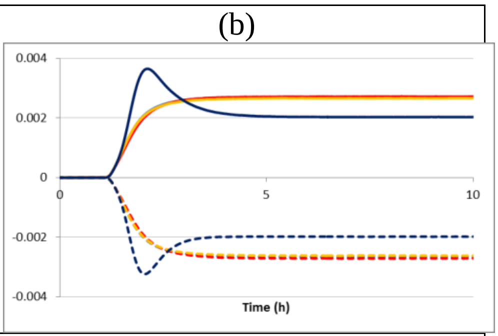
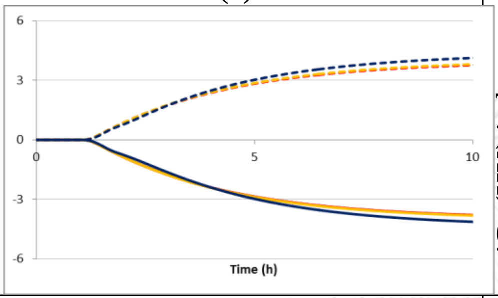
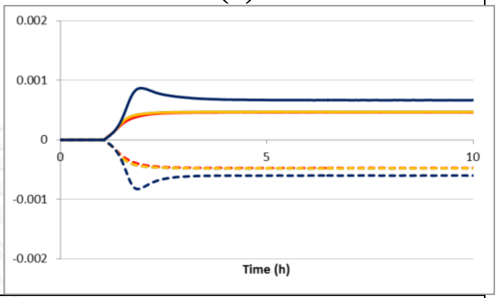
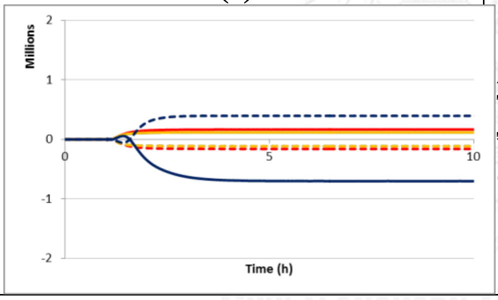
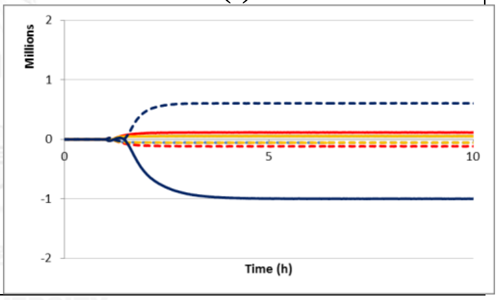
— Feed +5% R	- - - Feed -5% R
— Feed +5% R/D	- - - Feed -5% R/D
— Feed +5% RF	- - - Feed -5% RF
— Feed +5% RCC	- - - Feed -5% RCC

Table 5.16 The dynamic responses for temperature changes, plus 37.21°C and minus 27.21°C of column C2.

Distillate flow	(a)  	*n*C$_4$ (LK) in Distillate	(b) 
Bottom flow	(c)  	*i*C$_5$ (HK) in bottom	(d) 
Qc (kJ/h)	(e)  	Qr (kJ/h)	(f) 

— Temperature +5% R	- - - Temperature -5% R
— Temperature +5% R/D	- - - Temperature -5% R/D
— Temperature +5% RF	- - - Temperature -5% RF
— Temperature +5% RCC	- - - Temperature -5% RCC

Table 5.17 The dynamic responses for LK composition changes, plus nC_4 0.7941 and minus nC_4 of column C2.

Distillate flow	(a)  	nC_4 (LK) in Distillate	(b) 
Bottom flow	(c)  	iC_5 (HK) in bottom	(d) 
Qc (kJ/h)	(e)  	Qr (kJ/h)	(f) 

— LK mole faction +5% R

- - - LK mole faction -5% R

— LK mole faction +5% R/D

- - - LK mole faction -5% R/D

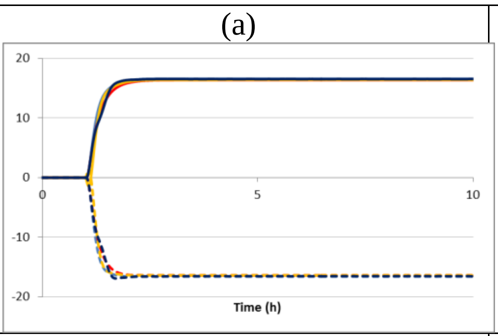
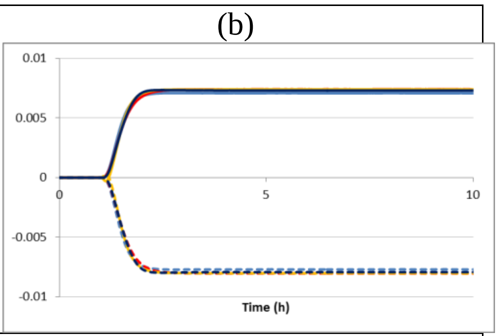
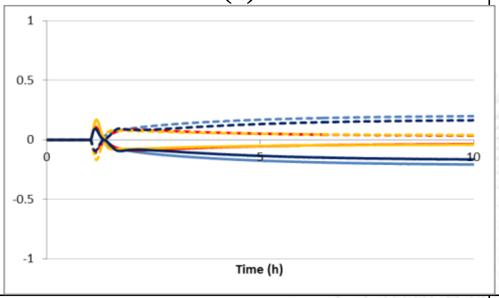
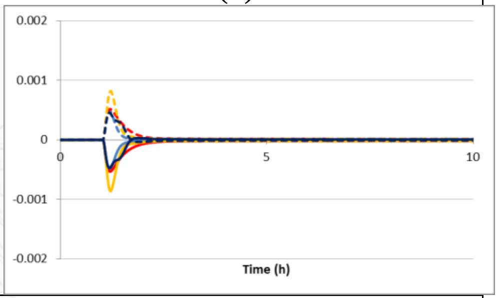
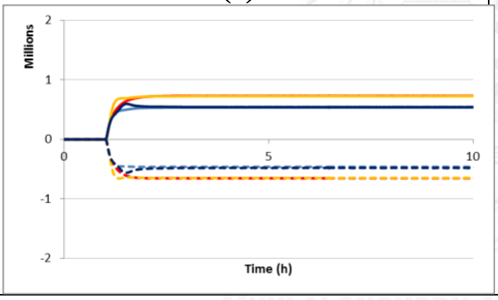
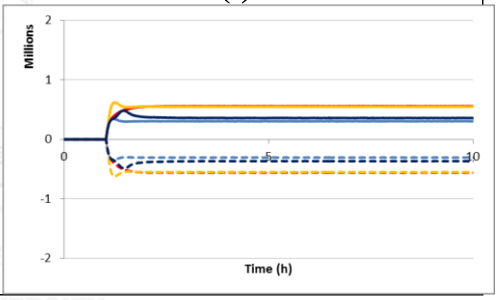
— LK mole faction +5% RF

- - - LK mole faction -5% RF

— LK mole faction +5% RCC

- - - LK mole faction -5% RCC

Table 5.18 The dynamic responses for LK component flow rate changes, plus nC_4 343.42 and minus nC_4 310.72 kmol/h of column C2.

Distillate flow	(a)  	nC_4 (LK) in Distillate	(b) 
Bottom flow	(c)  	iC_5 (HK) in bottom	(d) 
Qc (kJ/h)	(e)  	Qr (kJ/h)	(f) 

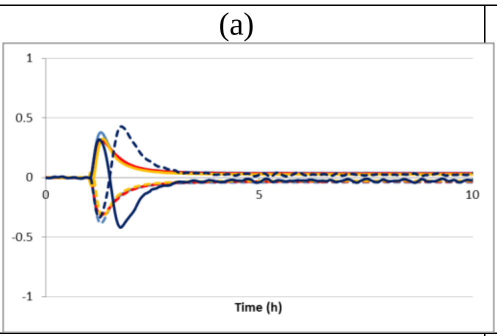
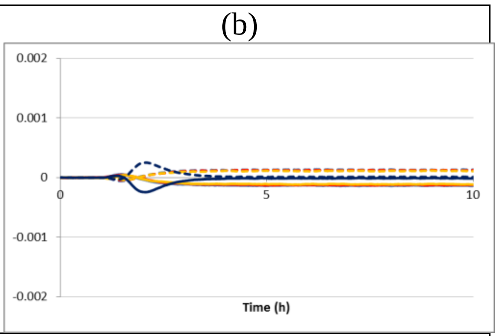
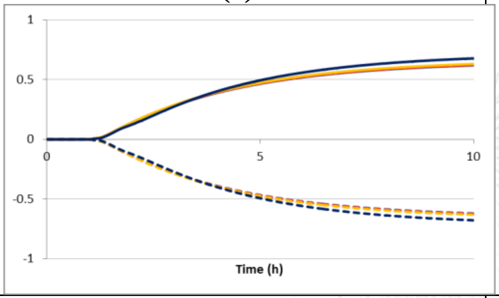
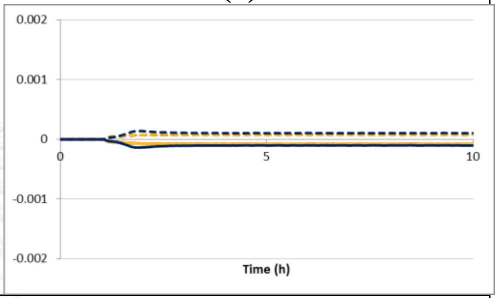
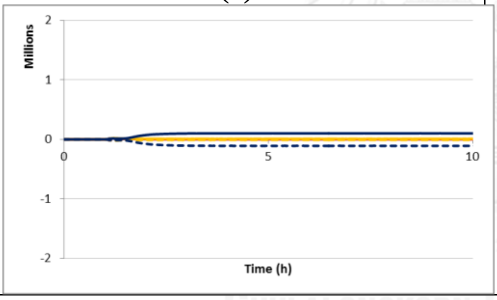
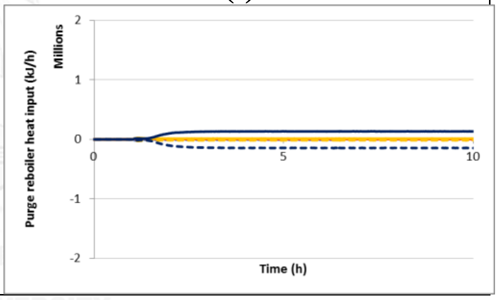
— LK component flow +5% R
- - - LK component flow -5% R

— LK component flow +5% R/D
- - - LK component flow -5% R/D

— LK component flow +5% RF
- - - LK component flow -5% RF

— LK component flow +5% RCC
- - - LK component flow -5% RCC

Table 5.19 The dynamic responses for HK component flow rate changes, plus $iC_5$14.537 and minus $iC_5$13.153 kmol/h of column C2.

Distillate flow	(a) 	nC_4 (LK) in Distillate	(b) 
Bottom flow	(c) 	iC_5 (HK) in bottom	(d) 
Qc (kJ/h)	(e) 	Qr (kJ/h)	(f) 

━ HK component flow +5% R	╍ HK component flow -5% R
━ HK component flow +5% R/D	╍ HK component flow -5% R/D
━ HK component flow +5% RF	╍ HK component flow -5% RF
━ HK component flow +5% RCC	╍ HK component flow -5% RCC

Due to the change in fresh feed flow, R will not become choice because *n*C_4 component cannot change to set point. R/F, RCC and R/D can change respectively. From the change in temperature and LK component, the structures are all same. From the change in LK composition, offset of composition in distillate of RCC is less than others and offset of composition in bottom of R/F is less than others. Owing to the change in HK component, composition in distillate of RCC can be changed to set point. From the above reasons, the proper structure decided from the result of all the five tests is RCC for C2.

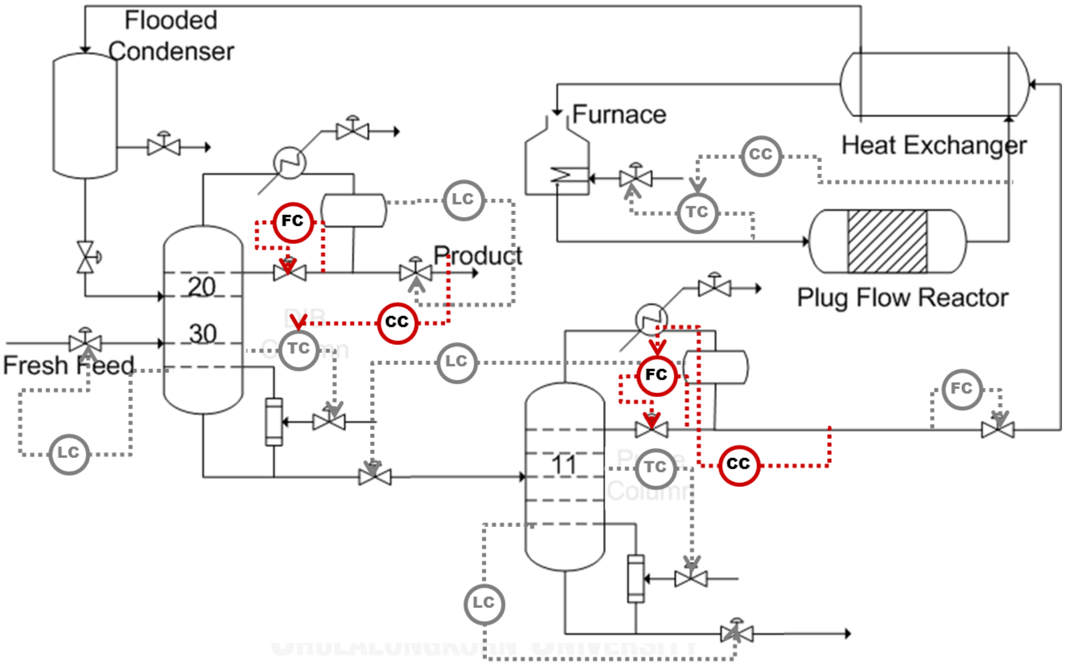


Figure 5.34 Flowsheet of butane isomerization process for handle the material disturbances.

Stage 3. Unit Level Design.

Step 5: Design the rest of the control loop.

