

Control of a Train of Distillation Columns for the Separation of Natural Gas Liquid

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ABSTRACT: The recovery of valuable heavier hydrocarbons from natural gas is accomplished in a series of five distillation columns. The first and second columns are cryogenic high-pressure columns in which the methane and ethane are sequentially taken overhead. The third column produces a propane distillate product. The fourth column removes isobutane and normal butane overhead for separation in the fifth column. The bottoms product from the fourth column is a mixture of hydrocarbons that are heavier than the C4 components.

This paper studies the dynamic control of this train of distillation columns. Effective control structures are developed and tested on all columns using conventional PI controllers. The product–quality loops in three of the columns use tray temperature control, but the first and last columns require composition controllers to effectively handle disturbances in feed composition. The difficult separation in the large, high reflux ratio de-isobutanizer is well-controlled by an unusual control structure that uses reboiler duty to control the reflux-drum level.

1. INTRODUCTION

Natural-gas processing is becoming more important in chemical engineering because of the increased supply and decreased cost of natural gas that have resulted from improved production methods. New chemical plants are being considered at locations near the Marcellus and Utica natural gas supply areas.

The composition of the natural gas varies from source to source but is mostly methane (85 mol % C1) with small amounts of ethane (C2), propane (C3), isobutane and normal butane (iC4 and nC4), and heavier hydrocarbons down to C7's. The C2 and heavier components are called natural gas liquid (NGL). They are more valuable than methane, so their recovery is important. It is achieved using a series of distillation columns.

The methane is separated out first in a high-pressure (25 atm) cryogenic distillation column using expansion to generate the low temperatures (180 K). A complex configuration of conventional and intermediate reboilers is used to provide vapor boilup and cool the feed. Two process–process heat exchangers, an expander, two flash drums, and two throttling valves are used to produce four different feed streams to the column. This column is called a de-methanizer. It is designed to achieve two objectives: (1) recover a specified desired percent of the ethane in the feed and (2) keep a low concentration of methane in its bottoms product.

It should be noted that the de-methanizer configuration used in this paper is a new design proposed by Nawaz et al.¹ that has economic advantages over some existing flowsheets (lower energy consumption).

The second column is called a de-ethanizer, and its function is to recover the ethane. It operates at 21 atm and 264 K, using refrigeration in the condenser. It is designed to achieve specified concentrations of propane impurity in the distillate and ethane impurity in the bottoms.

The third column is a de-propanizer whose function is to produce a propane distillate product. It operates at 17 atm with a reflux-drum temperature of 322 K so cooling water can be used in the condenser. It is designed to achieve specified concentrations

of isobutane impurity in the distillate and propane impurity in the bottoms.

The fourth column is a de-butanizer whose function is to take the isobutane and normal butane overhead for further separation in the downstream column. This iC4/nC4 separation is the most difficult (lowest relative volatility) of all the distillations, so it is economical to perform it after all other lighter and heavier components have been removed. The column operates at 7.1 atm with a reflux-drum temperature of 322 K, so cooling water can be used in the condenser. It is designed to achieve specified concentrations of isopentane impurity in the distillate and normal butane impurity in the bottoms.

The final column is a de-isobutanizer whose function is to produce an isobutane distillate product and a normal butane bottoms. The column operates at 6.6 atm with a reflux-drum temperature of 322 K so cooling water can be used in the condenser. It is designed to achieve specified concentrations of normal butane impurity in the distillate and isobutane impurity in the bottoms.

Since distillation is the premier separation method in the chemical and petroleum industries, the subject of distillation control has been extensively studied for many decades. Hundreds of papers and several books have explored many types of columns and many types of control structures. Most of these studies have looked at individual columns in isolation, but multicolumn systems have been examined in cases where recycle streams are present that connect the columns. Important examples are in azeotropic distillation systems (pressure-swing, extractive, and heterogeneous azeotropic units). Multiple columns in a plantwide environment have been explored in a variety of case

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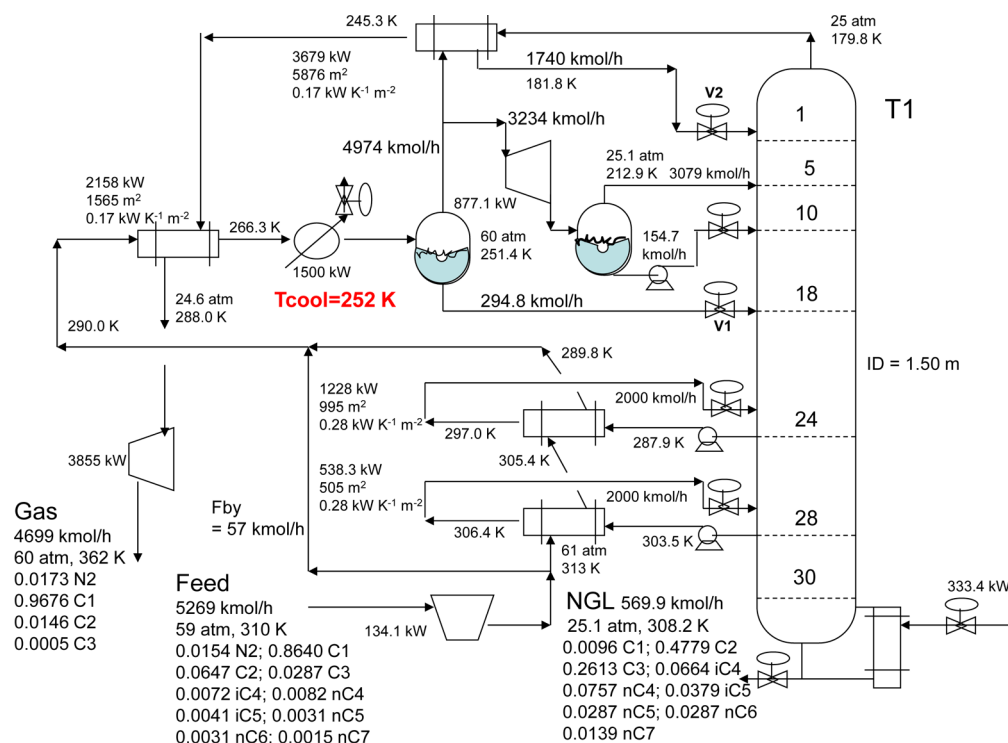


Figure 1. De-methanizer flowsheet.

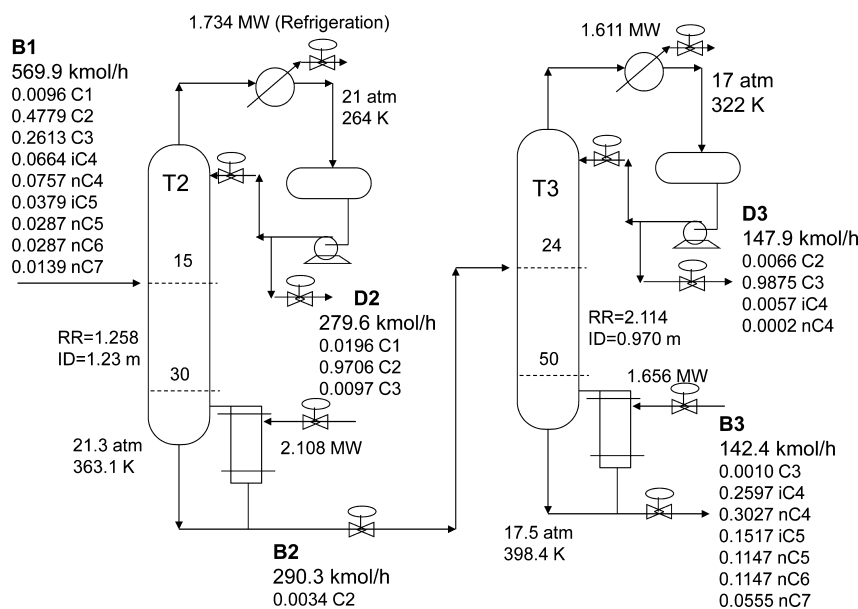


Figure 2. De-ethanizer and de-propanizer flowsheets.

studies of multiunit processes with reaction and separations sections.

