INTERACTIVE CONTROL ENGINEERING
Principles and Applications

Volume 10 Control in Upstream and Midstream Oil and Gas

CHAPTER 10

An NGL Fractionation Train Simulator for Plantwide Pairing and Shared Constraints

Pages 230–252

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How to cite this chapter. "An NGL Fractionation Train Simulator for Plantwide Pairing and Shared Constraints," in *Control in Upstream and Midstream Oil and Gas*, John D. Hedengren, Ed., vol. 10 of *Interactive Control Engineering: Principles and Applications*. Advanced Process Solutions, LLC, 2026, ch. 10, pp. 230–252.

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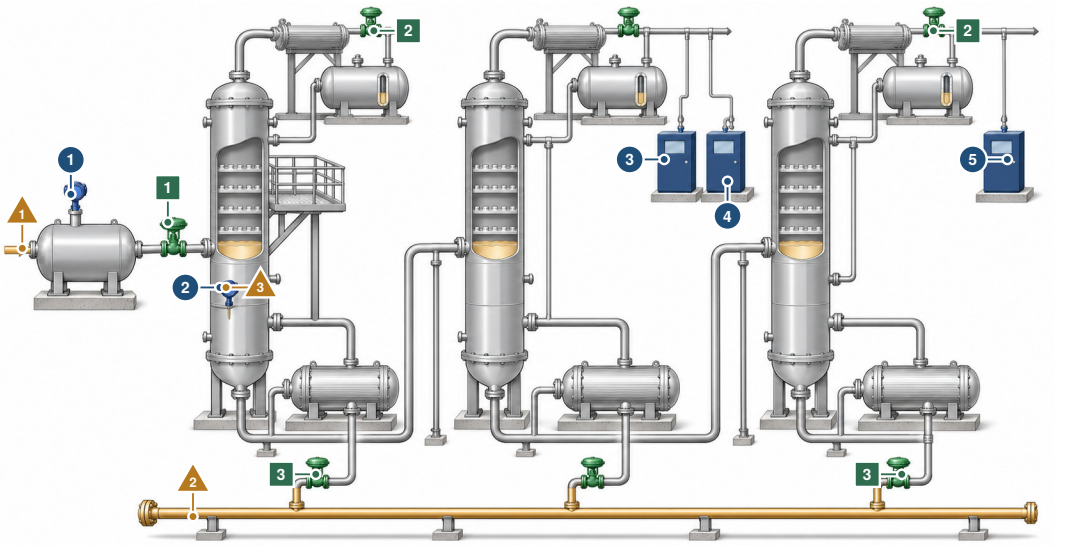
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CHAPTER 10

An NGL Fractionation Train Simulator for Plantwide Pairing and Shared Constraints



The controller moves 1 feed valve, which holds the surge drum level
2 three overhead product valves, which set the cuts
3 three reboiler steam valves, which hold the boilups the operator sets

The controller reads 1 surge drum level 2 first column's temperature, an estimate of the ethane in its bottoms
3 ethane in the propane 4 butane in the propane 5 pentane in the butane

What disturbs it 1 a richer, heavier or larger raw feed 2 a steam header that runs short 3 an estimate that drifts

The analyzer that judges the ethane in the propane sits on the second column's product line, while the valve that sets it is on the first column, and all three reboilers draw on one steam pipe. Illustration, not to scale.

IN PLAIN TERMS Which Column Owns Each Product

The main idea. A gas plant sends a mixed liquid through three tall columns in a row. Each column boils the lightest of what is left off its top. Every product is sold under a limit on the unwanted liquid it may carry.

The challenge. The ethane in the propane is checked at the second column, yet only the first column can take it out. A quick estimate of that ethane can drift without any alarm. The three columns also share one steam supply that can run short.

Why it matters. When nobody decides which product gives way, the one with no controller watching it is the one that fails.

THREE WAYS TO READ THIS CHAPTER

Explore	12 min	This page, the History boxes in Section 10.1 and Section 10.2, the Did you know? boxes in Sections 10.3 and 10.5, the Try it boxes in Section 10.5, and Key points at the end of the Conclusion. No equations.
Study	50 min	The whole chapter in order, stopping at the three pause points, then Exercises 1 to 12.
Practice	20 min	This page, Section 10.5 with Figure 10.4, Section 10.6 and Section 10.7 with its Open challenge box.