This step is designed to the control loops for safety and inventory controls. We can assign control loops within individual units. The rest control loops are designed and added to the process, which includes pressure and level controls. Quantifiers of each component are different location. A controller must be presented in all liquid accumulates.

The assigned control loops were depending on the number of remaining necessary for each unit. From step 2.1, 10 controllers were already installed, as the in the summation of column DOF shown in Table 5.20 controllers will be needed to be to complete the control structure.

Table 5.20 Remaining control variable

Unit	DOF	Have	Add
Independent stream	1	1	-
Furnace	1	1	-
Condenser	2	0	2
DIB column	5	4	1
Purge column	5	4	1
	14	10	4

One liquid level need to be controlled: flooded condenser. Three pressure control loops: flooded condenser, column C1 and C2. The various loops shown in Figure 5.35 are listed below with their controlled and manipulated variables.

- Flooded condenser level (LC) is controlled by manipulating the liquid streams leaving the condenser.

- Pressure control loops: Three pressure controllers (PC) must are installed in the condensers. In flooded condenser and both condenser of columns, pressure controls can be achieved by manipulating heat removal at the condenser.

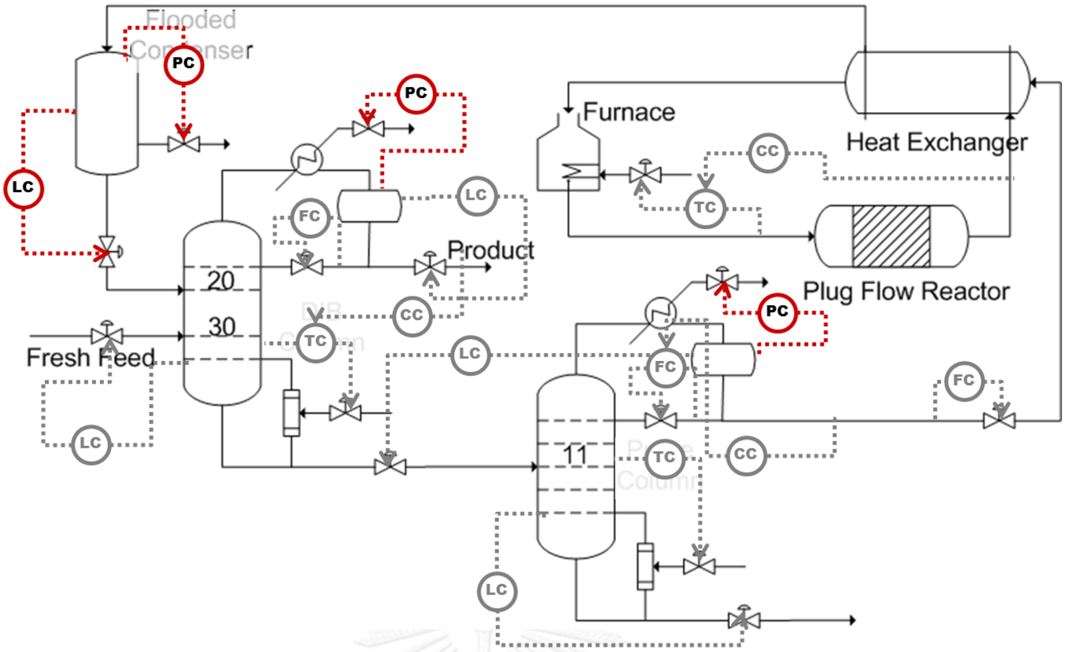


Figure 5.35 Flowsheet of butane isomerization process for the rest of the control variables.

Stage 4. Enhanced control design.

Step 6: Adding enhanced control.

There are one enhanced loop added in Step 3.5 and two enhanced loops in Step 4.2. They are the composition cascade loops at the reactor, C1 and C2 columns.

Stage 5. Energy Management and Optimization.

Step 7: Save energy via heat exchanger networks.

The energy recovery is already made in the original design of the feed-effluent heat exchanger (FEHE) to exchange heat between the reactor effluent and the reactor feed. The FEHE exchange the heat of the reactor effluent and preheat the reactor feed. The recovery heat load is 5,586 MJ/h. The cool side temperature difference is 91°C. Still, there is a room for improvement. We can reduce the temperature different to 30°C, practical minimum temperature differential approach. The temperature of the preheated reactor feed must less than 136.6°C, its vaporization temperature. Again, the temperature of the reactor effluent must not be cooled below 130°C, its condensation temperature. The current temperature of the cooled reactor effluent is 139°C, so we do not further improving the design.

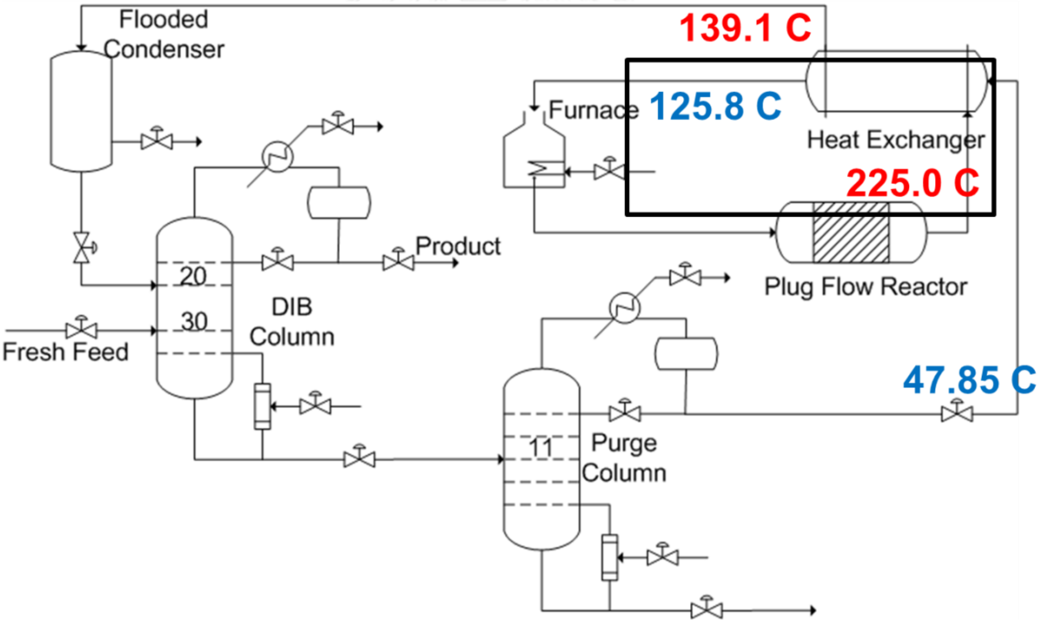


Figure 5.36 Flowsheet of butane isomerization process for energy management.

Step 8: Optimize for economics and/or improving control performance.

This step is omitted due to the scope of the Thesis.

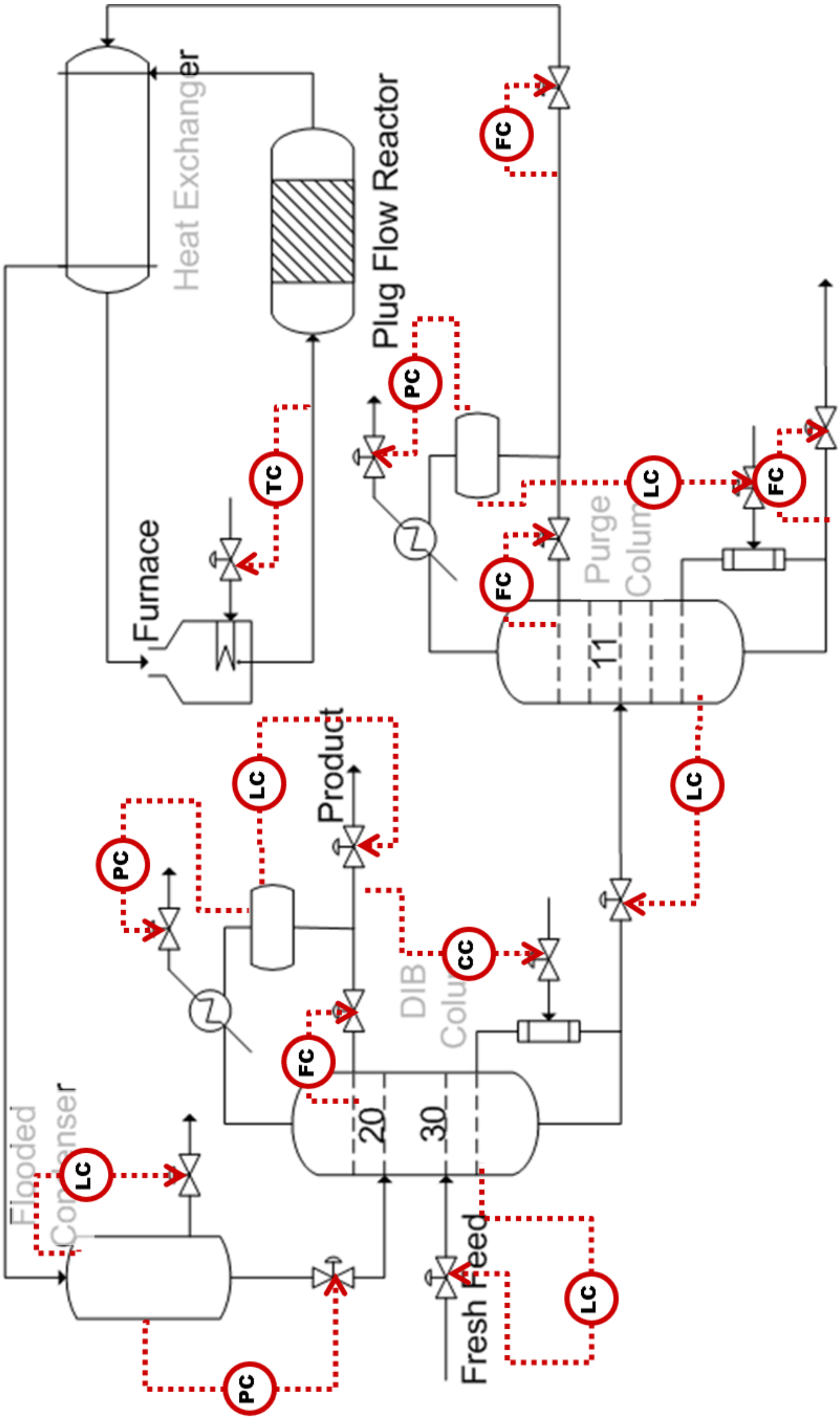


Figure 5.37 butane isomerization plantwide control structure BC

Table 5.21 Control structure lists of BC

Equipment	**Controller**	**Controlled Variables**	**Manipulated Variables**	**Type**	**Action**
Independent streams	FC01	Flow rate out of D2	Flow rate out of D2	PI	Reverse
Column C1	FC11	Reflux flow rate of C1	Reflux flow rate of C1	PI	Reverse
	CC11	$n\mathrm{C}_4$ mole fraction	Reboiler heat input of C1	PID	Reverse
	LC11	Reflux drum level of C1	Flow rate out of D1	P	Direct
	LC12	Bottom level of C1	Fresh feed of C1	P	Reverse
	PC11	Reflux drum pressure of C1	Condenser heat removal of C1	PI	Direct
Column C2	FC21	Reflux flow rate of C2	Reflux flow rate of C2	PI	Reverse
	FC22	Flow rate out of B2	Flow rate out of B2	PI	Reverse
	LC21	Reflux drum level of C2	Reboiler heat input of C2	P	Reverse
	LC22	Bottom level of C2	Column feed of C2	P	Reverse
	PC21	Reflux drum pressure of C2	Condenser heat removal of C2	PI	Direct
Reactor	TC31	Reactor inlet temperature	heat input of furnace	PID	Reverse
Flooded condenser	LC41	Condenser level	Condenser heat removal	P	Direct
	PC41	Condenser pressure	Flow rate out of flooded condenser	PI	Reverse