Because of the importance and widespread application of the NGL separation system, we would expect the control of this system to be extensively studied. Many papers have explored the economic sequencing of the separation from the standpoint of steady-state design. However, the author has been unable to find any studies in the open literature that develop effective control structures for the train of distillation columns. That is the purpose of this paper. The only study (Zu et al.²) of this particular sequence of distillation columns found in the literature

deals with emergency flaring issues, not normal control structures.

Some dynamic control studies of cryogenic distillation systems have been reported. Hanakuma et al.³ reported the use of a self-tuning fuzzy control system to control the bottoms temperature by manipulated reboiler duty in a de-methanizer in an ethylene plant. Dynamic modeling of a somewhat similar cryogenic liquefied natural gas process was reported by Michelsen et al.⁴ The open-loop dynamic behavior was explored, but no closed-loop control studies were reported. Panahi and Skogestad⁵ discuss control of a natural gas to liquid plant. A control structure for a cryogenic ethylene splitter is described by Song et al.⁶ Many

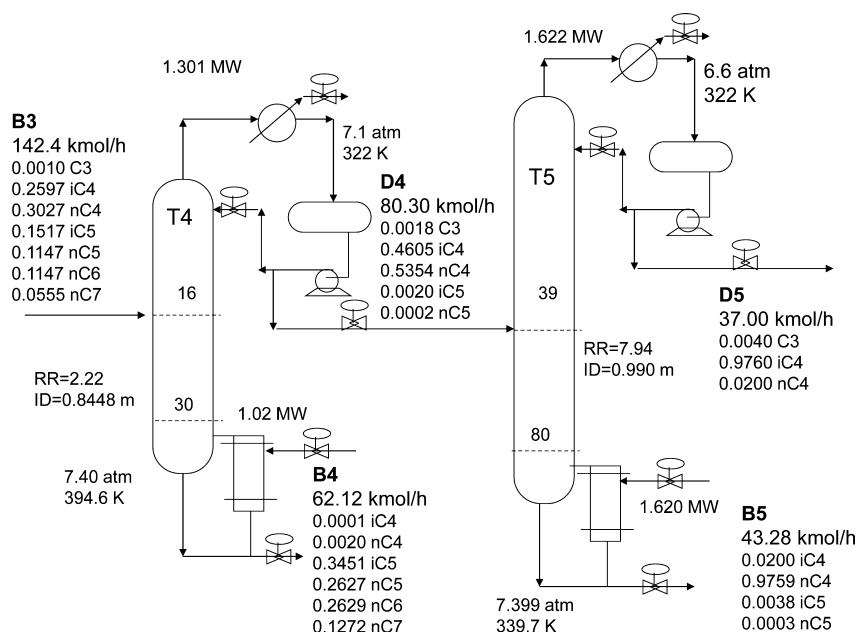


Figure 3. De-butanizer and de-isobutanizer flowsheets.

studies of the control of cryogenic heat-integrated liquid air columns have appeared.^{7,8}

Many papers dealing with individual de-ethanizers, de-propanizers, and de-isobutanizers have appeared.^{9–11} These separations are widely used in application examples for various techniques for control, identification, and optimization.

2. PROCESS STUDIED

The numerical example used in this paper uses the natural gas composition reported by Lee et al.¹³ and the de-methanizer configuration developed by Nawaz et al.¹ Figures 1–3 give the flowsheets of the five distillation columns in the system with their steady-state design conditions, energy duties, and equipment sizes. In the following sections, each of the columns is described in detail. Aspen simulations are used in this study with Peng–Robinson physical properties. Dynamic control studies uses the Gear numerical integration algorithm.

2.1. De-methanizer. The design of the de-methanizer is based on the following specifications and assumptions:

- (1) The recovery of C2 in the column bottoms NGL stream is set at 80%. This is achieved by varying the exit temperature from the externally cooled heat exchanger. Lower temperatures increase C2 recovery but increase refrigeration duty.
- (2) The composition of the bottoms stream is specified to produce a ratio of C1 to the sum of C1 and C2 of 0.02. Thus the bottoms composition on a C3-and-heavier-free basis is 2 mol % C1.
- (3) The column has 31 stages with four feed streams fed on different stages where stream and stage C1 compositions are similar.
- (4) The two side reboilers have liquid pump-around flow rates of 2000 kmol/h and are located at stages 24 and 28.
- (5) The pressure at the top of the column is 25 atm, and the stage pressure drop is 0.01 atm per tray.
- (6) Minimum approach temperature differentials at either the hot or the cold end of the heat exchangers are 2 K. Plate

heat exchangers are used because of the large area requirements.

- (7) The assumed overall heat-transfer coefficients for sizing gas–gas systems and gas–liquid systems are 0.17 and 0.28 kW K^{−1} m^{−2}, respectively.

The natural-gas feed flow rate is 5269 kmol/h with a composition of 86.4 mol % C1 and 47.79 mol % C2. There is a small amount of nitrogen (1.54 mol %). The remaining components run from C3 (2.87 mol %) down to a small amount of nC7. See Figure 1 for a detailed list of the feed composition. About 10% of the feed gas leaves the bottom of the de-methanizer as valuable NGL.

The supply pressure is 59 atm, and a small feed compressor boosts the pressure to 61 atm. The gas feed stream at 313 K is hotter than the temperatures in the stripping section of the column, so the use of intermediate (side) reboilers is attractive. Side reboilers have two advantages. They reduce energy consumption in the steam-heated reboiler at the base of the column, and they cool the feed so that less refrigeration is required.

Two side reboilers are used. The first takes a large pump-around liquid stream (2000 kmol/h) from stage 28 at 303.5 K and heats it to 306.4 K before returning it back to the column. The heat transferred in this heat exchanger is 538.3 kW, requiring 505 m² of area assuming an overall heat-transfer coefficient of 0.28 kW K^{−1} m^{−2} in this gas–liquid system and a minimum differential approach temperature of 2 K. The feed gas is cooled from 313 to 305.4 K and flows to the second side reboiler located higher in the column where the tray temperature is lower.

The second side reboiler takes a second large pump-around liquid stream (2000 kmol/h) from Stage 24 at 287.9 K and heats it to 297.0 K. The heat transferred in this heat exchanger is 1228 kW, requiring 995 m² of area. The feed gas is cooled from 305.4 to 289.8 K.

As shown in Figure 1, a small amount of the ambient-temperature feed gas (57 kmol/h) is bypassed around the two side reboilers. As discussed later in this paper, the side reboilers produce too much vapor boilup under some conditions (varying feed composition), and the duty of the steam-heated reboiler

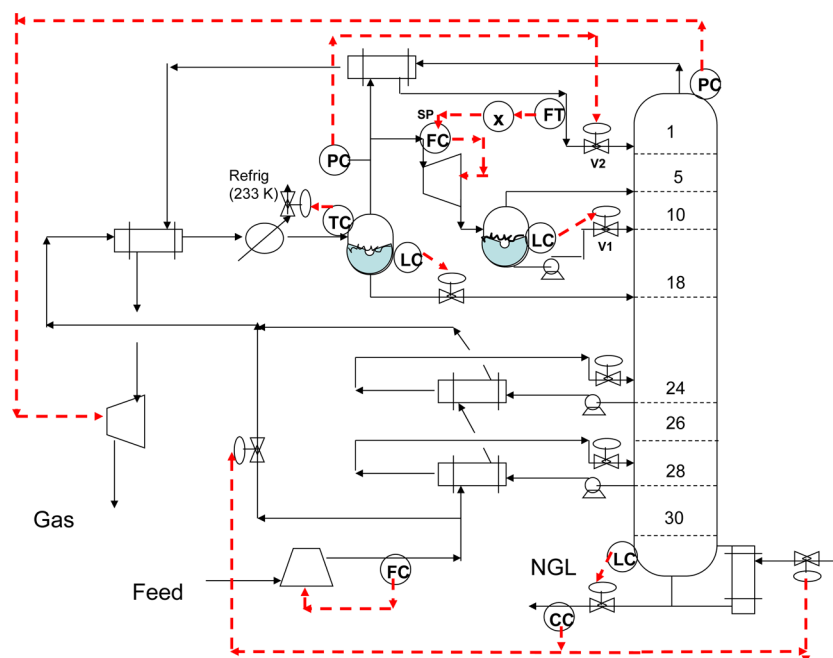


Figure 4. De-methanizer control structure.

would go to zero unless some bypassing is used. A control structure is developed that uses split-ranged valves to manipulate the two variables (reboiler steam and bypassing).