The control question Ethane in the propane product is measured on the depropanizer, set by the deethanizer cut one column upstream, and three reboilers share one steam header. Which valve owns each specification, which measurement can see a drifting soft sensor, and which product loses when the header runs short?

Learning objectives

- Show from the shortcut split why the depropanizer cut cannot control ethane in the propane, and compute the distillate that dilution would need.
- Distinguish the cut, which trades one key impurity against the other, from the boilup, which improves both at the price of steam.
- Compare the local, product analyzer and cascade structures on a drifted inferential, and explain why the failure appears as giveaway.
- Predict which specification a proportional cut of the steam header leaves broken, and say what layer would decide instead.
- Judge a predictive controller against a cascade by its depth over a limit as well as its minutes, and trace its deeper surges to its disturbance model.

Prerequisites Sections 1.4.4 (averaging level), 1.4.8 (constraint control), 1.4.9 (late and inferred measurements), 1.6.2 (the framework's PID) and 1.6.4 (the apps). Section 1.6.3 is needed only for the MPC. Chapter 9 makes the liquid this train splits.

The app https://apmonitor.com/apps/ngl-train/

See also Volume 6, Chapter 2 for a pairing decided before any loop is tuned, Volume 6, Chapter 7 for the recycle this train does not have, and Volume 6, Chapter 8 for plantwide structure on a plant with sampled analyzers. Volume 5, Chapter 3 for composition control through a slow analyzer, and Volume 5, Chapter 10 for a soft sensor whose bias a late measurement corrects.

10.1 Introduction

A control course teaches loop pairing on a single unit. Two inputs, two outputs, a relative gain array, and a choice. On a plant the question arrives in a different form and the form is harder: a product specification is written on a stream that leaves one column, and the valve that sets it belongs to a column upstream. On the train modeled here, ethane in the propane product is measured on the depropanizer overhead, and moving the depropanizer cut from 290 kmol/h to 345 kmol/h leaves the ethane flow in that distillate at 48.5 kmol/h at both ends. It does not move at all. The percentage falls from 16.7 to 14.0 mol% purely by dilution, and reaching the 2 mol% contract limit that way would need a distillate of about 2430 kmol/h against a column feed of 918.

The misconception this corrects is that a controlled variable can be paired with any manipulated variable that moves it. Ethane is lighter than the depropanizer's light key, so its recovery to that distillate is one for every cut and every boilup, and the only thing a local loop can do is change the denominator. A relative gain array computed on the depropanizer alone would not say so, because the gain is not zero. It is a dilution gain, and dilution is not control. Volume 6, Chapter 2 shows a second way the array misleads, on a pairing it rates as free of interaction. Tracing a specification back to the unit that makes it, before drawing a single loop, is the first step of every plantwide control design procedure [1, 2], and it is the step a unit-level course has no example for.

The structure generalizes well beyond gas processing. Units in series with specifications made upstream of where they are measured, plus a shared utility header that couples them all, describes refinery crude and vacuum trains, aromatics complexes and most of what an operator means by the plant rather than the column. Plantwide design procedures exist for exactly this class [1, 3, 4] and the constraint arbitration that a real train delegates to a multivariable controller [5] is deliberately left out of the loop structures so that its absence can be observed. An optional model predictive controller plans the three cuts against their contract limits but leaves the steam header to its proportional rule, so the header arbitration stays absent under it as well. This chapter studies an interactive three column

NGL train whose separation factors, fractionation coefficients and analyzer dynamics are hidden, so that the pairing must be discovered from experiment and a shared constraint resolves itself by arithmetic rather than by judgment. The shortcut split, ratio control and plantwide design procedures are established methods. The train is a shortcut split model with its instrument chain, three quality architectures and a steam header allocation rule, written out in the detail needed to rebuild it. Its scenarios build to a boiler trip whose casualty is the specification nobody chose to sacrifice, and every run exports its complete time history. An optional model predictive controller of a standard linear form is measured against the cascade on the same instruments, and the comparison is reported as a negative result with its mechanism.

Section 10.2 follows NGL fractionation and its control from the shortcut design methods to the plantwide procedures. Section 10.3 states the train model, its shortcut split and the shared header, and Section 10.4 the loops, the three ethane architectures and the predictive controller. Section 10.5 measures the scenarios, from the transport of one disturbance to the MPC against the cascade. Section 10.6 sets the hardware of a real train beside the model, Section 10.7 lists what the model leaves out, Section 10.8 draws three general lessons, Section 10.9 sums up, and the exercises in Section 10.10 put the scenarios to work.