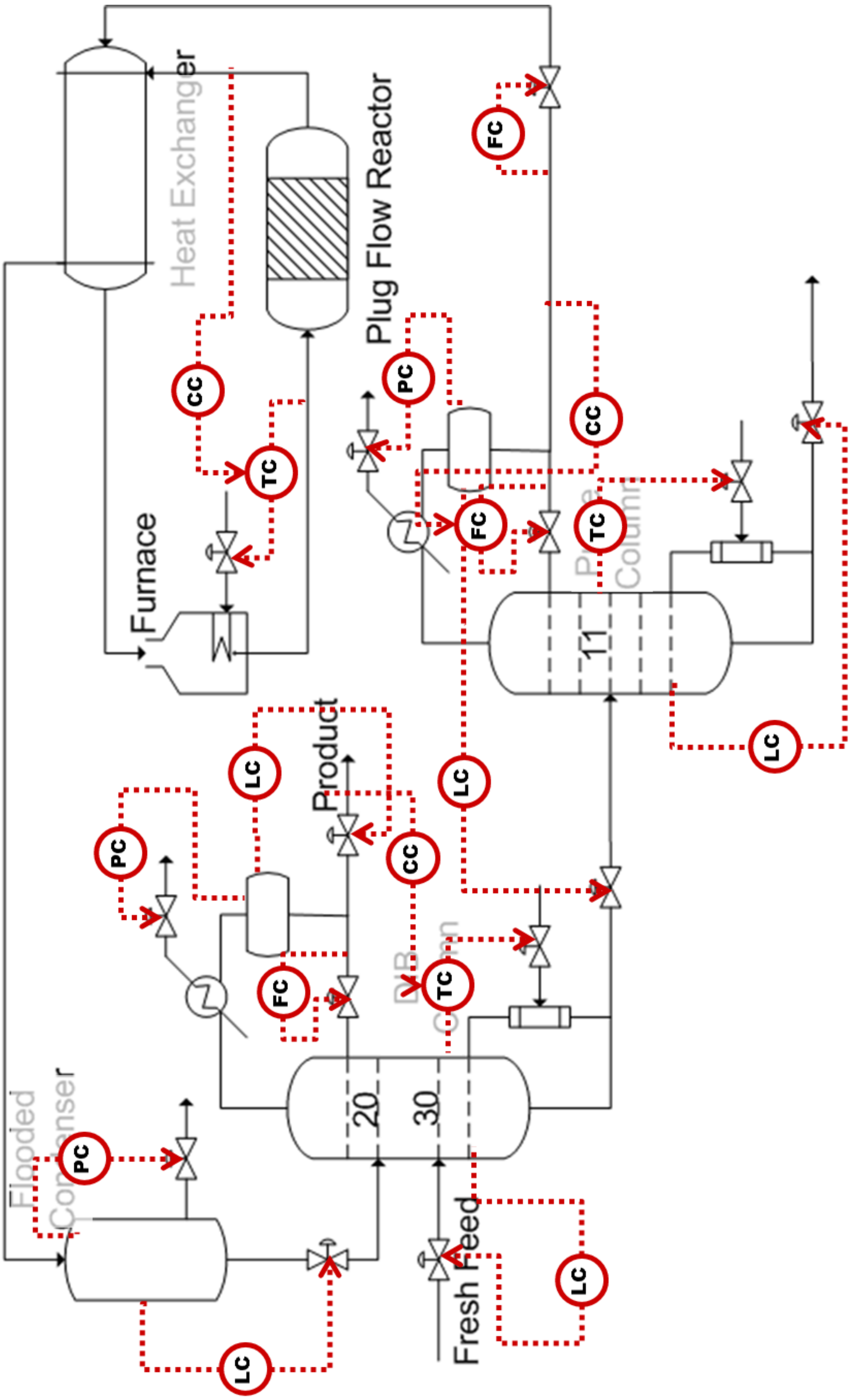


Figure 5.38 butane isomerization plantwide control structure CS1

Table 5.22 Control structure lists of CS1

Equipment	**Controller**	**Controlled Variables**	**Manipulated Variables**	**Type**	**Action**
Independent streams	FC01	Flow rate out of D2	Flow rate out of D2	PI	Reverse
Column C1	FC11	Reflux flow rate of C1	Reflux flow rate of C1	PI	Reverse
	TC11	Stage 44 C1 temperature	Reboiler heat input of C1	PID	Reverse
	LC11	Reflux drum level of C1	Flow rate out of D1	P	Direct
	LC12	Bottom level of C1	Fresh feed of C1	P	Reverse
	PC11	Reflux drum pressure of C1	Condenser heat removal of C1	PI	Direct
	CC11	Product composition	Set point of stage 44 C1 temperature loop	PID	Reverse
Column C2	FC21	Reflux flow rate of C2	Reflux flow rate of C2	PI	Reverse
	TC21	Stage 18 C1 temp.	Reboiler heat input of C2	PID	Reverse
	LC21	Reflux drum level of C1	Column feed of C2	P	Reverse
	LC22	Bottom level of C1	Flow rate out of B2	P	Direct
	PC21	Reflux drum pressure of C2	Condenser heat removal of C2	PI	Direct
	CC21	Recycle composition	Set point of fixed reflux flow loop	PID	Reverse
Reactor	TC31	Reactor inlet temperature	heat input of furnace	PID	Reverse
	CC32	Reactor outlet iC_4 mole fraction	Setpoint of reactor temperature loop	PID	Reverse
Flooded condenser	LC41	Condenser level	Flow rate out of flooded condenser	PI	Direct
	PC41	Condenser pressure	Condenser heat removal	PI	Reverse

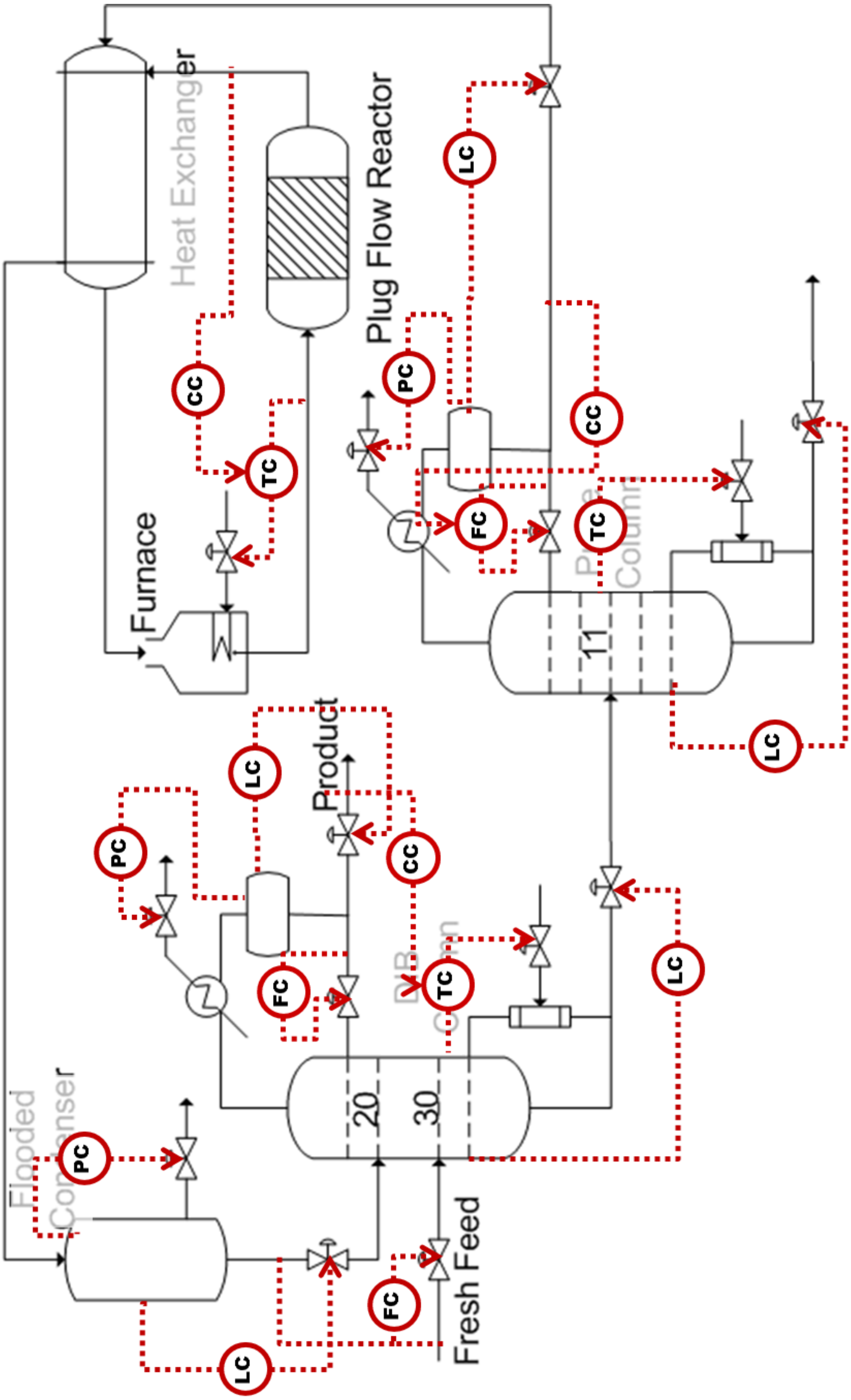


Figure 5.39 butane isomerization plantwide control structure CS2

Table 5.23 Control structure lists of CS2

Equipment	**Controller**	**Controlled Variables**	**Manipulated Variables**	**Type**	**Action**
Independent streams	FC01	Combined flow rate	Fresh feed flow rate	PI	Reverse
Column C1	FC11	Reflux flow rate of C1	Reflux flow rate of C1	PI	Reverse
	TC11	Stage 44 C1 temperature	Reboiler heat input of C1	PID	Reverse
	LC11	Reflux drum level of C1	Flow rate out of D1	P	Direct
	LC12	Bottom level of C1	Flow rate out of B1	P	Direct
	PC11	Reflux drum pressure of C1	Condenser heat removal of C1	PI	Direct
	CC11	Product composition	Set point of stage 44 C1 temperature loop	PID	Reverse
Column C2	FC21	Reflux flow rate of C2	Reflux flow rate of C2	PI	Reverse
	TC21	Stage 18 C1 temp.	Reboiler heat input of C2	PID	Reverse
	LC21	Reflux drum level of C1	Flow rate out of D2	P	Direct
	LC22	Bottom level of C1	Flow rate out of B2	P	Direct
	PC21	Reflux drum pressure of C2	Condenser heat removal of C2	PI	Direct
	CC21	Recycle composition	Set point of fixed reflux flow loop	PID	Reverse
Reactor	TC31	Reactor inlet temperature	heat input of furnace	PID	Reverse
	CC32	Reactor outlet iC4 mole fraction	Setpoint of reactor temperature loop	PID	Reverse
Flooded condenser	LC41	Condenser level	Flow rate out of flooded condenser	PI	Direct
	PC41	Condenser pressure	Condenser heat removal	PI	Reverse

Validate the designed control structures by rigorous dynamic simulation.

The simulation of this process is analyzed by using the commercial process simulator. All liquid level controllers and pressure loops are proportional control, liquid level with $K_C = 2$ for column base and reflux drum level loops and $K_C = 2$, $\tau_I = 10$ for pressure loops. Temperature loops contain 3 min deadtime. The product quality and safety loops are tuned by using a relay-feedback test to determine ultimate gains and periods. The Tyreus-Luyben tuning rules are implemented.

Table 5.24 Setpoint changes and disturbances

No.	**Type**	**Description**	**Change**
1	Setpoint changes	Fresh feed flow rate or Recycle flow rate	±15%
2	Disturbances	iC_5 feed composition	+0.02
		nC_4 feed composition	±10%
		Catalyst activity	-10%

5.3 Dynamic simulation results

To evaluate the efficiency, it is completed by regulating the process with disturbance and then determining the best plantwide control structure. The important plantwide disturbances in the butane isomerization process have 4 disturbances. Disturbance is fed into the system in order to test and assess the efficiency of control structure. In this step, the dynamic process will be run until it reaches the steady state of each structure. We have to adjust the fresh feed of each structure and fix it to nearly the other (about 250 kmol/h). Moreover, the product purity of each structure needs to be the same (each structure control have a composition control loop or cascade composition control loop for product purity) in order that we can effectively compare the results of the tests.

The first one is the step change in ±15% throughput changes. This can be done by changing the setpoint of the reactor feed flow (distillate flow of C2 or recycle flow) control loop and a fresh feed flow control loop. The second disturbance is fresh feed composition of heavy impurity changes. It is assumed that the fresh feed is not constant composition, but increase 0.02 composition of the heavy impurity by decrease 0.02 composition of reactant. The third disturbance is a fresh feed composition of reactant changes. It has increase and decrease 10% of reactant and product composition because the fresh feed is mixed butane. If the composition of reactant increases, composition of product in fresh feed will decrease. The last disturbance is catalyst activity decrease 10%. There are tests only in decreasing state because practically, catalyst will deactivate when the time passes. This test will provide us with a proper control structure that can control or solve this problem. Note that the plus disturbance response is represented by solid line and the minus disturbance response is represented by dot line. If there is an only disturbance, it will be represented by solid line. There are 2-3 data on the *y* axis, the first data is depicted in black, the second one is in dark gray and the third one is in light gray. Normally, the first data is on the primary axis and second (or the third) data is on the secondary axis.

The control structure efficiency is interested in two topics. First, how do these design variables and control loops affect the control of product quality ($i\mathrm{C}_4$ purity in the distillate of the DIB column)? Second, how do they affect the robustness of the control structures? All processes and control structures can be made to fail if they are upset with a large enough disturbances. It is desirable, though, for a process and control structure to handle relatively large disturbances. In order illustrate the dynamic of control structure designed using the new design procedure of Wongsri (2015).

5.3.1. Fresh feed flow rate or recycle flow rate disturbance

The control structures are tested for the responses of the fixed enter or reenter process follow step 3.1 with ±61 kmol/h (±15%) disturbances in the setpoint of the reactor feed flow rate controller for BC and CS1 and with +44.1 kmol/hr, -48.5 kmol/hr (เพื่อให้ recycle ±15% as same as BC and CS1) disturbances in the setpoint of the combined flow rate controller for CS2. CS1 control structure, there is changed from nominal value is 406.7 kmol/h to 467.7 kmol/h and 345.7 kmol/h as same as the recycle of BC. Other control structure, combine $n\mathrm{C}_4$ feed is changed from nominal value is 336.6 kmol/h to 380.7 kmol/h and 288.1 kmol/h. Table 5.25 shows the responses of the whole process for fixed flow disturbance. The step changes are made at time = 1 h and finished at time = 30 h the solid lines are responses of a 15% increase and the dashed lines are a 15% decrease.

In case increase fixed flow (IFF) disturbance in BC will result in decreasing C2 reflux drum level which will increase C2 reboiler heat input and push more heavy impurities to the top of the column. By doing so, C2 reflux drum level will be fulfilled resulting in the reduction of C2 reboiler level. Next, it increases C2 column feed (C1 bottom flow) to fulfill C2 reflux drum level and C2 reboiler level, but result in decreasing C1 reboiler level that increases fresh feed flow. Show that fresh feed flow will be increased, when recycle flow increases. From Table 5.25 (a), fresh feed flow increases in the beginning and reducing in the later cause C2 reboiler heat input is increased, result in heavy impurity will be boiled to the top of C2 and turn to process (reaction section), see Table 5.25 (g). Show that reactor inlet composition of reactant, conversion and product will be decreased, see Table 5.25 (d, k). The reactant remained and heavy impurity from C2 bottom are controlled to C1 bottom result in an increasing C1 reboiler level and signal to decrease fresh feed flow. When the process in to steady state, it shows fresh feed flow is less than the last steady state result in decreasing heavy impurity in process. That mean $n\mathrm{C}_4$ component in process is less than nominal value result in conversion does not change in nominal value.

For CS1, in case IFF has nearly the same change as that in BC But that has skipped step shown decreases C2 reflux drum level, increases C2 column feed, decreases C1 reboiler level and increase fresh feed flow, see Table 5.25 (a, e). Afterward, fresh feed flow dropped before then increase because beginning conversion will be decreased, resulting in $n\mathrm{C}_4$ component is returned to C1 and increases C1 reboiler level. But CS1 has a composition cascade loop to maintain reactor outlet $i\mathrm{C}_4$ composition result in adjusting fresh feed flow and product flow will be increased.