The mixed main and bypass streams flow to the next heat exchanger in which the cold methane product stream is used to further cool the feed. The cold stream to this feed pre-cooler is at 245.3 K with a flow rate of 4762 kmol/h. This is the gas product with a composition of 1.70 mol % nitrogen, 97.14 mol % C1, 1.11 mol % C2, and a trace of C3. This stream is heated to 288.0 K by the feed stream, which is cooled from 290.0 to 266.3 K. The heat-transfer rate is 2156 kW, requiring 1565 m² for an overall heat-transfer coefficient of 0.17 kW K⁻¹ m⁻² in this gas–gas system.

The feed then goes to a cooler that uses an external refrigerant to further cool the feed to 251.4 K, requiring 1500 kW of refrigeration duty. At 251.4 K and 60 atm, some liquid is formed. The feed stream goes to the first flash drum. The liquid phase at 294.8 kmol/h is flashed through valve V1 to near the operating pressure of the column (25 atm) and fed onto stage 18 of the 31-stage column. The temperature of the stream after the flash is 237.2 K with a vapor fraction of 0.2855.

The gas phase from the first flash drum is split into two streams. One is fed to an expander (3234 kmol/h) that produces 877.1 kW of power, which is used to help drive the main product compressor. The temperature of the expander exit is 212.9 K with a vapor fraction of 0.952. The stream leaving the expander goes to a second flash drum. The vapor from this drum (3079 kmol/h) is fed on stage 5. The liquid (154.7 kmol/h) is fed on stage 10.

The other portion of the gas stream from the first flash drum (1740 kmol/h) goes to a heat exchanger in which it is cooled and condensed at 181.8 K by the cold overhead vapor stream from the top of the column at 179.8 K. The heat exchanger transfers 3679 kW with an area of 5876 m² assuming an overall heat-transfer coefficient of 0.17 kW K⁻¹ m⁻² in this gas–gas system. The feed stream leaving this heat exchanger is all liquid. It flashes through valve V2 from 60 to 25 atm, which drops the temperature to 174.3 K and produces a vapor fraction of 0.1015.

At the base of the column a steam-heated reboiler has a heat duty of 333.4 kW. The base temperature is 308.2 K, so a low-

temperature heat source can be used. Notice that the heat added in the steam-heated reboiler is only 16% of the total heat added to the column in the three reboilers. The effect of this on dynamics and control is discussed in a later section of this paper. Column diameter is 1.50 m.

An important design optimization variable in the de-methanizer column is how the vapor stream leaving the first flash tank is split. Some is sent to the expander, and some is sent to the heat exchanger to be cooled by the overhead vapor. In the flowsheet given in Figure 1, the stream fed to the expander is 65% of the gas stream. The optimum split is determined by looking at the effects of the split on the recovered expander power, the reboiler duty, the refrigerated cooler duty, and the NGL production rate. The C2 recovery is kept constant by adjusting the refrigerated cooler temperature, and the composition of the NGL is kept constant at 2 mol % C1 on a C3-and-heavier-free basis. The minimum in the net energy (cooler + reboiler – expander) occurs at 65%, so this is selected as the operating point.

The de-methanizer is clearly the most complex of the columns. In the sections below, the designs of each of the four subsequent columns are discussed.

2.2. De-ethanizer. The column on the left in Figure 2 gives the flowsheet for the de-ethanizer column. It operates at 21 atm so as to be well under the critical pressure of ethane (48 atm) and avoid liquid/vapor hydraulic problems. At this pressure, the reflux-drum temperature is 264 K, so refrigeration is required (1.734 MW). The column base temperature is 363 K, so a low-temperature heat source is required (2.108 MW).

The C2/C3 separation is not difficult, and a column with 31 stages is used with feed introduced onto stage 15, the feed-tray location that minimizes the refrigerated condenser duty. The design specifications are 1 mol % C3 in the distillate and 0.34 mol % C2 in the bottoms. The required reflux ratio is 1.258. The purity of the ethane product D2 is 97.06 mol %. Column diameter is 1.23 m.

2.3. De-propanizer. The column on the right in Figure 2 gives the flowsheet for the de-propanizer column. It operates at

17 atm so that a reflux-drum temperature of 322 K is achieved, which permits the use of cooling water in the condenser. The C3/C4 separation is more difficult, so a column with 51 stages is used. The feed tray location that minimizes reboiler duty is stage 24.

The design specifications are 0.6 mol % C4 (sum of iC4 and nC4) in the distillate and 0.1 mol % C3 in the bottoms. The required reflux ratio is 2.114, and the reboiler duty is 1.656 MW of low-pressure steam. Column diameter is 0.97 m.

2.4. De-butanizer. The column on the left in Figure 3 gives the flowsheet for the de-butanizer column whose function is to take the iC4 and nC4 components overhead for subsequent separation in the downstream column. Because the iC4/nC4 separation is quite difficult (1.3 relative volatility at 322 K), conventional distillation wisdom recommends that this separation be performed when the mixture is essentially binary so as to minimize energy requirements.

The separation in the de-butanizer is between the nC4 and the iC5 and is fairly easy with a relative volatility of 2.2 at 322 K. A column with 31 stages is used, operating at 7.1 atm to give a reflux-drum temperature of 322 K. The feed tray location that minimizes reboiler duty is stage 16.

The design specifications are 0.2 mol % iC5 in the distillate and 0.2 mol % nC4 in the bottoms. The required reflux ratio is 2.20, and the reboiler duty is 1.02 MW of low-pressure steam. Column diameter is 0.845 m.

2.5. De-isobutanizer. The column on the right in Figure 3 gives the flowsheet for the de-isobutanizer column whose function is to make the difficult separation between iC4 and nC4. A column with 81 stages is used, operating at 6.6 atm to give a reflux-drum temperature of 322 K. The feed tray location that minimizes reboiler duty is stage 39.

The design specifications are 2 mol % nC4 in the distillate and 2 mol % iC4 in the bottoms. A high reflux ratio of 7.94 is required. Reboiler duty is 1.62 MW of low-pressure steam. Column diameter is 0.990 m.

3. CONTROL STRUCTURE DEVELOPMENT

The development of control structures for each of the five distillation columns is discussed in the following sections. All liquid holdups are sized to provide 5 min of holdup when half-full. All level controllers are proportional with $K_C = 2$ except in the de-isobutanizer, as discussed later.

3.1. De-methanizer Control Structure. A recent paper¹⁴ studied the de-methanizer control issues, and the proposed control scheme is shown in Figure 4. The specific loops are described below.

- (1) Feed flow rate is controlled by manipulating the power to the feed compressor.
- (2) Column pressure is controlled by manipulating power to the gas product compressor, which discharges into a 60 atm pipeline.
- (3) The temperature in the first flash drum is controlled by manipulating the refrigerant duty in the cooler.
- (4) The pressure in the first flash drum is controlled by valve V2 in the reflux line to the column.
- (5) The level in the first flash drum is controlled by manipulating the valve in the liquid line from the drum.
- (6) The flow rate through the expander is flow-controlled by manipulating speed and is ratioed to the reflux flow rate. The set point of the flow controller comes from a multiplier whose two inputs are the reflux flow rate and a

constant 1.858 (the steady-state design ratio to keep the vapor split at 65% through the expander).