> ⧖ HISTORY **1964 to 1973: Control Structure Becomes the Question**
>
> A textbook of 1964, *Techniques of Process Control*, gave the first comprehensive discussion of plantwide control, in a chapter on overall process control that treated material balance control, production rate control and buffer tanks used as low-pass filters. In 1973 a critique of process control theory said that the central issue for new theories was the control structure: which variables to measure, which inputs to manipulate and which links to make between the two. A review of 2000 found industrial practice still close to the 1964 chapter, and noted that even a plant with no recycle needs the plantwide view, because in a string of units in series each one disturbs the next [6]. The train of this chapter is such a string.

10.2 Background

Raw natural gas liquids arrive from a gas plant, such as the turboexpander plant of Chapter 9, as a mixture of ethane, propane, butanes and a C5-plus cut called natural gasoline. Three columns in series take it apart, each one's bottoms being the next one's feed, and each product is sold against a contract written in mole percent. Nothing recycles, which makes the train tractable: one pass down it is the steady state.

The industry that produces these streams grew with the gas transmission network and with petrochemical demand for light feedstock. Ethane recovered from natural gas became the dominant steam cracker feed in North America, and the economics of recovering it rather than leaving it in the sales gas have driven gas plant design for decades [7]. Growth in gas processing elsewhere has followed the same pattern more recently [8]. The fractionation train itself is old technology, and its design methods are older still: the Fenske minimum stage count [9], the Underwood minimum reflux [10] and the Gilliland correlation joining them [11] are the shortcut sequence a designer still uses to size these columns before any rigorous simulation is run, and shortcut methods remain a subject of active development [12].

Control developed on a different track. The individual columns were run first on tray temperatures and manual cut adjustments, with laboratory samples confirming the contract. Process chromatographs made continuous product measurement possible and made the modern quality loop practical, but a chromatograph on the propane product measures a specification whose valve is on another column, and the retrofit that follows is a plantwide cascade rather than a local loop. The plantwide control literature then formalized that step, as the History box at the end of this section tells. Heuristic rules for the same decisions were collected in parallel [4]. Soft sensors entered the same trains as local inferentials, usually pressure-compensated tray temperatures, and the review literature on their maintenance and drift is directly relevant to what scenario 6 of this application shows [13], and

Table 10.1 Classification of Variables at the Design Operating Point

Symbol	Role	Nominal	Range
x_{C2} in propane	CV, analyzer	1.80 mol%	0.2–3.0 SP
x_{C4} in propane	CV, analyzer	2.30 mol%	0.5–4.0 SP
x_{C5} in butane	CV, analyzer	1.90 mol%	0.4–4.0 SP
$\hat{x}_{C2,B}$	CV, inferential	0.50 mol%	0.05–4.0 SP
h (V-101)	CV, level	50 %	20–80 %
D_1/F_1	MV	8.16 %	2–25 %
D_2/F_2	MV	35.14 %	20–55 %
D_3/F_3	MV	49.80 %	30–72 %
$(V/F)_{1,2}$	MV, economic	1.80, 1.90	0.8–3.2
$(V/F)_3$	MV, economic	1.40	0.6–2.6
F_0	DV, measured	1000 kmol/h	600–1400 kmol/h
z_{C2}	DV, unmeasured	0.08	0.02–0.20
V_{max}	DV, measured	5200 kmol/h	3000–7000 kmol/h
inferential bias	DV, unmeasured	1×	0.5–2×

Section 1.4.9 places them among the other inferred and late measurements of this volume. Constraint arbitration under a shared utility header arrived last, and arrived as multivariable control rather than as a regulatory structure [5], which is why a train without it behaves the way Section 10.5 reports.

> HISTORY **1997 and 2000: Procedures for the Whole Plant**
>
> In 1997 a heuristic design procedure set out to give a control structure for an entire flowsheet rather than for single units. Its nine steps centered on energy management, production rate, product quality, constraints, inventories, component balances and optimization [1]. The first steps establish the control objectives and count the control degrees of freedom by the independent valves, and the quality step recommends the usual rule of pairing close [6], a rule the ethane in this chapter's propane puts to the test. In 2000 self-optimizing control asked which controlled variables, held at constant setpoints, keep operation near the optimum with an acceptable loss [2].