CS2 is different from BC and CS1 because of mode operation. CS2 is on-supply mode to maintain $n\mathrm{C}_4$ component flow by adjusting fresh feed flow. When the setpoint of fixed flow change, the combined feed flow control is signal to increase fresh feed flow (in case IFF) immediately, see Table 5.25 (a). Then increases C1 reboiler level, C1 bottom flow, C2 reflux drum level and recycle flow respectively and decreases conversion in the begin. After that CS2 had nearly the same as CS1.

Table 5.25 Dynamic test results for TPM flowrate changes.

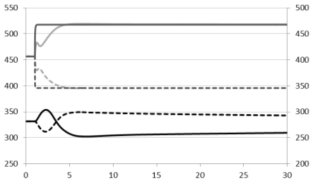
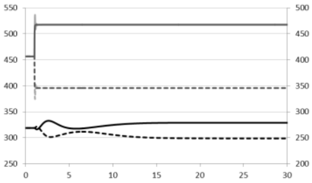
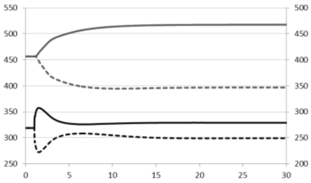
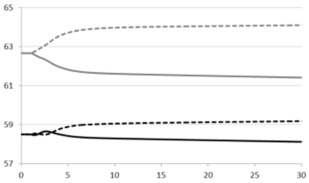
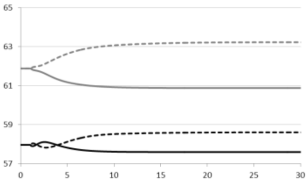
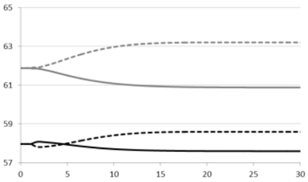
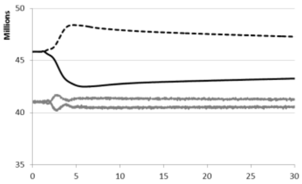
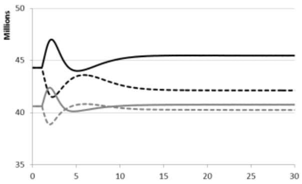
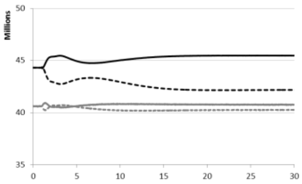
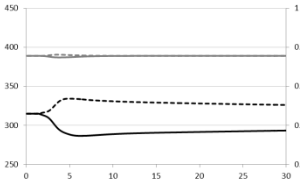
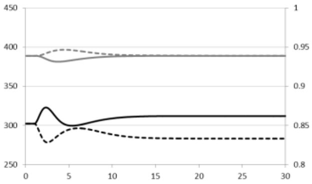
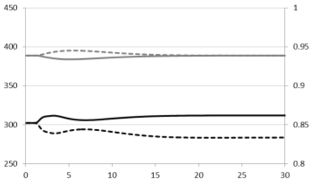
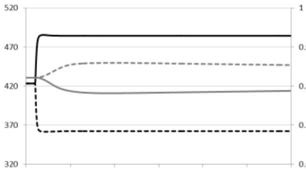
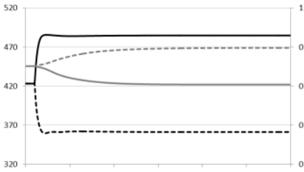
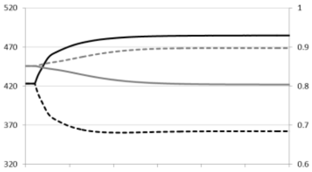
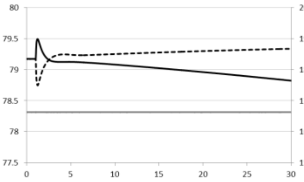
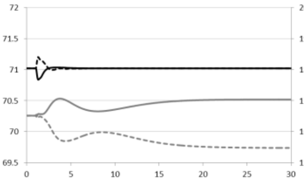
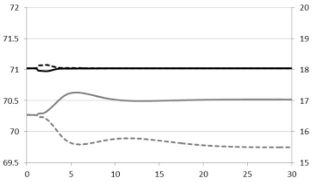
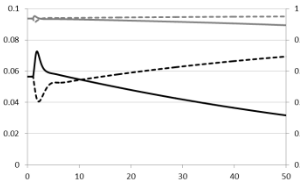
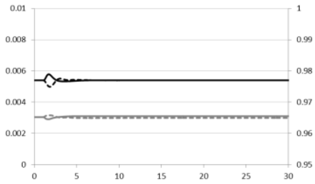
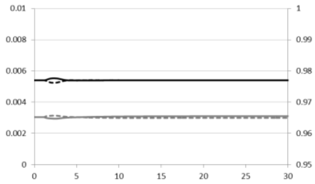
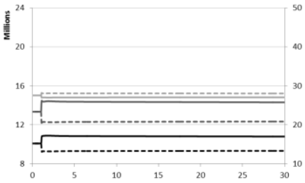
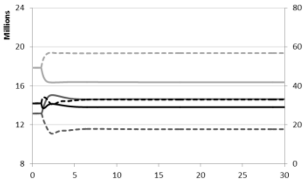
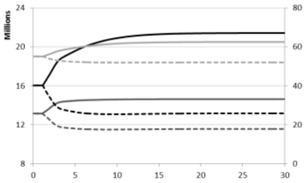
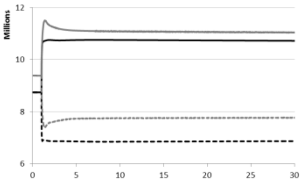
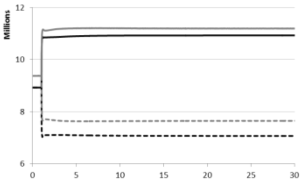
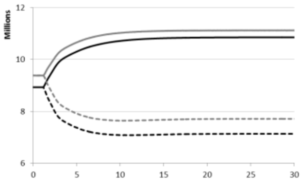
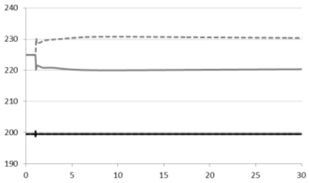
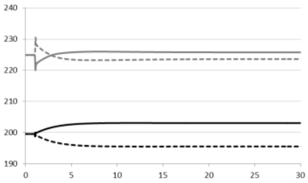
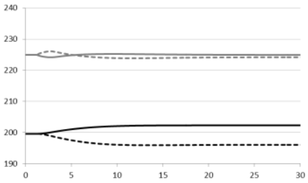
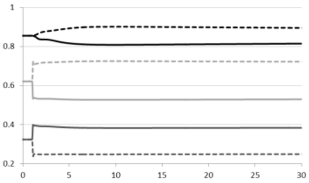
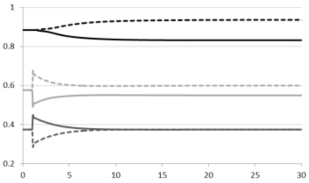
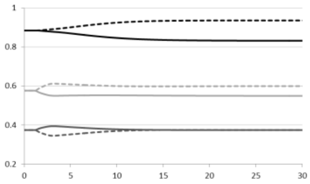
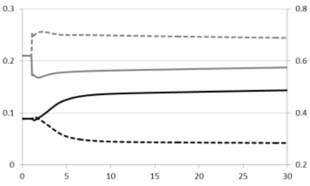
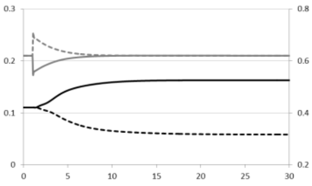
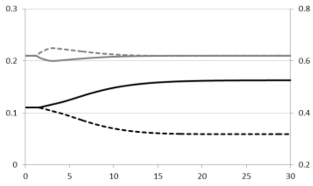
	BC	**CS1**	**CS2**
a - Fresh feed flow & Recycle flow & Flooded condenser outlet flow (kmol/h)			
b - C1 tray-28 & tray-44 temperature (C)			
c - C1 condenser heat removal & reboiler heat input (kJ/h)			
d - Product flow (kmol/h) & Product purity (iC_4)			
e - C1 bottom flow (kmol/h) & Bottom purity (nC_4)			
f - C2 tray-18 temperature (C) & Purge flow (kmol/h)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.25 (continued) Dynamic test results for TPM flowrate changes.

	BC	CS1	CS2
g – Reactor inlet composition (iC_5) & Purge purity			
h - C2 condenser heat removal & reboiler heat input (kJ/h) & reflux drum level (%)			
i - Furnace heat input & Flooded condenser heat removal (kJ/h)			
j - Reactor inlet & outlet temperature (C)			
k - Reactor inlet & outlet composition (nC_4) & conversion			
l- Reactor inlet & outlet composition (iC_4)			
	Time (hr)	Time (hr)	Time (hr)

5.3.2. Fresh feed purity disturbances

These disturbances are composition of fresh feed changes. The first one is increase composition of heavy impurity and decrease composition of reactant. The other one is composition of reactant changes, if composition of reactant increases will be decrease composition of product.

5.3.2.1 Heavy impurity feed composition disturbance

The control structures are tested for the responses of the iC_5 composition in fresh feed flow with 0.02 increase composition from 0.05 to 0.07. Table 5.26 shows the responses of the whole process for impurity feed composition disturbance. The step changes are made at time = 1 h and finished at time = 25 hr the solid lines are responses of 0.02 increase composition

For BC, when composition of heavy impurity in fresh feed increase that will not be response to C1 reboiler level because all of changes send to C1 bottom. In the first, C1 reboiler level is not change, therefore the fresh feed is not changing too. Disturbance will be flowing in C2. Heavy impurity that is increase will be increase C2 reboiler level result in decreasing C2 column feed flow. Because C2 bottom flow, purge flow, is fixed and C2 column feed flow is a little decrease, see Table 5.26 (e). It will be shown 3 effects are decreased C2 reboiler level, decrease C2 reflux drum level and increase C1 reboiler level. When C1 reboiler is increasing, fresh feed flow will be decreased. All levels of C2 are little decrease because column feed flow is decreased and outlet flows (distillate and bottom flow) are fixed. When C2 reboiler level decreases, column feed flow will be increased and change into nominal value. When C2 reflux drum level decrease, C2 reboiler heat input will be increased to a little more than nominal value, see Table 5.26 (h). It results in changing the heavy impurity to the top of the column to distillate flow (recycle flow) and go to the reactor, see Table 5.26 (g). That means decreases reaction (conversion) and reactor outlet composition of the product, see Table 5.26 (k, l). Finally, product flow will be decreased until process is not available, see Table 5.26 (d). BC has bad control loop is fixed flow loop at C2 bottom flow. It should change because heavy impurity has not the enough way to leave the plant. It will be accumulated in the process.

For CS1, it is as same as BC until C2. C2 reflux drum level will be decreased because reactant component in feed is decreases. The process will be increased C2 column feed for make up level and turn back to decrease C1 reboiler level result in increasing fresh feed flow for make up nC_4 component flow into the nominal value. Another action is increase component flow of heavy impurity (CFHI) but in this structure, CFHI will not accumulate because it has a control loop for impurity purge. It is level control loop to adjusting purge flow and maintain C2 reboiler level. For CS2, the purge has nearly the same as that in CS1. CS2 is on-supply mode, but CS1 is on-internal mode. Fresh feed flow will be changed immediately because it used combined flow control. From this disturbance test can be conclude that CS1 and CS2 are better than BC because they can purge the impurity.

Table 5.26 Dynamic test results for increase in iC_5 composition in fresh feed.