- (7) The level in the second flash drum is controlled by manipulating the valve in the liquid line leaving the drum.
- (8) Liquid level in the bottom of the column is controlled by manipulating the bottoms NGL product.
- (9) The valves in the side-reboiler pump-around streams are held in fixed positions.
- (10) The methane impurity composition of the bottoms is controlled by manipulating the reboiler duty and the bypass flow rate, using a split-range valve configuration. When the duty of the steam-heated reboiler is reduced to a low level, the flow rate of the feed gas bypassed around the side reboilers is increased.

The first flash-drum temperature controller has 1-min deadtimes and is tuned by running relay-feedback tests and using Tyreus–Luyben settings. Its tuning constants are $K_C = 3.51$ and $\tau_I = 9.2$ min with a 100 K temperature transmitter span and a maximum heat removal of 715 500 cal/s. A 3-min deadtime is used in the composition controller (CC), which is similarly tuned giving $K_C = 0.785$ and $\tau_I = 41$ min with a 5 mol % C1 composition transmitter span.

The use of tray temperature controllers instead of a bottoms composition controller was explored in the previous paper but found to be ineffective for maintaining bottoms composition, particularly for disturbances in feed composition.

3.2. De-Ethanizer Control Structure. The column is separating C2 from C3. The possibility of using a simple single-end tray-temperature control structure is explored by first looking at the temperature and composition profiles. The C2 and C3 compositions are changing fairly rapidly around stage 23, where the temperature profile is also changing. So stage 23 is selected to test.

If a single-end control structure is to be effective, the remaining control degree of freedom must be fixed in an appropriate manner. The use of feed composition sensitivity analysis provides guidance in this determination. Using steady-state Aspen Plus simulation, the C2 and C3 compositions of the feed to the column are changed from their design values while maintaining the specified key-component impurity levels in the distillate and bottoms by using two Aspen *Design spec/vary* functions. The upper part of Table 1 gives the results of these calculations for the de-ethanizer column. The percent variation of the required reflux flow rate and reflux ratio over the range of feed compositions is shown to be 8.1% and 29.0%, respectively. This indicates that a single-end control structure with a fixed reflux-to-feed ratio should be better at handling feed composition changes than a fixed reflux ratio structure.

The control structure for the de-ethanizer is shown on the left side of in Figure 5 and has the following loops:

- (1) Feed comes in on level control from the upstream column.
- (2) Pressure is controlled by manipulating refrigeration duty in the condenser.
- (3) Reflux is ratioed to feed.
- (4) Reflux-drum level is controlled by manipulating the flow rate of the distillate product.
- (5) Base level is controlled by manipulating the flow rate of the bottoms.
- (6) Stage 23 temperature is controlled by reboiler duty. The controller has a 1-min deadtime and is tuned by a running relay-feedback test and using Tyreus–Luyben settings. Tuning constants are $K_C = 0.59$ and $\tau_I = 16$ min with a 100

Table 1. Feed Composition Sensitivity Analysis

De-ethanizer				
	feed ethane (mol fraction)	propane (mol fraction)	reflux (kmol/h)	reflux ratio
design	0.5278	0.2111	341.7	1.114
	0.4778	0.2611	357.7	1.258
	0.4278	0.3111	370.8	1.487
% change over range			8.12	29.0
De-propanizer				
	feed propane (mol fraction)	feed iC4 (mol fraction)	reflux (kmol/h)	reflux ratio
design	0.4934	0.1546	329.3	2.210
	0.5234	0.1246	315.6	2.114
	0.5534	0.0946	301.1	1.803
% change over range			8.90	20.4
De-butanizer				
	feed nC4 (mol fraction)	feed iC5 (mol fraction)	reflux (kmol/h)	reflux ratio
design	0.2526	0.2017	182.0	2.538
	0.3027	0.1517	173.6	2.220
	0.3526	0.1017	162.0	1.888
% change over range			11.5	29.5
De-isobutanizer				
	feed iC4 (mol fraction)	feed nC4 (mol fraction)	reflux (kmol/h)	reflux ratio
design	0.5005	0.4954	286.6	7.240
	0.4605	0.5354	288.0	7.940
	0.4205	0.5754	289.0	8.751
% change over range			0.80	19.0

K temperature transmitter span and a maximum heat removal of 1 010 000 cal/s.

A variety of disturbances are introduced into the process to test the effectiveness of this single-end tray-temperature control structure, and results are presented in a later section of this paper.

3.3. Depropanizer Control Structure. The key components in this column are C3 and iC4. The possibility of using a simple single-end tray-temperature control structure is explored by first looking at the temperature and composition profiles. The

C3 and iC4 compositions are changing fairly rapidly around stage 33, where the temperature profile is also changing. So stage 33 is selected to test.

Feed composition sensitivity analysis is used to see if a single-end control structure could be effective. Using steady-state Aspen Plus simulation, the C3 and iC4 compositions of the feed to the column are changed from their design values while maintaining the specified key-component impurity levels in the distillate and bottoms by using two Aspen *Design spec/vary* functions. The second part of Table 1 gives the results of these calculations for the de-propanizer column. The percent variation of the required reflux flow rate and reflux ratio over the range of feed compositions is shown to be 8.9% and 20.4%, respectively. This indicates that a single-end control structure with a fixed reflux-to-feed ratio should be better at handling feed composition changes than a fixed reflux ratio structure.

The control structure for the de-propanizer is shown on the right side of Figure 5 and has the following loops:

- (1) Feed comes in on level control from the upstream column.
- (2) Pressure is controlled by manipulating condenser duty. This is shown as manipulating cooling water, but other alternative setups are widely used such as flooded condensers, hot-vapor bypass, and vent-bleed systems.
- (3) Reflux is ratioed to feed.
- (4) Reflux-drum level is controlled by manipulating the flow rate of the distillate product.
- (5) Base level is controlled by manipulating the flow rate of the bottoms.
- (6) Stage 33 temperature is controlled by reboiler duty. The controller has a 1-min deadtime and is tuned by a running relay-feedback test and using Tyreus–Luyben settings. Tuning constants are $K_C = 0.89$ and $\tau_I = 13$ min with a 100 K temperature transmitter span and a maximum heat removal of 798 500 cal/s.

Evaluation of the effectiveness of this single-end tray-temperature control structure is presented in a later section of this paper.

3.4. De-butanizer Control Structure. The key components in this column are nC4 and iC5. The possibility of using a simple single-end tray-temperature control structure is explored by first looking at the temperature and composition profiles. The

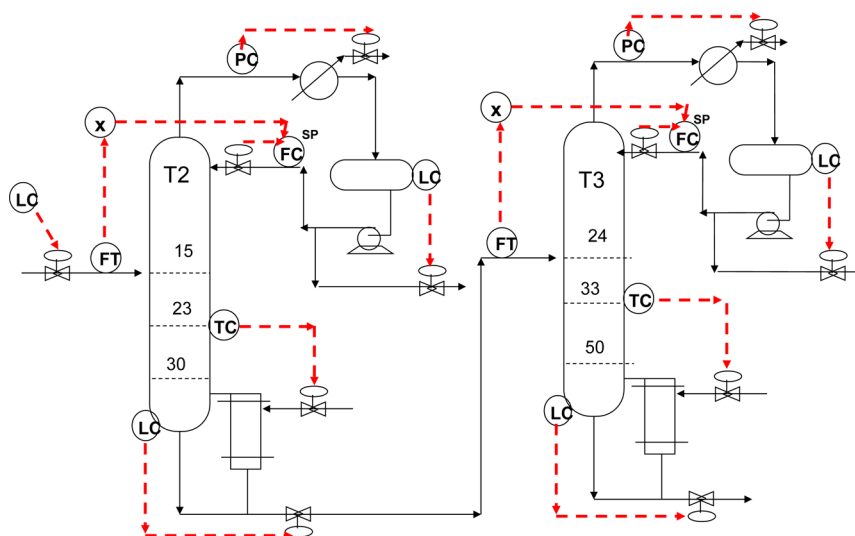


Figure 5. De-ethanizer and de-propanizer control structures.