10.3 Process Model

The train takes 1000 kmol/h of raw NGL at 8 mol percent ethane, 32 percent propane, 30 percent butanes and the balance natural gasoline. A surge drum feeds deethanizer T-101, whose bottoms feed depropanizer T-102, whose bottoms feed debutanizer T-103. Four products leave under five contract limits, two of them on the propane. Figure 10.1 shows the arrangement and every symbol used in the equations.

> DID YOU KNOW? **The Products of One Train, by the Million Barrels**
>
> In 2025 the natural gas processing plants of the United States produced an average of 2.3 million barrels a day of propane, 0.7 million of normal butane, 0.5 million of isobutane and 0.9 million of natural gasoline [14]. A train like this chapter's makes all of them, with the two butanes lumped into one product.

Table 10.1 classifies the variables. Flows are in kmol/h, compositions in mole percent and time in minutes. Nineteen states are integrated: four component holdups in the surge drum, four distillate component flows per column, each a first-order lag on its shortcut split, and three reboiler thermal lags. The bottoms of each column follow by difference from its feed. The analyzer samplers, their averaging filters and the local inferential lag are discrete states updated at each controller scan. A case starts at the steady state of the settings in force, with the drum at its level setpoint, the boilups at what the header passes and every instrument settled on its start-up reading.

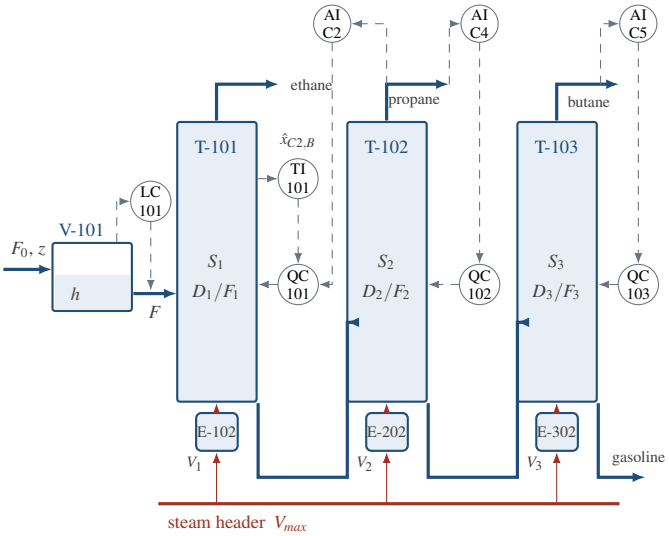


Figure 10.1 NGL fractionation train. Raw liquids F_0 of composition z enter surge drum V-101 at level h and leave as train feed F. Deethanizer T-101, depropanizer T-102 and debutanizer T-103 each carry a separation factor S_i set by the boilup ratio and a cut ratio D_i/F_i that places the material balance. Reboilers E-102, E-202 and E-302 draw boilup V_i from one steam header of capacity V_{max}. LC-101 holds the drum level with the train feed. AI-C2 and AI-C4 are the propane product chromatographs and AI-C5 the butane product chromatograph, TI-101 is the deethanizer local inferential $\hat{x}_{C2,B}$, and QC-101 to QC-103 manipulate the three cut ratios, QC-101 from TI-101, AI-C2 or both in cascade. The ethane specification is measured on the T-102 overhead and set by the T-101 cut.

10.3.1 The Shortcut Split

Each column is represented by two numbers, a separation factor S that says how sharply it fractionates and a cut rate D that says where the material balance is placed. Everything lighter than the light key goes entirely overhead, everything heavier than the heavy key goes entirely to the bottoms, and the two keys share according to Equation (10.1).

$$\frac{r_L}{1-r_L} = S\,\frac{r_H}{1-r_H} \tag{10.1}$$

where r_L and r_H are the recoveries of the light and heavy keys to the distillate. The equation is solved by bisection on r_H subject to the overall material balance $D = \sum_i r_i F z_i$.

The first sentence of that paragraph contains the whole plantwide lesson. Ethane is lighter than the depropanizer's light key, so $r_{C2} = 1$ for that column, always, at every cut and every boilup.