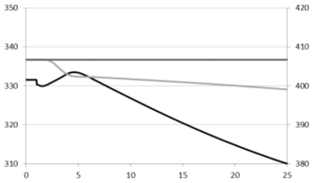
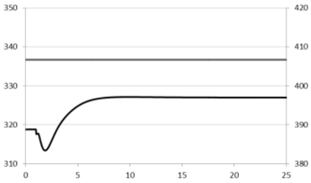
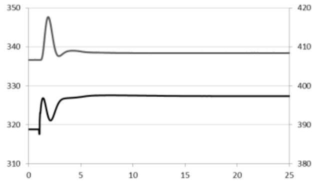
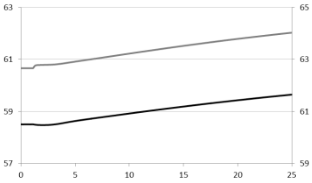
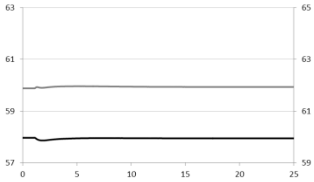
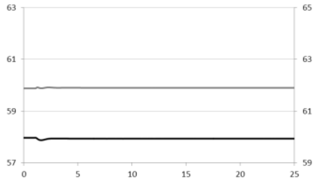
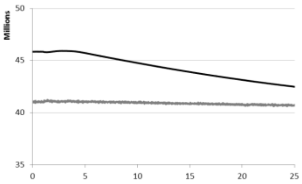
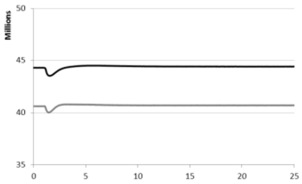
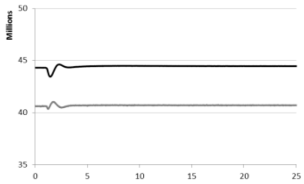
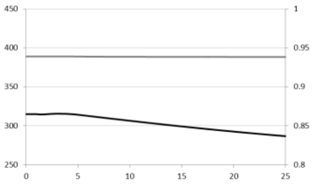
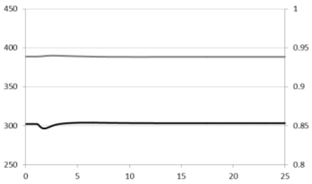
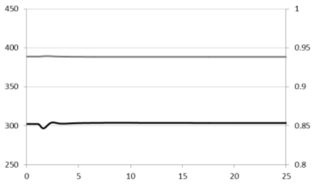
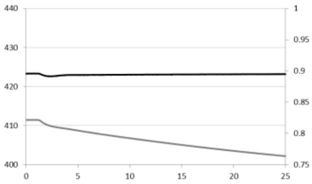
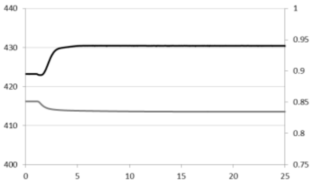
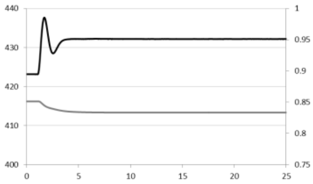
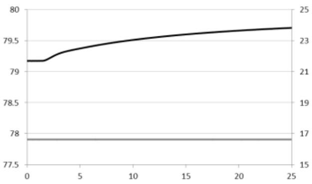
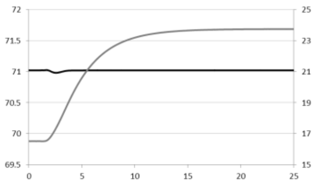
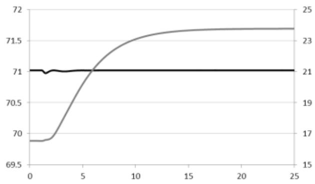
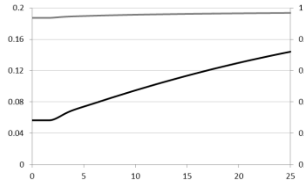
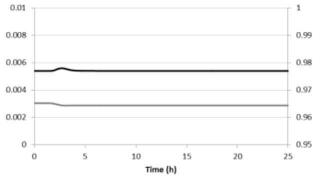
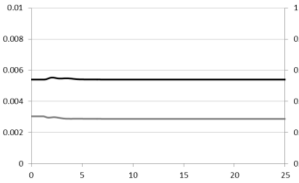
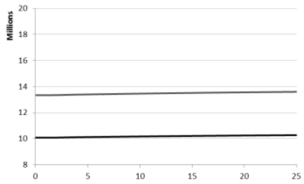
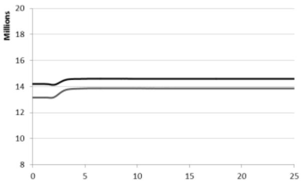
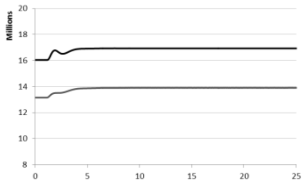
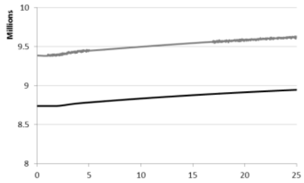
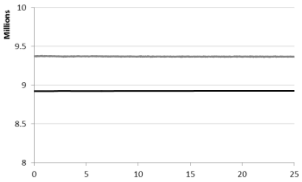
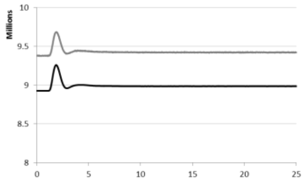
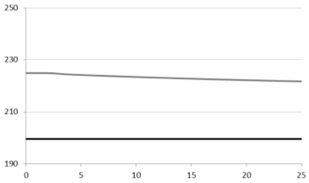
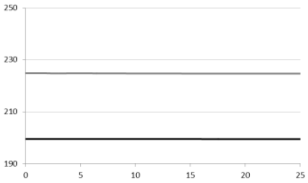
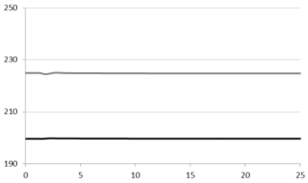
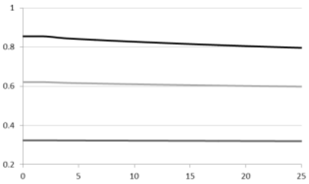
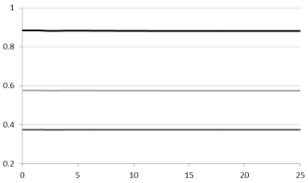
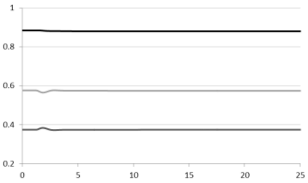
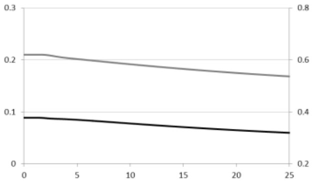
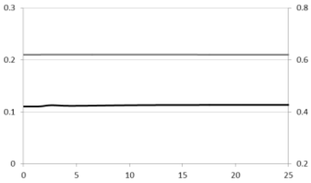
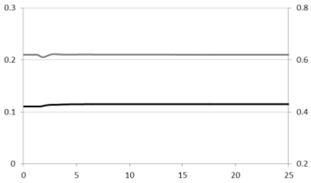
	BC	CS1	CS2
a - Fresh feed flow & Recycle flow & Flooded condenser outlet flow (kmol/h)			
b - C1 tray-28 & tray-44 temperature (C)			
c - C1 condenser heat removal & reboiler heat input (kJ/h)			
d - Product flow (kmol/h) & Product purity (iC_4)			
e - C1 bottom flow (kmol/h) & Bottom purity (nC_4)			
f - C2 tray-18 temperature (C) & Purge flow (kmol/h)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.26 (continued) Dynamic test results for increase in iC_5 composition in fresh feed.

	BC	CS1	CS2
g – Reactor inlet composition (iC_5) & Purge purity			
h - C2 condenser heat removal & reboiler heat input (kJ/h)			
i - Furnace heat input & Flooded condenser heat removal (kJ/h)			
j - Reactor inlet & outlet temperature (C)			
k - Reactor inlet & outlet composition (nC_4) & conversion			
l- Reactor inlet & outlet composition (iC_4)			
	Time (hr)	Time (hr)	Time (hr)

5.3.2.2 Reactant feed composition disturbance

The control structures are tested for the responses of the $n\text{C}_4$ composition in fresh feed flow with ±0.069 (±10%) Table 5.27 shows the responses of the whole process for reactant feed composition disturbance. The step changes are made at time = 1 h and finished at time = 30 hr the solid lines are responses of a 10% increase and the dashed lines are a 10% decrease.

When $n\text{C}_4$composition in BC change, it will be changed C1 bottom level and result in changing fresh feed flow. But fresh feed flow changed over all of the process will be in new steady state. With case increase $n\text{C}_4$ composition (INBC) will be increased $n\text{C}_4$ component in C1 reboiler, C1 bottom flow, C2 reflux drum, reactor inlet and increase conversion, see Table 5.27 (a, e, k). Reactor outlet composition of product is little decrease and then increase because reactor inlet composition of product decreases in fresh feed when $n\text{C}_4$ composition increased. Reactor outlet component of product is as same as before and of fresh feed is decrease will be decreased C1 reflux drum level cause product flow is decreased, see Table 5.27 (a, l). Because INBC will be decreased CFHI in fresh feed, there is no heavy impurity accumulation problem. But in case decrease $n\text{C}_4$ composition (DNBC) will be increased fresh feed flow and increase heavy impurity accumulation in the process, but less than heavy impurity composition disturbance test, see Table 5.27 (g). CS1 has nearly the same change as that in BC. There are differences in purge flow will be changed from C2 reboiler level. In case DNBC, there is no heavy impurity accumulation problem because purge flow will be increased for increase heavy impurity component. Reactor outlet composition of product is kept constant and result in changing into new steady state faster than BC. CS2 has nearly the same change as that in CS1. There are differences in fresh feed flow. They will be changed since disturbance is appearing by combined flow loop. They are more changes in the first time and run into steady state with offset more than CS1. In case INBC, the combined flow loop will be directed to decrease fresh feed flow cause to increase C1 reboiler level. Therefore C1 bottom flow, C2 reflux drum level and reactor inlet flow are decreased, see Table 5.27 (a, e). But reactor inlet composition of reactant will be increase result in increasing conversion. Therefore composition cascade loop will be signalled to decrease reactor inlet temperature and furnace heat input, see Table 5.27 (k, j, i)

Table 5.27 Dynamic test results for $n\mathrm{C}_4$ composition changes in fresh feed (±0.069).

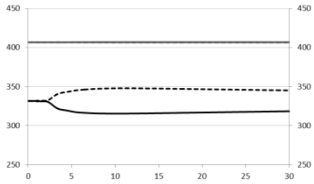
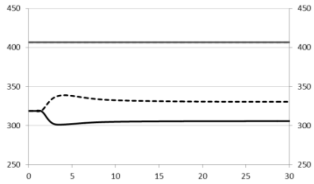
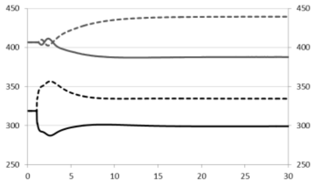
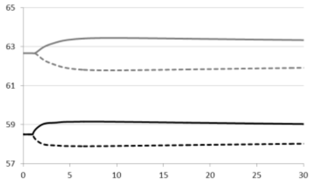
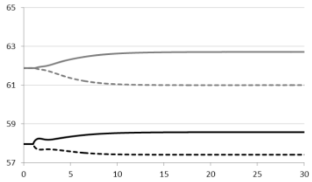
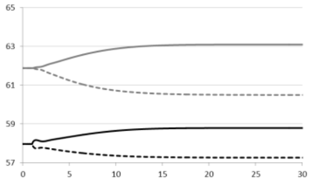
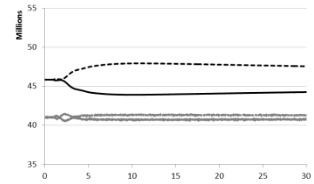
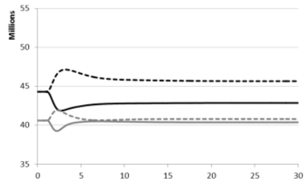
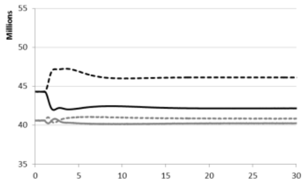
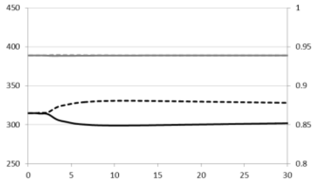
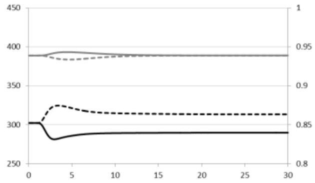
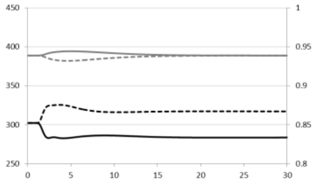
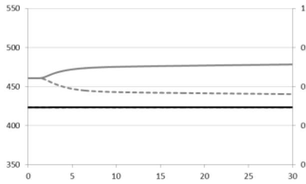
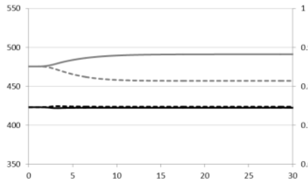
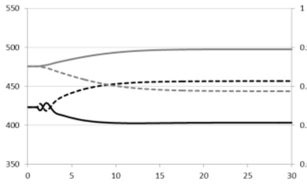
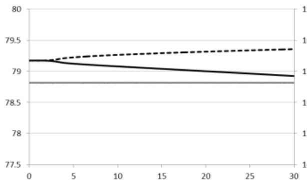
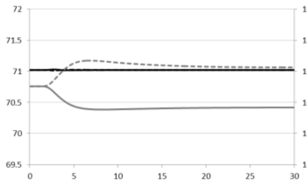
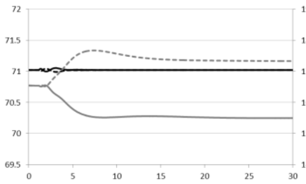
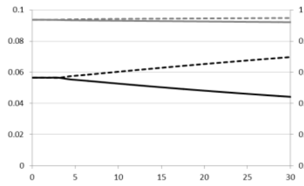
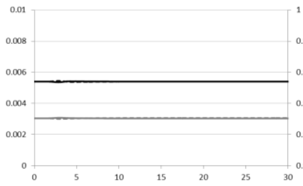
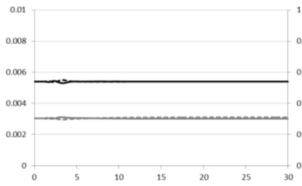
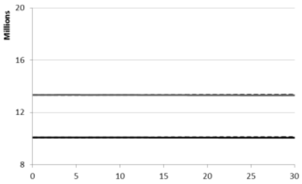
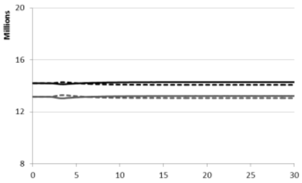
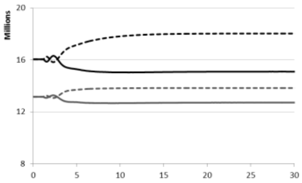
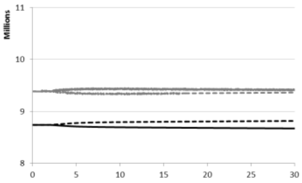
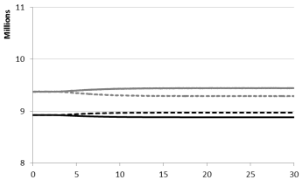
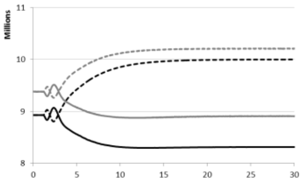
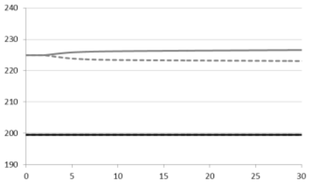
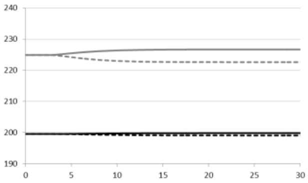
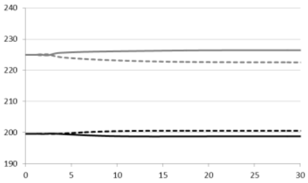
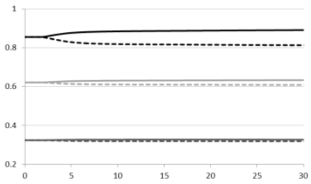
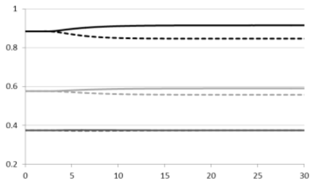
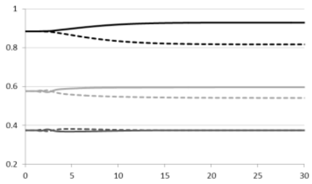
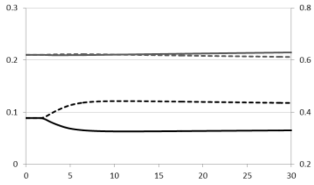
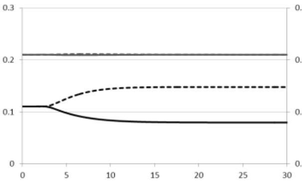
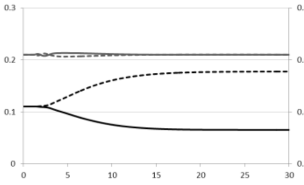
	BC	CS1	CS2
a - Fresh feed flow & Recycle flow & Flooded condenser outlet flow (kmol/h)			
b - C1 tray-28 & tray-44 temperature (C)			
c - C1 condenser heat removal & reboiler heat input (kJ/h)			
d - Product flow (kmol/h) & Product purity (iC_4)			
e - C1 bottom flow (kmol/h) & Bottom purity (nC_4)			
f - C2 tray-18 temperature (C) & Purge flow (kmol/h)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.27 (continued) Dynamic test results for nC_4 composition changes in fresh feed.