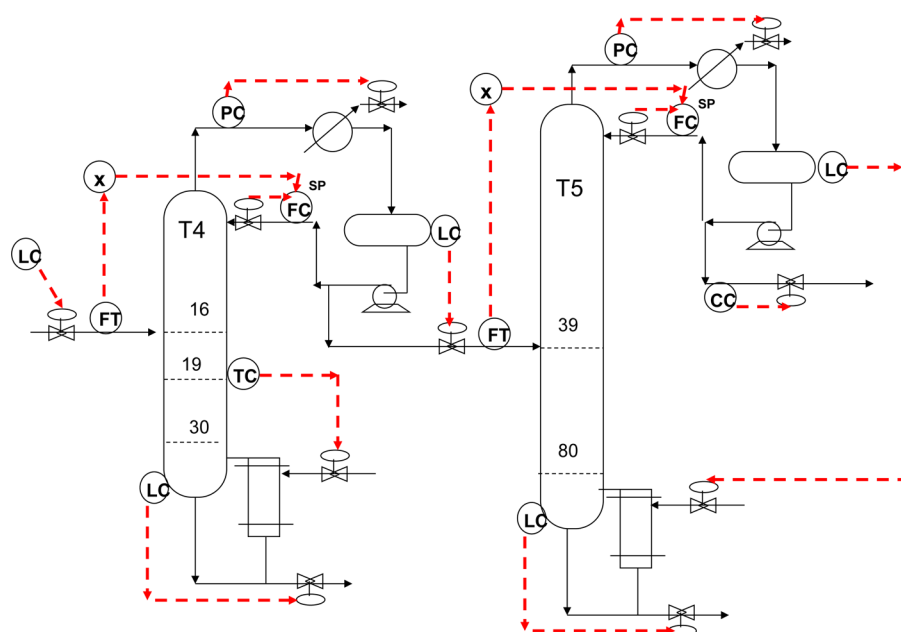


Figure 6. De-butanizer and de-isobutanizer control structures.

nC4 and iC5 compositions are changing fairly rapidly around stage 19, where the temperature profile is also changing. So stage 19 is selected to test.

Feed composition sensitivity analysis is used to see if a single-end control structure could be effective. Using steady-state Aspen Plus simulation, the nC4 and iC5 compositions of the feed to the column are changed from their design values while maintaining the specified key-component impurity levels in the distillate and bottoms by using two Aspen *Design spec/vary* functions. The third part of Table 1 gives the results of these calculations for the de-butanizer column. The percent variation of the required reflux flow rate and reflux ratio over the range of feed compositions is shown to be 11.5% and 29.5%, respectively. This indicates that a single-end control structure with a fixed reflux-to-feed ratio should be better at handling feed composition changes than a fixed reflux ratio structure.

The control structure for the de-butanizer is shown on the left side of Figure 6 and has the following loops:

- (1) Feed comes in on level control from the upstream column.
- (2) Pressure is controlled by manipulating condenser duty.
- (3) Reflux is ratioed to feed.
- (4) Reflux-drum level is controlled by manipulating the flow rate of the distillate product.
- (5) Base level is controlled by manipulating the flow rate of the bottoms.
- (6) Stage 19 temperature is controlled by reboiler duty. The controller has a 1-min deadtime and is tuned by running a relay-feedback test and using Tyreus–Luyben settings. Tuning constants are $K_C = 0.77$ and $\tau_I = 16$ min with a 100 K temperature transmitter span and a maximum heat removal of 488 000 cal/s.

Evaluation of the effectiveness of this single-end tray-temperature control structure is presented in a later section of this paper.

3.5. De-isobutanizer Control Structure. This column represents the most difficult separation and the most challenging control problem. The reflux ratio in this superfractionator is large, so conventional distillation control wisdom recommends

controlling the reflux-drum level using reflux. The temperature profile is flat because the changes in compositions from tray to tray are very small. So the use of temperatures to infer compositions is not going to be effective.

The key components in this column are iC4 and nC4. Feed composition sensitivity analysis is used to see if a single-end composition control structure could be effective or if a dual-composition control scheme is required. Using steady-state Aspen Plus simulation, the iC4 and nC4 compositions of the feed to the column are changed from their design values while maintaining the specified key-component impurity levels in the distillate and bottoms by using two Aspen *Design spec/vary* functions. The bottom part of Table 1 gives the results of these calculations for the de-isobutanizer column. The percent variation of the required reflux flow rate and reflux ratio over the range of feed compositions is shown to be 0.8% and 19%, respectively. This indicates that a single-end composition control structure with a fixed reflux-to-feed ratio should be significantly better at handling feed composition changes than a fixed reflux ratio structure.

It is important to note that there are two competing rules from distillation control wisdom. The first suggests that the reflux-drum level should be controlled by reflux because of the very high reflux ratio. The second suggests that the reflux should be ratioed to the feed. It appears that we cannot follow both of these rules. However, there is a control structure that permits both.

The control structure for the de-butanizer is shown on the right side of Figure 6 and has the following loops:

- (1) Feed comes in on level control from the reflux drum of the upstream column.
- (2) Pressure is controlled by manipulating condenser duty.
- (3) Reflux is ratioed to feed.
- (4) Reflux-drum level is controlled by manipulating reboiler duty using a proportional controller with a gain of 10.
- (5) Base level is controlled by manipulating the flow rate of the bottoms.
- (6) Distillate nC4 impurity is controlled by manipulating the control valve in the distillate line. The controller has a 3-min deadtime and is tuned by running a relay-feedback test