10.3.2 Fractionation and Energy

The separation factor is built from a stage count and an efficiency that saturates in boilup, Equation (10.2).

$$S = \alpha^{N\eta}, \qquad \eta = 1 - \exp\left[-k\left(\frac{V}{F} - \left(\frac{V}{F}\right)_0\right)\right] \tag{10.2}$$

The coefficients k_i are calibrated so that each column reproduces its design split at its design boilup ratio, and $(V/F)_0 = 0.50$ is the boilup below which nothing separates. Because η saturates, the first increment of steam buys a great deal of separation and the fourth buys very little, which is the economic content of the boilup setpoint.

Table 10.2 Model Parameters

Symbol	Meaning	Value	Class
α_1	C2/C3 volatility	2.10	adjustable
α_2	C3/C4 volatility	2.40	adjustable
α_3	C4/C5 volatility	2.60	adjustable
N_i	theoretical stages, each	25	adjustable
k_1, k_2, k_3	fractionation coefficients	0.4703, 0.3384, 0.4587	hidden
$(V/F)_0$	boilup with no separation	0.50	adjustable
τ_1	deethanizer composition	25 min	adjustable
τ_2	depropanizer composition	30 min	adjustable
τ_3	debutanizer composition	45 min	adjustable
θ	analyzer cycle	10 min	adjustable
x_{C2}^{max}	ethane in propane	2.0 mol%	visible
x_{C4}^{max}	butane in propane	2.5 mol%	visible
x_{C5}^{max}	pentane in butane	2.0 mol%	visible
$x_{C4,gas}^{max}$	butane in gasoline (RVP)	2.0 mol%	visible
x_{C3}^{max}	propane in ethane	5.0 mol%	visible

10.3.3 Cut Against Boilup

The two handles are not interchangeable and the difference is measurable. On the deethanizer, six more kmol/h of cut takes the ethane left in the bottoms from 1.60 kmol/h to 0.62 kmol/h and pushes the propane lost overhead from 3.20 kmol/h to 8.22 kmol/h, which is a trade along a fixed curve. Raising the boilup ratio from 1.6 to 2.2 at a fixed cut takes the ethane from 2.97 kmol/h to 0.45 kmol/h and the propane from 4.57 kmol/h to 2.05 kmol/h, so both improve, which no cut change can do. That asymmetry is why the quality loop goes on the cut and the boilup stays an economic setpoint.

🕮 **Pause.** Try Exercise 10.2 before reading on.

10.3.4 Why the Manipulated Variable Is a Ratio

The quality loops manipulate D/F rather than D. An 18 percent rate increase moves the correct cut flow of every column by 18 percent and leaves the correct cut ratio unchanged at a fixed feed composition, because the split depends only on the cut and boilup ratios, so a ratio station removes the disturbance before any controller sees it. This is feedforward written into the pairing rather than added beside it, and it is the same argument the reflux ratio makes on a superfractionator at a different scale.

10.3.5 The Shared Header

Three reboilers draw on one steam header of capacity V_{max}. When the sum of the three demands exceeds it, every boilup is multiplied by the same factor, Equation (10.3).

$$\phi = \min\left(1,\ \frac{V_{max}}{V_1 + V_2 + V_3}\right), \qquad V_i^{app} = \phi\, V_i \tag{10.3}$$

Nothing in this rule decides which column should give way. That is deliberate, and Section 10.5 reports what the plant does about it.

10.3.6 Assumptions

Each column is a shortcut split with a single separation factor rather than a stage-by-stage model, so tray hydraulics, flooding and efficiency loss at turndown are absent. The column responds as a first-order lag on composition, which represents the dominant time constant and not the shape of the approach.

The train has no recycle. Real gas plants return some streams and a depropanizer overhead is sometimes recycled for reprocessing, which would change the plantwide analysis considerably.

Component lists are lumped into four pseudo-components, so the isobutane and normal butane split that a real debutanizer has to manage is invisible.

The steam header allocation is a proportional cut with no priority. Real plants carry a load shedding scheme or a multivariable controller that decides which column to starve, and the absence of one here is what scenario 8 exists to show.

The natural gasoline Reid vapor pressure specification is represented by a butane mole fraction limit rather than by a vapor pressure calculation.