	BC	CS1	CS2
g – Reactor inlet composition (iC_5) & Purge purity			
h - C2 condenser heat removal & reboiler heat input (kJ/h)			
i - Furnace heat input & Flooded condenser heat removal (kJ/h)			
j - Reactor inlet & outlet temperature (C)			
k - Reactor inlet & outlet composition (nC_4) & conversion			
l- Reactor inlet & outlet composition (iC_4)			
	Time (hr)	Time (hr)	Time (hr)

5.3.3. Catalyst activity disturbance

The control structures are tested for the responses of deactivate of catalyst with decrease 10% kinetics from 4×10^8 to 3.6×10^8. Table 5.28 shows the responses of the whole process for deactivate of catalyst disturbance. The step changes are made at time = 1 h and finished at time = 25 hr the solid lines are responses of 10% decrease kinetic

When kinetic is decreased by 10%, it will be decreased conversion, see Table 5.28 (k). Reactor outlet composition of product will be decreased. Because product purity is controlled, product flow will be decreased, see Table 5.28 (d, l). Due to conversion is decreased, nC_4 component from the reactor will return to C1 and will be increasing C1 reboiler level. It will be decrease fresh feed flow, see Table 5.28 (a). When it decreases cause decrease heavy impurity to process, see Table 5.28 (g). That mean nC_4 composition and iC_4 composition will be increased slowly, see Table 5.28 (e, k, l). Therefore reactor outlet composition will be close to the setpoint in the future. CS1 and CS2 have nearly the same change as that in BC. Their reactor outlet composition of product can be increase into setpoint. Because they have composition cascade loop to control reactor outlet composition by adjusting setpoint of temperature loop to increase reactor inlet temperature and result in increasing conversion. Therefore, overall process will be changed into old steady state except reactor inlet temperature loop (reactor inlet temperature and furnace heat input).

Table 5.28 Dynamic test results for decrease in catalyst activity.

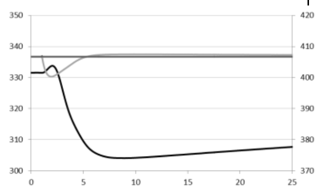
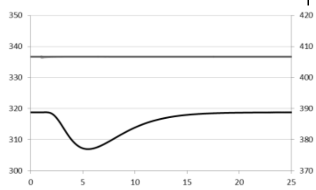
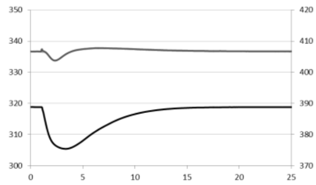
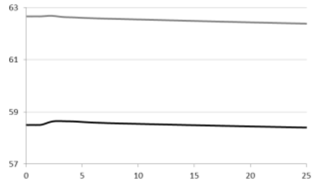
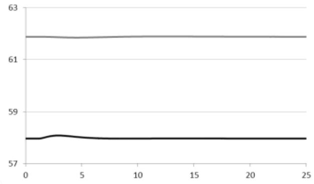
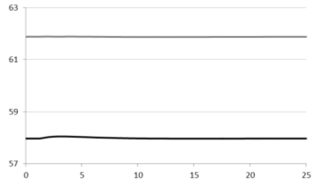
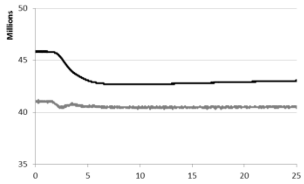
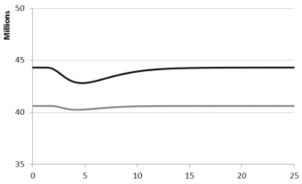
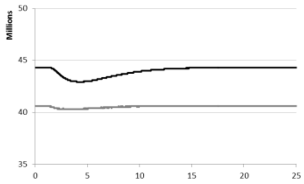
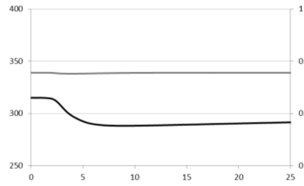
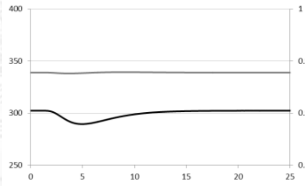
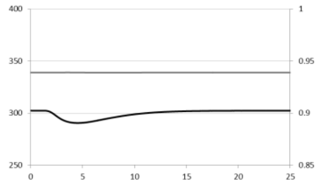
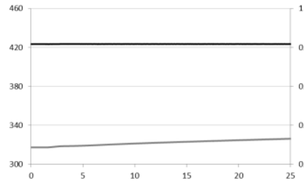
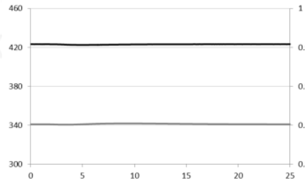
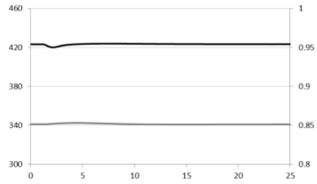
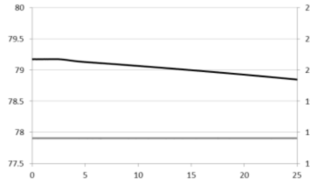
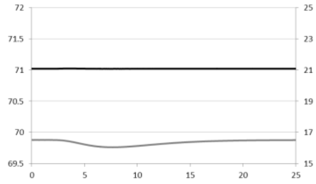
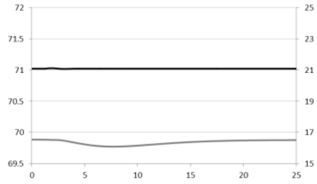
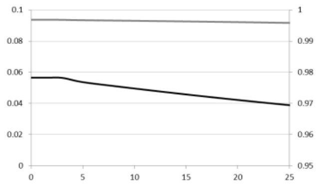
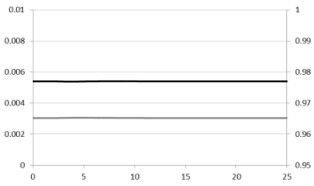
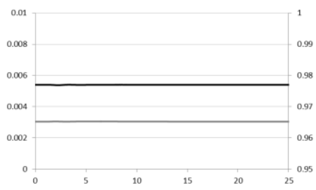
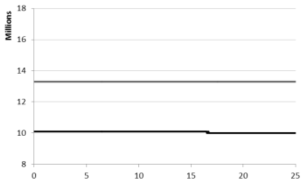
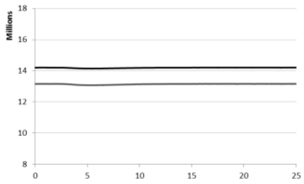
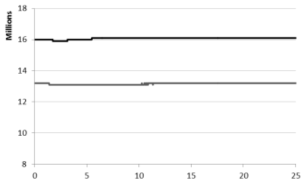
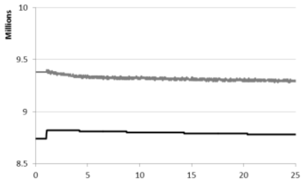
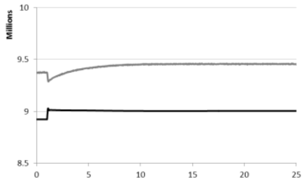
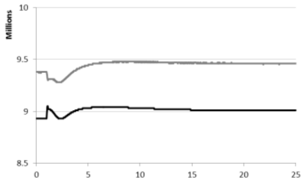
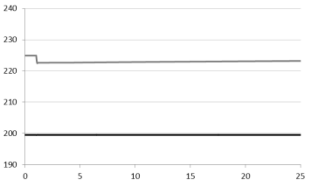
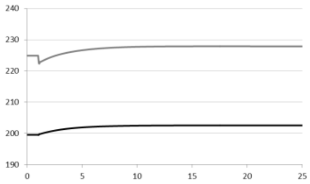
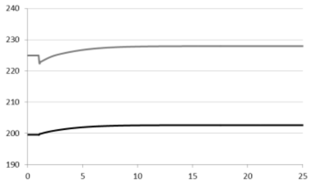
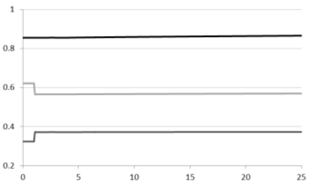
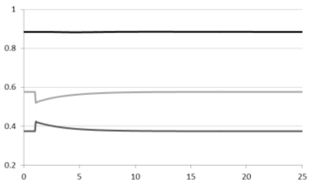
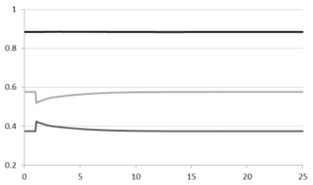
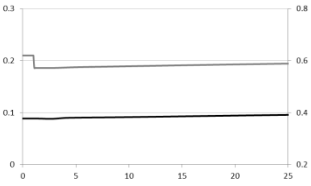
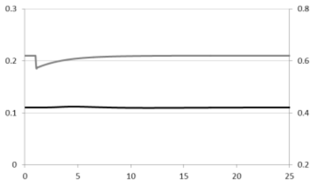
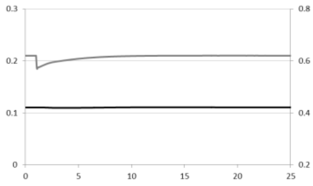
	BC	CS1	CS2
a - Fresh feed flow & Recycle flow & Flooded condenser outlet flow (kmol/h)			
b - C1 tray-28 & tray-44 temperature (C)			
c - C1 condenser heat removal & reboiler heat input (kJ/h)			
d - Product flow (kmol/h) & Product purity (iC_4)			
e - C1 bottom flow (kmol/h) & Bottom purity (nC_4)			
f - C2 tray-18 temperature (C) & Purge flow (kmol/h)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.28 (continued) Dynamic test results for decrease in catalyst activity.

	BC	CS1	CS2
g – Reactor inlet composition (iC_5) & Purge purity			
h - C2 condenser heat removal & reboiler heat input (kJ/h)			
i - Furnace heat input & Flooded condenser heat removal (kJ/h)			
j - Reactor inlet & outlet temperature (C)			
k - Reactor inlet & outlet composition (nC_4) & conversion			
l- Reactor inlet & outlet composition (iC_4)			
	Time (hr)	Time (hr)	Time (hr)

5.4 Control Structure Performance Evaluation

A common problem in control system design is establishing the appropriate value of controller gain. In general a low value of gain produces a slow system response, while high gain values can cause an excessive-oscillatory response to the possibility of instability. Somewhere between these extremes is a value of the gain that produces the best system response.

The tuning of different parameters in a controller is not exact, which is usually a compromise that is largely dependent on the process control. There is a method used to evaluate and compare the control structure, performance of different parameters. One method is using quantitative performance for dynamic process simulations. This type of performance assessment is usually used in an academic research and uses different criteria to minimize the value of error from the setpoint. One of the four basic criteria are in common use (IAE, ISE, ITAE and ITSE) is an integral absolute error (IAE). Use for systems that need to suppress all errors equally.

In this research, IAE method and IAE results are used to evaluate the dynamic performance of whole control structures. Variables in the process have different variable types, (temperature, pressure and composition) so to compare it, we must divide by span (the largest expected change in disturbance) of each variable. Control loops which are considered are safety loop.

Integral absolute error is well known and widely used. For the formulation of a dynamic performance as written below:

$$\text{IAE} = \int e\ |(t)|\ dt$$

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

For change in material disturbances consists of TPM flow rate, heavy impurity, reactant feed composition and catalyst activity. The IAE results for the handles each disturbance on 30 hr. and maintain product quality are shown in Table 5.30-5.33.

Table 5.29 IAE position for handle disturbance

Control loops	**IAE Position**
Temperature (a)	Reactor inlet temperature
Temperature (b)	Temperature of C1 sensitive tray Temperature of C2 sensitive tray
Pressure (a)	Condenser pressure of C1 column Condenser pressure of C2 column
Pressure (b)	Pressure of flooded condenser
Composition (a)	Product purity (nC_4 composition)
Composition (b)	Recycle purity (iC_5 composition) Reactor outlet composition (iC_4 composition)

Table 5.30 Summation of the IAE results for handle disturbance to the change in TPM flowrate.

Control Structures	**Control loops**					
	Temperature		**Pressure**		**Composition**	
	(a)	**(b)**	**(a)**	**(b)**	**(a)**	**(b)**
BC	0.0643	74.2015	4.7361	827.384	0.0024	4.1192
CS1	0.1061	0.7029	5.1547	1.6297	0.0890	0.3209
CS2	0.0794	0.8693	5.7645	1.5655	0.0882	0.2743

Table 5.31 Summation of the IAE results for handle disturbance to the change in feed heavy impurity.