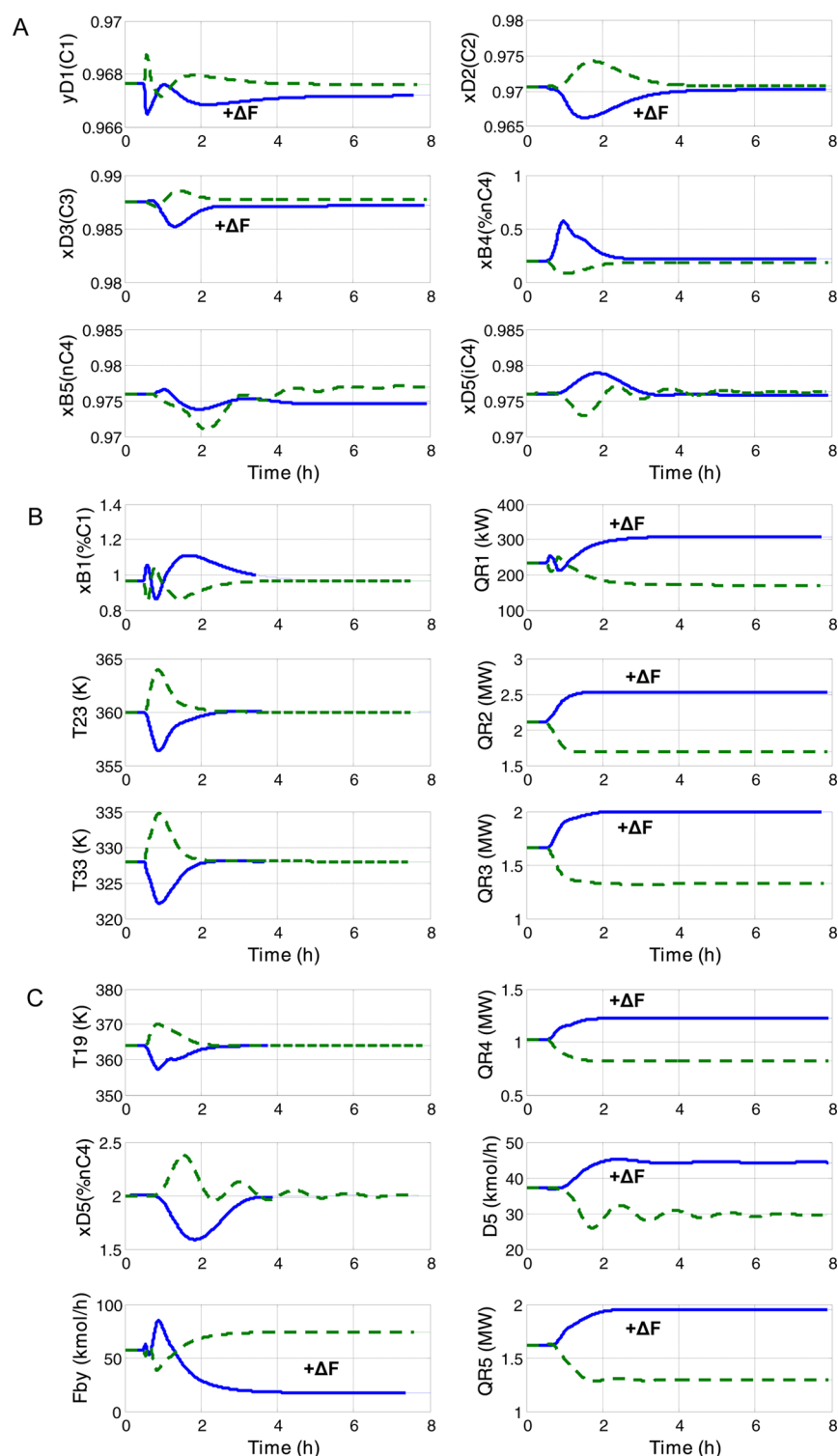


Figure 7. Feed flow rate disturbances: (A) Product compositions; (B) de-methanizer (DM), de-ethanizer (DE), and de-propanizer (DP) loop variables; (C) de-butanizer (DB) and de-isobutanizer (DIB) loop variables.

and using Tyreus–Luyben settings. Tuning constants are $K_C = 1.6$ and $\tau_I = 50$ min with a 5 mol % composition transmitter span.

This control structure is unusual because of the use of reboiler heat input to control the reflux-drum level. Since the reflux ratio is large, the reflux-drum level would normally be controlled by manipulating the reflux flow rate. However, the reflux-to-feed control structure is indicated by the feed composition sensitivity

analysis to require very small changes in this ratio, so it is logical to fix this ratio. However, some other effective variable must be selected to control the liquid level in the reflux drum. Using distillate would be ineffective because its flow rate is much smaller than the reflux. Using reboiler duty works well for this job because the changes in vapor flow rates in the column are quite rapid.

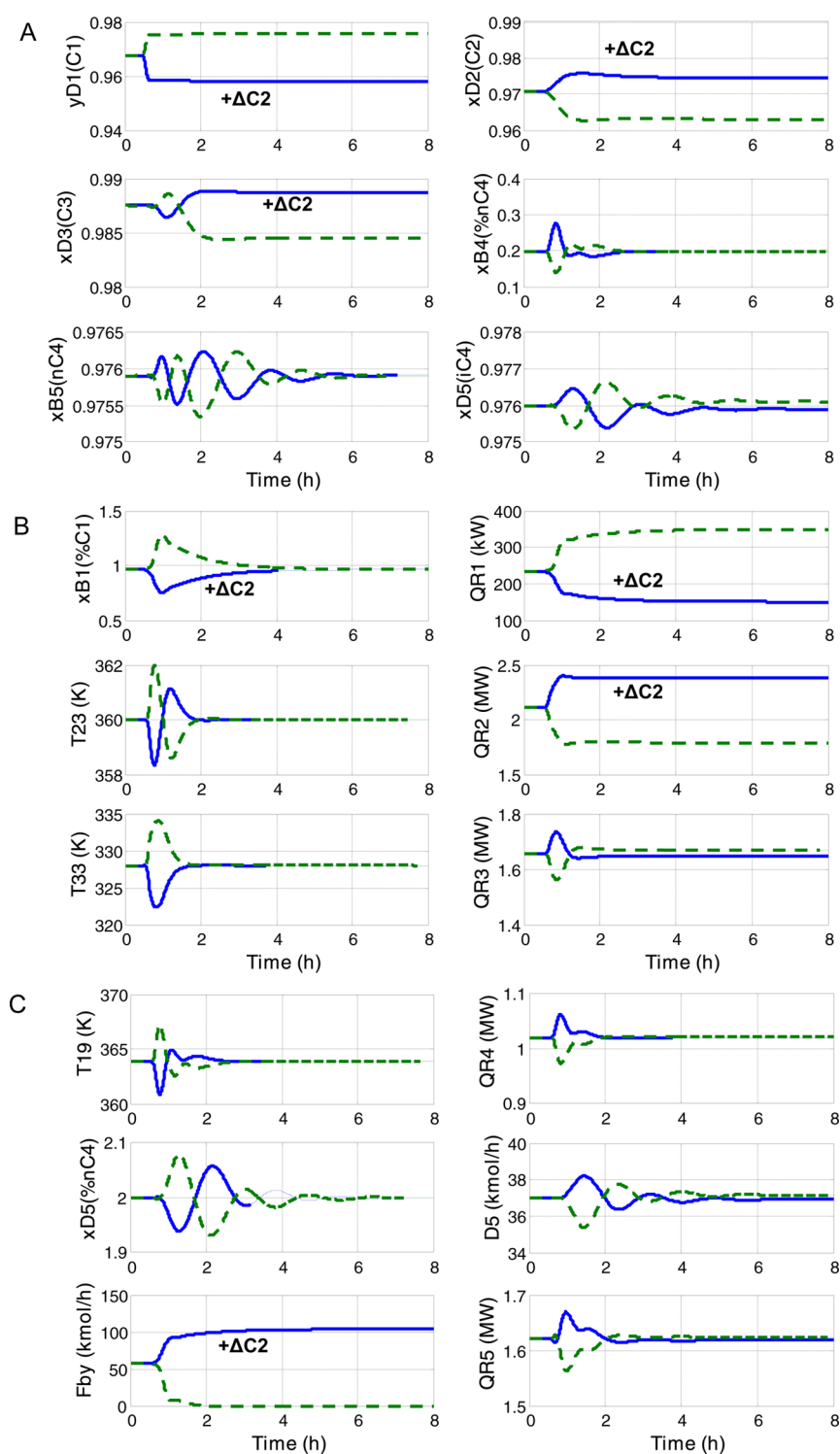


Figure 8. Feed composition C2 disturbances: (A) Product compositions; (B) DM, DE, and DP loop variables; (C) DB and DIB loop variables.

The distillate composition is controlled by the flow rate of distillate. A gain of 10 is used in the reflux-drum level controller to give faster response of vapor boilup to level changes since the changes in distillate flow rate need to be quickly detected by the level controller, which then changes the vapor rate up the column and affects distillate composition. Notice that the pressure controller manipulating condenser duty is “nested” inside the composition control loop. The pressure controller must be on automatic for the composition controller to work.

Evaluation of the effectiveness of this single-end composition control structure is presented in the next section of this paper.

4. DYNAMIC PERFORMANCE

The effectiveness of the control structures for the five columns is tested by making large changes in throughput and in feed composition..