Analyzers are exact samplers with a cycle time and an averaging filter, and the inferential carries an explicit bias fault. Where a case adds analyzer repeatability noise, the noise enters after the sampler, so a published result wanders between samples (0.030 mol% RMS about the held sample in the challenge) where a real chromatograph reports one value per cycle. A real chromatograph has more ways to be wrong than a gain error.

10.4 Controller Design

10.4.1 Velocity Form PID

Every loop uses the framework's velocity-form PID of Section 1.6.2, Equations (1.16) to (1.19), with the derivative taken on the measurement, executed every $\Delta t_c = 0.2$ min. The integral is seeded with the output in force when the case starts, and the tracking time of the back-calculation is $T_t = \tau_I$, so each scan at a limit removes $\Delta t_c / \tau_I$ of the clipped excess from the integral alone. Clamping the running output instead would rectify symmetric analyzer noise into a one-way drift away from whichever limit the output rests against. No cut ratio and no setpoint written by the trim reaches its limit in any of the ten original cases as shipped or in the cascade runs of Section 10.5.7, and under the predictive controller the loops only track the cuts it applies, so the tracking term does not act in any result reported here. In manual the integral is set so that the output equals the operator's cut ratio exactly, which makes the transfer to automatic bumpless. Derivative action is not used on any loop here ($\tau_D = 0$), because the measurements are sampled and a derivative on a staircase is a sequence of impulses [15]. Integration is fixed-step fourth-order Runge-Kutta at 0.05 min on the engine of Section 1.6.1, so every run reproduces exactly.

10.4.2 The Three Ethane Architectures

The ethane specification selector offers four closed-loop configurations besides off. Comparing the three below is the subject of Section 10.5, and the fourth, the model predictive controller of Section 10.4.3, is compared with the cascade in Section 10.5.7.

Local closes a loop from the deethanizer inferential, a lagged estimate of ethane left in the bottoms, to the deethanizer cut ratio, with action $a = -1$ in Equation (1.16) because more cut is less ethane left behind, and $K_c = 1.20$ % of feed per mol% and $\tau_I = 28$ min.

Analyzer closes a loop from the propane product chromatograph, one column downstream, directly to the deethanizer cut ratio, also with $a = -1$, $K_c = 0.60$ % of feed per mol% and $\tau_I = 120$ min. It is slow because it waits for a composition to travel a column and an analyzer to sample it. Volume 5, Chapter 3 tunes a composition loop against the same kind of slow analyzer on a single column.

Cascade puts the analyzer on the setpoint of the local loop, with $a = +1$ on the setpoint, $K_c = 0.40$ and $\tau_I = 240$ min inside 0.05 mol% to 4.0 mol%, which keeps the fast local response and removes the inferential's offset. The trim is not told when the inner loop's cut reaches its limit.

The butane in propane loop manipulates the depropanizer cut, with $a = +1$, $K_c = 0.50$ and $\tau_I = 40$ min, and the pentane in butane loop manipulates the debutanizer cut, with $a = +1$, $K_c = 0.60$ and $\tau_I = 45$ min. These are the tunings of scenarios 4 to 9. The surge drum level loop, $K_c = 25$ kmol/h per percent and $\tau_I = 90$ min, is detuned so that the drum absorbs rate changes rather than passing them to three columns at once, the averaging level control of Section 1.4.4.

10.4.3 Model Predictive Control

A fifth setting of the ethane selector, never the default and loaded only by scenario 10, hands the cut ratios to one model predictive controller (MPC), a plant-specific form of the framework's option of Section 1.6.3. Once a minute ($T_s = 1$ min, every fifth controller scan) a quadratic program plans the deethanizer cut ratio $u_1 = D_1/F_1$, and $u_2 = D_2/F_2$ and $u_3 = D_3/F_3$ whenever the butane and pentane loops are in automatic, and the cuts take the plan's first move and keep it for that minute, as in the receding-horizon practice of industry [5]. A cut whose loop is in manual stays where the operator set it, and its impurity leaves the objective. The ratio stations, the cut ranges and the header rule of (10.3) act on its moves as on the loops' outputs, and it moves no boilup. Predictive control has been evaluated on an industrial depropanizer [16], and the control of a whole NGL train has been studied [17].