Control Structures	**Control loops**					
	Temperature		**Pressure**		**Composition**	
	(a)	**(b)**	**(a)**	**(b)**	**(a)**	**(b)**
BC	0.0013	36.1258	2.2077	972.988	0.0010	3.1508
CS1	0.0006	0.0929	2.0107	0.0906	0.0040	0.0016
CS2	0.0214	0.1593	2.2628	0.3134	0.0011	0.0112

Table 5.32 Summation of the IAE results for handle disturbance to the change in reactant feed composition.

Control Structures	**Control loops**					
	Temperature		**Pressure**		**Composition**	
	(a)	**(b)**	**(a)**	**(b)**	**(a)**	**(b)**
BC	0.00005	49.7579	3.9885	181.505	0.0015	0.5657
CS1	0.0094	0.2408	4.2798	0.1948	0.0649	0.0309
CS2	0.0709	0.8548	4.9869	0.9393	0.0984	0.0925

Table 5.33 Summation of the IAE results for handle disturbance to the change in catalyst activity.

Control Structures	**Control loops**					
	Temperature		**Pressure**		**Composition**	
	(a)	**(b)**	**(a)**	**(b)**	**(a)**	**(b)**
BC	0.0027	10.4740	2.1827	136.888	0.0013	1.4115
CS1	0.0279	0.0328	1.9447	0.1508	0.0031	0.1279
CS2	0.0314	0.0707	1.9325	0.1958	0.0012	0.1278

Note that IAE of temperature (a), pressure (a) and composition (a) are control loops used for generating desired product quantity. However, the previous results showed us that those factors can control the product quality. From the table above, the IAE of both CS1 and CS2 are higher than that of BC. In the case of increasing heavy impurity, the IAE of CS1 is the best because CS1 and also CS2 are able to prevent the snowball effect, shown in Composition (b), but BC. IAE in the (b) column of the other variables is to show the variation of the important process variables, e.g. the tray temperatures (see Table 5.29).

In conclusion, arguably the best structure is CS1, since it has overall minimum IAE.

CHAPTER 6

Conclusion

The new plantwide control structure design procedure by Wongsri, based on heuristics, insight, and process control principles, is applied to design a control system for a butane isomerization plant. The procedure emphases on the use of quantifiers, which substantially indicate the amount of material in the system, and their handlers to design plant level inventory control. How to find the quantifiers is thoroughly explained and illustrated. Furthermore the procedure provides the disturbance management principles to design the quality loops. Luyben design is unable to prevent the snowball effect when the impurity iC_5 component in feed increased (Luyben, 1994).

The disturbance rejection loops are designed using the disturbance management to shift the thermal disturbance to the nearest exits and to shift the material disturbance to maintain the fixture plant status. Then the unit level loops are designed.

Two new designs using Wongsri's procedure are obtained. The major difference is on the location of TPM. For CS1, the TPM location is the recycle flow and for CS2, the TPM location is the combined feed to C1 column. It show that a proper TPM is the recycle flow rate since it provides better control performance i.e. slower responses and lager offsets.

Contrast to Luyben's conclusion (Luyben, 1999), recycle flowrate as a straightforward effect on feed and throughput on the condition that a component handler is chosen properly. The control performance of CS1 is better than that of Luyben's design on temperature control on both columns, heavy impurity control in reactor feed. Notably, increasing TPM increases the product rate in our design, while this decrease the product rate in Luyben's design. However Luyben's structure is able to keep the product purity better than our design. Our design the product purity is deviated about ±0.01 before it settled down.

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APPENDIX

APPENDIX A

PROCESS STREAMS AND EQUIPMENTS DATA

The two columns and a reactor must be sized for dynamic simulation and results are summarized in Table A.1

Table A.1 Equipment data of butane isomerization process

Section	Equipment	Specifications	
Separation	DIB column (C1)	Number of trays	50
		Feed to tray	20, 30
		Condenser Pressure (kPa)	689.5
		Reboiler Pressure (kPa)	723.9
		Diameter (m)	4.877
		Reflux drum volume (m^3)	113.3
		Reboiler volume (m^3)	96.28
		Reflux ratio	7.8
		Mole fraction of Distillate (iC_4)	0.935
		Mole fraction of Bottom (nC_4)	0.85
	Purge column (C2)	Number of trays	20
		Feed to tray	11
		Condenser Pressure (kPa)	448.2
		Reboiler Pressure (kPa)	457.1
		Diameter (m)	4.79
		Reflux drum volume (m^3)	22.65
		Reboiler volume (m^3)	20.95
		Reflux ratio	0.8
		Mole fraction of Distillate (nC_4)	0.885
		Mole fraction of Bottom (iC_5)	0.965
Reaction	Reactor	Diameter (m)	7.59
		Length (m)	2.134
		Volume (m^3)	27.135

Table A.2 Duty table of steady state operating for butane isomerization process

Variables	**R1**	**C1**		**C2**	
		Condenser	**Reboiler**	**Condenser**	**Reboiler**
Outlet Temperature (K)	498	322.8	338.2	317.9	347.8
Heat duty (MW)	0	-12.31	11.28	-16.04	13.16

Table A.3 Stream table for steady state operating of butane isomerization process

	1	**2**	**3**	**4**	**5**	**6**	**7**
Vapor Fraction	0.00	0.00	1.00	0.00	0.000	0.00	1.00
Temperature (°C)	32.21	63.44	50.52	49.79	49.79	65.20	45.42
Pressure (kPa)	769.9	840.70	689.48	689.48	689.48	729.40	448.16
Mole Flow (kmol/hr)	318.80	406.70	2247.3	302.28	1946.0	423.22	726.70
Mass Flow (kg/hr)	18664	23669	129957	17480	112533	24853	42294
Volume Flow (m^3/hr)	32.24	41.52	231.32	31.12	200.38	42.64	72.79
Mole Flow (kmol/hr)							
C_3	6.376	0	47.398	6.375	41.042	0	0
iC_4	76.51	252.15	2110	283.82	1827.1	44.83	80.10
nC_4	219.98	152.35	89.89	12.09	77.84	360.25	642.68
iC_5	15.94	2.196	0	0	0	18.14	3.924
Mole Fraction							
C_3	0.02	0.00	0.02	0.02	0.02	0.00	0
iC_4	0.24	0.62	0.94	0.94	0.94	0.106	0.1102
nC_4	0.69	0.375	0.04	0.04	0.04	0.851	0.8844
iC_5	0.05	0.005	0.00	0.00	0.00	0.043	0.0054

Table A.3 (continued) Stream table for steady state operating of butane isomerization process

	8	**9**	**10**	**11**	**12**	**13**	**14**
Vapor Fraction	0.00	0.00	0.00	0.00	1.00	1.00	1.00
Temperature (°C)	48.68	44.91	78.74	126.17	199.53	224.88	139.15
Pressure (kPa)	4884.8	448.16	457.22	3435.0	3229.1	3204.4	2970.0
Mole Flow (kmol/hr)	406.70	319.98	16.516	406.70	406.70	406.70	406.70
Mass Flow (kg/hr)	23670	18622	1183.6	23670	23670	23670	23670
Volume Flow (m^3/hr)	40.74	32.05	1.90	40.74	40.74	41.52	41.52
Mole Flow (kmol/hr)							
C_3	0	0	0	0	0	0	0
iC_4	44.83	35.27	0.0034	44.83	44.83	252.15	252.15
nC_4	359.68	282.98	0.5712	359.68	359.68	152.35	152.35
iC_5	2.196	1.728	15.942	2.196	2.196	2.196	2.196
Mole Fraction							
C_3	0	0	0	0	0	0	0
iC_4	0.1102	0.1102	0.0002	0.1102	0.1102	0.62	0.62
nC_4	0.8844	0.8844	0.0346	0.8844	0.8844	0.3746	0.3746
iC_5	0.0054	0.0054	0.9652	0.0054	0.0054	0.0054	0.0054

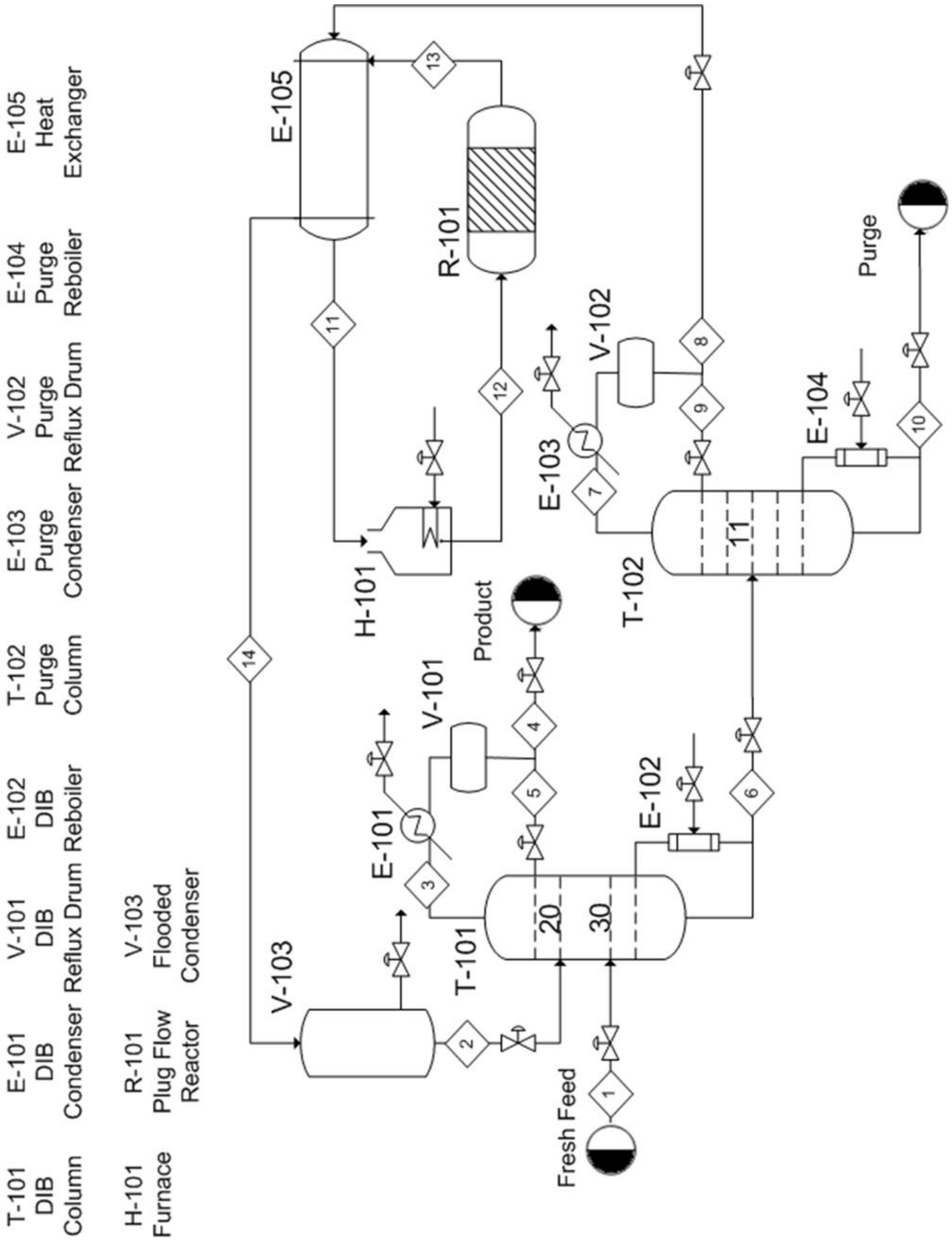

Figure A.1 Butane isomerization process flow diagram for steady state simulation

APPENDIX B

CONTROLLER TYPE AND TUNING PARAMETERS

B.1 Tuning Controller

Each process has its own dynamic characteristics that condition the tuning condition. If we do not have any preliminary tuning constant we have to find some start with. Each tuning method will end up with a different of tuning parameter. The first widely used technique for PID tuning was published by Ziegler-Nichols in 1942.

Flow controllers: The dynamic of flow measurement are fast processes. Therefore use PI with small gain and fast integral time.

Level controllers: These controllers are integrating processes, use P or PI controller.

Pressure controllers: These controllers are normally very fast loops that normally require PI, which high gain and fast integral time.

Temperature controllers: These controllers are normally very slow loops that normally require derivative time.

B.2 Tuning Parameter

Ȧström and Hägglund presented a relay feedback system to generate sustained oscillation as an alternative to the conventional continuous cycling technique for controller tuning in 1984. This relay feedback test was soon referred as auto-tune variation via the system as shown in Figure. B.1. Luyben popularizes the relay feedback method and calls this method “ATV” (Auto-Tune Variation). The auto-tune variation techniques are used to determine the ultimate period and the ultimate gain, which applied to Ziegler–Nichols (ZN) and Tyreus and Luyben (TL). Tuning values for proportional, integral and derivative controller parameters has been determined from the ultimate period and the ultimate gain.

Theoretical Analysis:

Read off the amplitude of PV signal wave “a” and OP signal wave “h”, it is used to calculate ultimate gain (Ku) from equation:

$$K_u = \frac{4h}{\pi a} \tag{B-1}$$

When Pu is period taken from limit cycle (ultimate period)

Ku is controller gain that produces the limit cycle (ultimate gain)

Kc is the controller path gain

Ti is the controller's integrator time constant

Td is the controller's derivative time constant

given two measured feedback loop parameters derived from measurements:

1. The period Pu of the oscillation frequency at the stability limit.
2. The gain margin Ku for loop stability

with the goal of achieving good regulation (disturbance rejection).