4.1. Feed Flow Rate Disturbances. Figure 7 gives dynamic responses for 20% disturbances in the set point of the feed gas

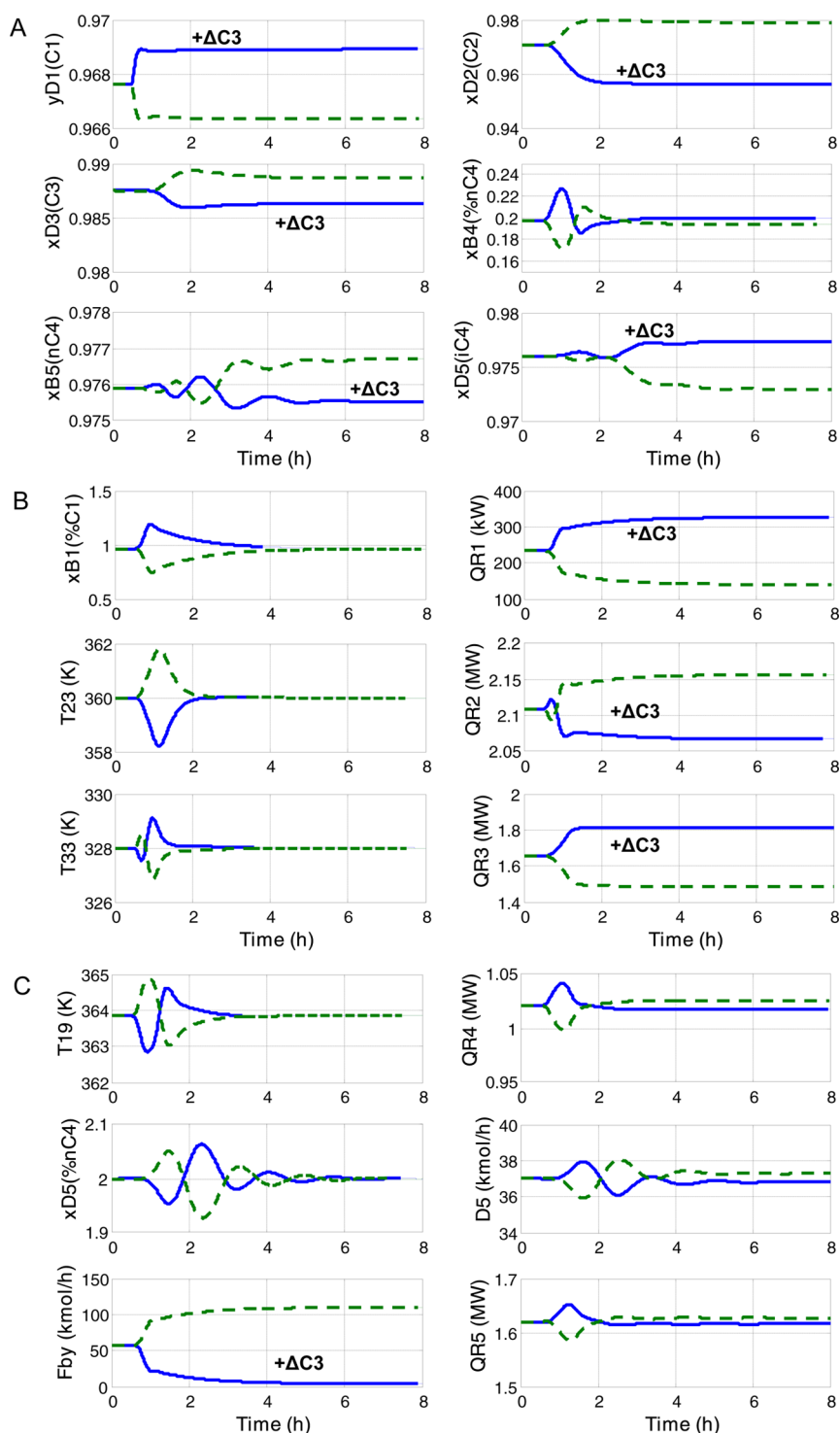


Figure 9. Feed composition C3 disturbances: (A) Product compositions; (B) DM, DE, and DP loop variables; (C) DB and DIB loop variables..

flow controller at time equal to 0.5 h. The solid lines are for 20% increases, and the dashed lines are for 20% decreases. Figure 7A shows how all the product compositions leaving the process respond to these large disturbances. The upper left graph shows how the C1 composition $y_{D1(C1)}$ of the product gas stream going overhead in the de-methanizer responds. This composition is not controlled, but it changes very little from the normal 96.76 mol % C1.

The upper right graph shows that the composition of the ethane product $x_{D2(C2)}$ also stays very close to the normal 97.06

mol % C2. Remember that this composition is not directly controlled, but stage 23 temperature is controlled and a reflux-to-feed ratio is maintained. This structure works well for throughput disturbances.

The middle left graph in Figure 7A shows that the composition of the propane product $x_{D3(C3)}$ is also held quite close to the design value of 98.75 mol % C3 when using a stage-33 temperature controller and a reflux-to-feed ratio control structure.

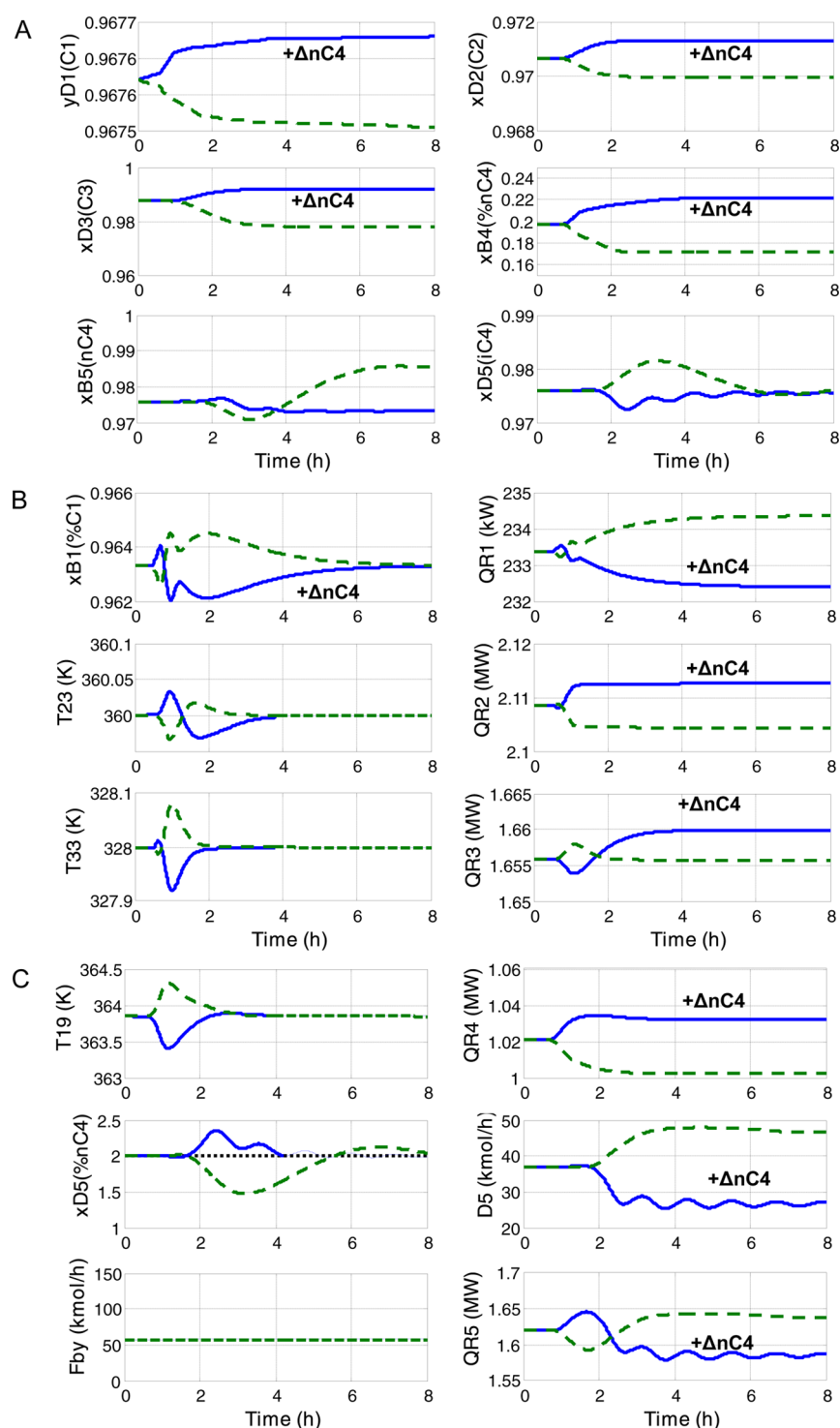


Figure 10. Feed composition C4 disturbances: (A) Product compositions; (B) DM, DE, and DP loop variables; (C) DB and DIB loop variables.

The middle right graph shows that the losses of nC4 in the bottoms from the debutanizer $x_{B4(nC4)}$ remain small. Note the composition is given in mole percent for this impurity.