The controller's model is identified on the application. Each cut ratio was stepped by ± 0.1 percent of feed on the deethanizer and ± 0.2 on the other two, in manual with the noise off, where the products sit on their targets of 1.80, 2.30 and 1.90 mol% (cut ratios of 7.489, 35.62 and 49.79 %), and each 480 min response of the logarithm of an impurity was fitted by least squares on first-order lags at the declared time constants of Table 10.2. Logarithms are used because a cut changes an impurity by a factor, as the inferential bias does. The deethanizer cut reaches the downstream products only through its bottoms as the inferential reads them, because the same cut leaves different bottoms on a different feed. The result is Equation (10.4).

$$
\begin{aligned}
\tau_m \frac{\mathrm{d}a_{jm}}{\mathrm{d}t} &= u_j - u_j^0 - a_{jm}, \qquad \tilde{y}_B = \textstyle\sum_{m=1}^{3} \gamma_m a_{1m} + d \\
\tau_2 \frac{\mathrm{d}p}{\mathrm{d}t} &= \tilde{y}_B - p, \quad \tau_3 \frac{\mathrm{d}q}{\mathrm{d}t} = \tilde{y}_B - q, \quad \tau_3 \frac{\mathrm{d}r}{\mathrm{d}t} = p - r \\
\tilde{y}_{C2} &= 1.0137\,p - 0.0281\,a_{22} \\
\tilde{y}_{C4} &= -0.4214\,p + 0.8195\,a_{22} + 0.0081\,a_{23} \\
\tilde{y}_{C5} &= -0.0827\,p - 0.0786\,q - 0.0592\,r \\
&\quad - 0.9079\,a_{22} + 1.3635\,a_{23} + 0.5566\,a_{33}
\end{aligned}
\tag{10.4}
$$

where a_{jm} is cut j through the lag τ_m, d is a disturbance at the deethanizer bottoms held constant over a plan, $\gamma = (-1.3673,\ -0.0855,\ 0.0137)$, p, q and r carry the bottoms downstream, and $\tilde{y}_B$, $\tilde{y}_{C2}$, $\tilde{y}_{C4}$ and $\tilde{y}_{C5}$ are the model's changes in the logarithms of $\hat{x}_{C2,B}$, x_{C2}, x_{C4} and x_{C5} of Table 10.1. The coefficients on a_{jm} are in natural logarithm per percent of feed and those on p, q and r are dimensionless, a term of $-6 \times 10^{-7}\,a_{23}$ in $\tilde{y}_{C2}$ is omitted, and the model is anchored at the first readings of a run, which it takes as a steady state, so u_j^0 is the cut ratio at $t = 0$. Recomputed by central differences on the declared train, the steady log gains differ from the model's by at most 0.2 %. Free fits to one-sided steps of 0.1 % of feed return 25.8 min from the deethanizer cut to the inferential, 29.3 min from the depropanizer cut to the butane and 44.2 min from the debutanizer cut to the pentane, against the declared 25, 30 and 45, and the model's own responses depart from these steps by 0.1 to 6.1 % RMS of their span. At the design cut ratios, with 0.50 mol% ethane in the propane, the deethanizer cut moves the ethane 1.41 times as hard as this linear model says.

The model is stepped at every scan with the cut ratios applied. The inferential updates d at every scan, and each analyzer result updates an output bias β_i on its own product, as shown in Equation (10.5).

$$
\begin{aligned}
d &\leftarrow d + \lambda_I \left(\ln \hat{x}_{C2,B} - \ln \hat{x}^0_{C2,B} - \tilde{y}_B \right) \\
\beta_i &\leftarrow \beta_i + \lambda_A \left(\ln x_i^{an} - \ln x_i^0 - \tilde{y}_i(t_s) - \beta_i \right)
\end{aligned}
\tag{10.5}
$$

where $\lambda_I = 0.2$ per scan, $\lambda_A = 0.5$ per result, a superscript 0 marks a first reading, and $\tilde{y}_i(t_s)$ is the model's prediction recorded when the sampler drew the sample whose result x_i^{an} is read 7.5 min later, once the 1.5 min averaging has settled. Compared at the moment of reading instead, the ethane in

scenario 10 would peak at 3.60 rather than 3.28 mol%. The analyzer biases find a drifted inferential, taking up half of the 25 percent drift of scenario 10 in 47.8 min and 90 percent in 97.9 min. The estimate of product i is $\hat{y}_i = \ln x_i^0 + \tilde{y}_i + \beta_i$, with one integrating disturbance per measured signal, the number shown sufficient for offset-free control [18], so the analyzer dead time lives in the estimator and the plan predicts the true product that the contract judges.