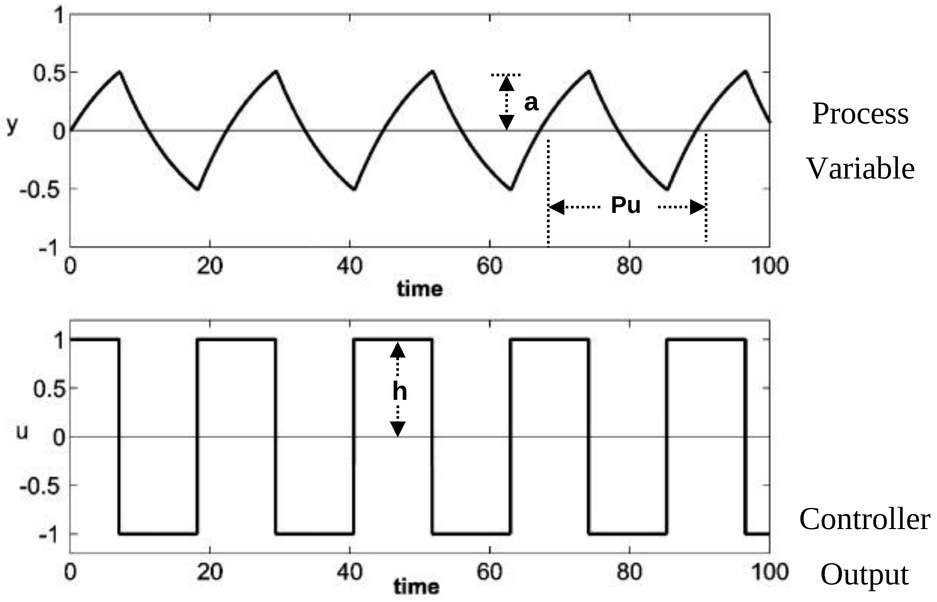

Figure B.1 Input and Output from Relay-Feedback Test.

B.2.1 Ziegler-Nichols Method

. Ziegler and Nichols published a classic paper in 1942, which introduced the continuous cycling method for controller tuning. They explained two empirical methods, which are used for obtaining controller parameters of P, PI, and PID controllers. These are the Ziegler-Nichols closed loop and open loop method. The closed-loop method is the most useful, and it is described below.

The Ziegler-Nichols rule is a heuristic tuning. The Ziegler-Nichols method is perhaps the first rigorous method to tune PID controllers. It is based on the following trial-and-error procedure. The ultimate gain (Ku) of a controller (which is the gain which causes sustained oscillations in the signals in the control system without the control signal reaching the maximum or minimum limits) must be found, and the ultimate (or critical) period (Pu) of the sustained oscillations is measured. Then, the controller is tuned by using Ku and Pu to calculate the controller gain, Kc. This method can be used in P-only, PI and PID controllers. The formulas for this method shown in Table B.1

For drawbacks of the Ziegler-Nichols method are small stability margin that the method may not be quick to use because it requires trial and error. The closed-loop behavior of this technique tends to be oscillatory and sensitive to uncertainty. The user has to make sure that the control signal does not reach its maximum and minimum values during the experiment.

Table B.1 Formulas for the controller parameters in the Ziegler-Nichols method

Controller	$\mathbf{K_C}$	$\boldsymbol{\tau_I}$	$\boldsymbol{\tau_D}$
P	K_U / 2.0	-	-
PI	K_U / 2.2	P_U / 1.2	-
PID	K_U / 1.7	P_U / 2.0	P_U / 8

B.2.2 Tyreus and Luyben Method

Tyreus and Luyben have suggested tuning parameter rules that result in less oscillatory responses, improves robustness and that are less sensitive to changes in the process condition, often called the TLC tuning rules. They formulas for the controller parameters are shown in Table B.2.

The Tyreus and Luyben tuning method is based on oscillations as in the Ziegler-Nichols method, but with modified formulas for the controller parameters to obtain better stability in the control loop compared with the Ziegler-Nichols method.

Table B.2 Formulas for the controller parameters in the Tyerus-Luyben method

Controller	$\mathbf{K_C}$	$\boldsymbol{\tau_I}$	$\boldsymbol{\tau_D}$
PI	K_U / 3.2	2.2 P_U	-
PID	K_U / 2.2	2.2 P_U	P_U / 6.3

Drawbacks of Tyreus-Luyben tuning method are sluggish disturbance compensation that the method may not be quick to use because it requires trial and error. The user has to make sure that the control signal does not reach its maximum and minimum values during the experiment.

Table B.3 Parameter tuning for Butane isomerization process of BC

Equipment	**Controller**	**Nominal value**	**PV Range**	**Tuning parameter**		
				K_c	**τ_I**	**τ_D**
Independent streams	FC01	406.7 kmol/hr	0-788 kmol/hr	0.5	0.3	-
Column C1	FC11	1946 kmol/hr	0-4000 kmol/hr	0.5	0.3	-
	CC11	0.04	0-0.1	0.15	72.9	16.2
	LC11	50 %	0-100 %	2	-	-
	LC12	47.89 %	0-100 %	2	-	-
	PC11	689.5 kPa	489.5-889.5 kPa	2	10	-
Column C2	FC21	320 kmol/hr	0-630.4 kmol/hr	0.5	0.3	-
	FC22	16.63 kmol/hr	0-27.2 kmol/hr	0.5	0.3	-
	LC21	27.81 %	0-100 %	2	-	-
	LC22	38.82 %	0-100 %	2	-	-
	PC21	448.2 kPa	248.2-648.2 kPa	2	10	-
Reactor	TC31	199.5 °C	148.9-248.9 °C	0.278	0.404	0.0897
Flooded condenser	LC41	50 %	0-100 %	2	-	-
	PC41	840.7 kPa	640.7-1040.7 kPa	2	10	-

Table B.4 Parameter tuning for Butane isomerization process of CS1

Equipment	**Controller**	**Nominal value**	**PV Range**	**Tuning parameter**		
				K_c	**τ_I**	**τ_D**
Independent streams	FC01	406.7 kmol/hr	0-788 kmol/hr	0.5	0.3	-
Column C1	FC11	1946 kmol/hr	0-4000 kmol/hr	0.5	0.3	-
	TC11	61.97 °C	10.19-110.19 °C	1.51	8.47	1.88
	LC11	50 %	0-100 %	2	-	-
	LC12	47.89 %	0-100 %	1	-	-
	PC11	689.5 kPa	489.5-889.5 kPa	2	10	-
	CC11	0.04	0-0.1	0.029	66.7	14.8
Column C2	FC21	320 kmol/hr	0-630.4 kmol/hr	0.5	0.3	-
	TC21	71.02 °C	21.02-121.02 °C	1.61	6.45	1.43
	LC21	50 %	0-100 %	1	-	-
	LC22	50 %	0-100 %	2	-	-
	PC21	448.2 kPa	248.2-648.2 kPa	2	10	-
	CC21	0.0054	0-0.05	3.46	18.7	4.15
Reactor	TC31	199.5 °C	148.9-248.9 °C	0.278	0.404	0.0897
	CC32	0.62	0-1	0.035	8.93	1.98
Flooded condenser	LC41	50 %	0-100 %	30	7	-
	PC41	840.7 kPa	640.7-1040.7 kPa	2	10	-

Table B.5 Parameter tuning for Butane isomerization process of CS2

Equipment	**Controller**	**Nominal value**	**PV Range**	**Tuning parameter**		
				K_c	**τ_I**	**τ_D**
Independent streams	FC01	372.3	0-673	0.5	3	-
Column C1	FC11	1946 kmol/hr	0-4000 kmol/hr	0.5	0.3	-
	TC11	61.88	10.19-110.19 °C	1.51	8.47	1.88
	LC11	50 %	0-100 %	2	-	-
	LC12	49.59 %	0-100 %	2	-	-
	PC11	689.5 kPa	489.5-889.5 kPa	2	10	-
	CC11	0.04	0-0.1	0.029	66.7	14.8
Column C2	FC21	320 kmol/hr	0-630.4 kmol/hr	0.5	0.3	-
	TC21	71.02 °C	21.02-121.02 °C	1.61	6.45	1.43
	LC21	49.28 %	0-100 %	2	-	-
	LC22	50 %	0-100 %	2	-	-
	PC21	448.2 kPa	248.2-648.2 kPa	2	10	-
	CC21	0.0054	0-0.05	3.46	18.7	4.15
Reactor	TC31	199.6 °C	148.9-248.9 °C	0.278	0.404	0.0897
	CC32	0.62	0-1	0.035	8.93	1.98
Flooded condenser	LC41	50 %	0-100 %	30	7	-
	PC41	840.7 kPa	640.7-1040.7 kPa	2	10	-

APPENDIX C.

IAE Results Data

Note that IAE of temperature (a) is reactor inlet temperature, (b) and (c) are temperatures at C1 and C2 sensitive tray, pressure (a) and (b) are pressure at the top of C1 and C2, (C) is pressure of flooded condenser, composition (a) is nC_4 composition in product flow, (b) is iC_5 composition in reactor inlet and (c) is iC_4 composition in reactor outlet.

Table C.1 The IAE of temperature Results for handle disturbance to the change in TPM flowrate

Control Structures	**Change**	**Control loops**			
		Temperature			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+15%	0.0375	29.8175	4.9450	74.2658
	-15%	0.0268	36.1569	3.2820	
CS1	+15%	0.0550	0.1282	0.1603	0.8091
	-15%	0.0511	0.2355	0.1789	
CS2	+15%	0.0385	0.2819	0.1012	0.9487
	-15%	0.0408	0.3695	0.1167	

Table C.2 The IAE of pressure Results for handle disturbance to the change in TPM flowrate

Control Structures	**Change**	**Control loops**			
		Pressure			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+15%	0.9384	1.4482	317.9489	832.1203
	-15%	0.8785	1.4710	509.4354	
CS1	+15%	1.0089	1.5193	0.8353	6.7845
	-15%	1.0991	1.5274	0.7944	
CS2	+15%	0.7476	1.9366	0.7640	7.3300
	-15%	0.9033	2.1769	0.8015	

Table C.3 The IAE of composition Results for handle disturbance to the change in TPM flowrate

Control Structures	**Change**	**Control loops**			
		Temperature			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+15%	0.0014	0.1991	1.6068	4.1215
	-15%	0.0010	0.1119	2.2014	
CS1	+15%	0.0379	0.0005	0.1501	0.4098
	-15%	0.0511	0.0005	0.1698	
CS2	+15%	0.0380	0.0002	0.1143	0.3625
	-15%	0.0503	0.0003	0.1595	

Table C.4 The IAE of temperature Results for handle disturbance to the change in feed heavy impurity.

Control Structures	**Change**	**Control loops**			
		Temperature			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+0.02	0.0013	24.8231	11.3026	36.1270
CS1	+0.02	0.0006	0.0379	0.0550	0.0935
CS2	+0.02	0.0214	0.1057	0.0535	0.1806

Table C.5 The IAE of pressure Results for handle disturbance to the change in feed heavy impurity.

Control Structures	**Change**	**Control loops**			
		Pressure			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+0.02	1.0586	1.1491	972.9885	975.1961
CS1	+0.02	0.7648	1.2459	0.0906	2.1014
CS2	+0.02	0.8322	1.4305	0.3134	2.5761

Table C.6 The IAE of composition Results for handle disturbance to the change in feed heavy impurity.

Control Structures	**Change**	**Control loops**			
		Composition			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+0.02	0.0010	1.6131	1.5377	3.1518
CS1	+0.02	0.0040	0.0003	0.0013	0.0055
CS2	+0.02	0.0011	0.0003	0.0109	0.0122

Table C.7 The IAE of temperature Results for handle disturbance to the change in reactant feed composition.

Control Structures	**Change**	**Control loops**			
		Temperature			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+10%	0.0000	20.0515	3.8651	49.7579
	-10%	0.0001	22.5437	3.2976	
CS1	+10%	0.0039	0.0878	0.0252	0.2502
	-10%	0.0055	0.1014	0.0265	
CS2	+10%	0.0315	0.3235	0.0655	0.9257
	-10%	0.0394	0.3947	0.0711	

Table C.8 The IAE of pressure Results for handle disturbance to the change in reactant feed composition.

Control Structures	**Change**	**Control loops**			
		Pressure			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+10%	0.8552	1.2284	93.3371	185.4935
	-10%	0.6920	1.2129	88.1680	
CS1	+10%	0.8480	1.1916	0.0947	4.4745
	-10%	0.9783	1.2618	0.1001	
CS2	+10%	0.9057	1.7301	0.4195	5.9262
	-10%	0.9848	1.3663	0.5198	

Table C.9 The IAE of composition Results for handle disturbance to the change in reactant feed composition.

Control Structures	**Change**	**Control loops**			
		Composition			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	+10%	0.0008	0.1795	0.1077	0.5673
	-10%	0.0008	0.1896	0.0889	
CS1	+10%	0.0317	0.0001	0.0118	0.0958
	-10%	0.0332	0.0001	0.0189	
CS2	+10%	0.0460	0.0002	0.0447	0.1910
	-10%	0.0524	0.0002	0.0475	

Table C.10 The IAE of temperature Results for handle disturbance to the change in catalyst activity.

Control Structures	**Change**	**Control loops**			
		Temperature			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	-10%	0.0027	5.0212	5.4528	10.4766
CS1	-10%	0.0279	0.0175	0.0153	0.0607
CS2	-10%	0.0314	0.0546	0.0161	0.1021

Table C.11 The IAE of pressure Results for handle disturbance to the change in catalyst activity.

Control Structures	**Change**	**Control loops**			
		Pressure			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	-10%	0.9463	1.2364	136.8886	139.0713
CS1	-10%	0.7243	1.2204	0.1508	2.0955
CS2	-10%	0.7242	1.2084	0.1958	2.1284

Table C.12 The IAE of composition Results for handle disturbance to the change in catalyst activity.

Control Structures	**Change**	**Control loops**			
		Composition			**Sum IAE**
		(a)	**(b)**	**(c)**	
BC	-10%	0.0013	0.3113	1.1002	1.4127
CS1	-10%	0.0031	0.0001	0.1278	0.1309
CS2	-10%	0.0012	0.0000	0.1278	0.1290

VITA

Mr. Kitt Kijnithikul was born in bongkok, Thailand on September 23, 1988. I graduated at Chulalongkorn University, Thailand and received the degree of Bachelor of Engineering in the field of Chemical Engineering in 2011. After that I entered the Graduate School of Chulalongkorn University to propose the Master degree of Engineering in Chemical Engineering and completed in 2015.