The bottoms two graphs in Figure 7A show the compositions of the two product streams leaving the de-isobutanizer. The distillate purity $x_{D5(iC4)}$ is not directly controlled at 97.6 mol % iC4, but the impurity of nC4 is directly controlled by the manipulated distillate flow rate. Reflux-drum level is controlled by manipulating reboiler duty and a reflux-to-feed ratio is

maintained. Distillate $x_{D5(iC4)}$ purity stays very close to its specification.

The bottoms purity $x_{B5(nC4)}$ is not directly controlled, but it is maintained fairly close to the normal 97.59 mol % nC4. Parts B and C of Figure 7 show how the three temperature controllers in the de-ethanizer, de-propanizer, and de-butanizer columns respond by manipulating reboiler duties. Also shown are the two composition loops. One in the de-methanizer controls bottoms C1 impurity $x_{B1(C1)}$ by manipulating both reboiler duty and bypass flow (shown in the bottom left graph in Figure 7C)..

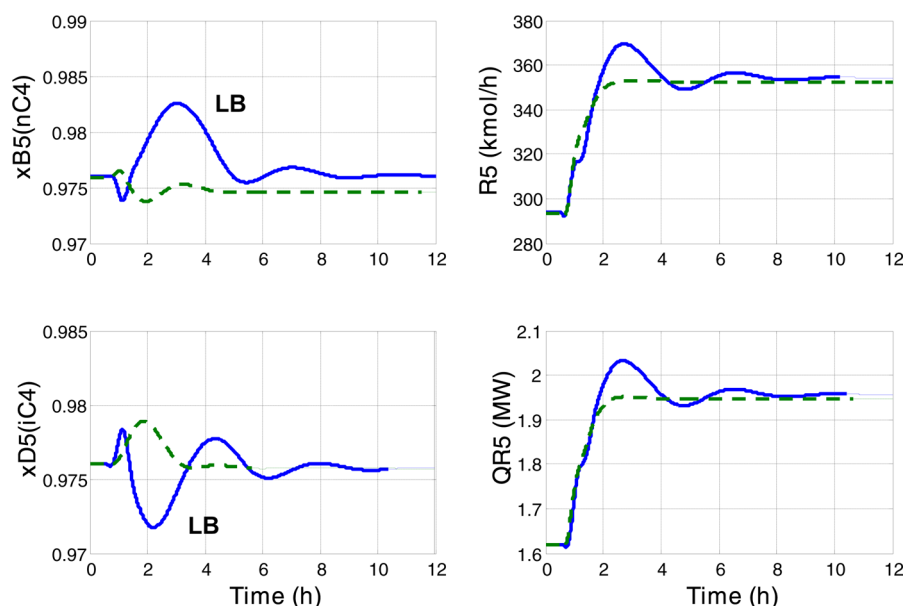


Figure 11. Comparison of proposed and LB control structures: +20% feed flow rate.

The second composition controller is in the de-isobutanizer, which maintains nC4 impurity in the distillate $x_{D4}(nC4)$ by manipulating distillate flow rate.

These results demonstrate that the proposed control structure handles large throughput disturbances quite well.

4.2. Feed Composition C2 Disturbances. Figure 8 gives results when the composition of the feed gas is changed in terms of C2. The solid lines are when C2 feed composition is increased from the design 6.47 to 8.47 mol % C2. The composition of the C1 in the feed is correspondingly reduced. The dashed lines are when C2 feed composition is decreased from the design 6.47 to 4.47 mol % C2. The composition of the C1 in the feed is correspondingly increased. Note that these are large disturbances on a relative basis (31%).

Figure 8A shows that there is a decrease in the C1 composition of the gas product from the de-methanizer when there is more C2 and less C1 in the feed. The purity of the ethane product and the purity of the propane product shift up or down slightly when there is more or less C2 in the feed. The other product compositions are essentially unaffected. Higher C2 feed concentration results in a lower heat input to the de-methanizer reboiler (Figure 8B) and more bypassing (Figure 8C). Reboiler duty in the ethanizer increases.

4.3. Feed Composition C3 Disturbances. Figure 9 gives results when the composition of the feed gas is changed in terms of C3. The solid lines are when the C3 feed composition is increased from 2.87 to 3.37 mol % C3 with a corresponding reduction in the C2 feed composition from 6.47 to 5.97 mol % C2. The dashed lines are when the C3 feed composition is decreased from 2.87 to 2.37 mol % C3 with a corresponding increase in the C2 feed composition from 6.47 to 6.97 mol % C2. Note that these are large disturbances on a relative basis (17%).

Figure 9A shows that there are quite small shifts in the product composition. Figure 9B shows that more energy is required in the de-propanizer reboiler and less in the ethanizer reboiler when C3 feed composition increases. Reboiler duty in the de-methanizer is higher and bypassing is lower (Figure 9C).

4.4. Feed Composition C4 Disturbances. Figure 10 gives results when the composition of the feed gas is changed in terms of iC4 and nC4. The solid lines are when the nC4 feed

composition is increased from 0.82 to 1.02 mol % nC4 with a corresponding reduction in the iC4 feed composition from 0.72 to 0.52 mol % iC4. The dashed lines are when the nC4 feed composition is decreased from 0.82 to 0.62 mol % nC4 with a corresponding increase in the iC4 feed composition from 0.72 to 0.92 mol % C2. Note that these are large 27% disturbances on a relative basis.

Figure 10A shows that the purities of all product streams are maintained quite close to their specifications. Heat duties in all columns except the de-isobutanizer are little affected (Figure 10B). In the de-isobutanizer, reboiler duty and distillate flow rate decrease when there is more nC4 and less iC4 in the feed.

All of these disturbances are effectively handled by the control structures used on the five columns. Stable regulatory plantwide control is achieved with product compositions maintained close to their specifications.

4.5. Alternative De-isobutanizer Control Structure.

Since the control structure on the de-isobutanizer (DIB) is somewhat unusual (reflux-drum level control using reboiler duty), it might be useful to compare its performance with a conventional control structure [L,B] recommended by Horowitz et al.¹⁵

The [L,B] structure is a dual-composition control structure in which distillate impurity $x_{D5}(nC4)$ is controlled by manipulating reflux flow rate and bottoms impurity $x_{B5}(iC4)$ is controlled by manipulating reboiler duty. Reflux-drum level is controlled by manipulating distillate flow rate, and base level is controlled by manipulating reboiler duty. Note that the reflux is ratioed to the feed with the distillate composition controller changing the ratio.

The two composition loops contained 3-min deadtimes and were tuned sequentially using relay-feedback tests and Tyreus–Luyben tuning rules. The resulting integral times are very large (118 min in the x_{D5} loop and 100 min in the x_{B5} loop). Remember that only one composition controller is used in the proposed structure.

Figure 11 gives a comparison of the two alternative control structures for a 20% increase in feed flow rate to the de-methanizer. The LB results show larger peak transients in the purities of the two product streams from the DIB because of the

large integral times in the two composition controllers and the interaction between the two composition loops.

5. CONCLUSION

A control structure for the five-column distillation train in a natural gas liquid refining process has been developed and tested. Effective product-quality control is achieved for the six product streams. Conventional PI controllers provide stable regulatory dynamic responses. Temperature controllers in three of the columns provide effective inferential composition control. Direct composition measurement and control are required in the demethanizer and the de-isobutanizer. The difficult separation in the large, high reflux ratio de-isobutanizer is well-controlled by an unusual control structure that uses reboiler duty to control the reflux-drum level.

Some general comments about designing a control structure for trains of distillation columns may be useful. In this process, all of the columns operate in series with the flows moving downstream from column to column. Therefore each column can be studied in isolation in terms of developing an appropriate control structure and finding stable tuning parameters. Life gets much more difficult when there are recycle streams from downstream column back upstream. A good example is the use of a heavy naphtha absorber to recover propane and heavier components from atmospheric crude unit distillate product while rejecting ethane. In this process, a portion of the heavy naphtha bottoms from the second column is recycled back to the first column. The control structure must be developed considering both columns.

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Notes

The authors declare no competing financial interest.

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