Each decision minimizes Equation (10.6) over $N = 120$ steps, a horizon of 120 min, with the moves held in eight blocks of 1, 2, 3, 5, 8, 13, 20 and 68 steps, a program of 24 moves and three slacks.

$$\begin{aligned} \min_{u,\varepsilon} \; & \sum_{k=1}^{N} \sum_{i} w_i \left(\hat{y}_{i,k} - \ln x_i^*\right)^2 + \sum_{b=1}^{8} \sum_{j=1}^{3} S_j \, g_j^2 \, \Delta u_{j,b}^2 \\ & + \sum_{i} \left(\rho_2 \, \varepsilon_i^2 + \rho_1 \, \varepsilon_i\right) \\ \text{s.t. } & u_j^{lo} \le u_{j,b} \le u_j^{hi}, \qquad \varepsilon_i \ge 0 \\ & \hat{y}_{i,k} \le \ln x_i^{max} + \phi_{i,k} + \varepsilon_i, \quad k = 1, 6, 11, \ldots \end{aligned} \tag{10.6}$$

where $\hat{y}_{i,k}$ is the predicted logarithm of product i at step k, x_i^* is the learner's target clipped to the contract limit x_i^{max} of Table 10.2, w_i is 1 when the plan moves the impurity's own cut and 0 otherwise (its band is then dropped), $\Delta u_{j,b}$ is the move of cut j into block b, the first measured from the cut applied now, $g = (1.459,\ 0.828,\ 0.557)$ is each cut's steady log gain on its own impurity, $S_1 = 20000$, $S_2 = S_3 = 1$, $\rho_1 = 10^4$ and $\rho_2 = 10^5$. The bounds are the ratio stations' ranges of Table 10.1 and are hard, and an optional deethanizer rate limit is off as shipped. The contract limits are soft constraints, and the shared solver carries a single slack for each output over the whole horizon. The funnel $\phi_{i,k} = \max(0,\ \hat{y}_{i,k}^h - \ln x_i^{max})(1 - s_{i,k})$, with $\hat{y}^h$ the prediction with the cuts held and $s_{i,k}$ the normalized step response of impurity i to its own cut, lets an excess already on its way decay as fast as that cut could remove it. Without the funnel the plan spends 750 min over the ethane limit in scenario 10 instead of 153.

Tracking the targets replaced a linear price on giveaway, which with one slack per impurity either paid the plan to ride just over a limit or, priced to forbid that, left no move price that both approached a target and rode out a surge. Giveaway is therefore measured, not priced. The slack is still bought against the move price, in 130 of the 1440 plans of scenario 10, all between 210 and 565 min, none above 0.031 in the logarithm and none with a cut at a bound, and the first of them re-solved with the deethanizer move priced a hundredth as much buys none. The prices were chosen once, on scenario 10, as the least summed loop IAE with no more cut travel and no more minutes over any limit than the shipped cascade, then the largest prices within 10 percent of that IAE.

The controller keeps a plan only if the solver reports success with an optimal status, every number it returns and every prediction converted back to mole percent is finite, the slacks are non-negative, and every planned move over the horizon, not only the first, lies inside its range. Otherwise the cascade and the two downstream loops, which track the applied cuts at every scan with no preload, take over until a plan is accepted again. With the noise off, forced rejections at five instants of scenario 10 hand over within 0.017 % of feed, and none of the measured runs rejected a plan. A switch to the MPC in mid-run is the plan's first decision rather than a bump, but it can move the debutanizer cut by 0.72 % of feed at once, about twice the largest move between consecutive plans of the day. The first plan is solved at $t = 0$. In Chrome a plan takes 2.1 ms, which occupies 0.21 % of a processor core when the run plays one simulated minute per second and 25 % at the top speed of 120 simulated minutes per second. The plant on the run's parameters, the instruments and the controller's identified model are three separate models, and replaying the 7201 measurement packets of scenario 10 in front of four different hidden plants reproduces all 1439 plans bit for bit.

Table 10.3 Depropanizer Cut Swept on a Rich Feed. The Ethane Flow in the Distillate Does Not Move.

Quantity	Cut 290 kmol/h	Cut 345 kmol/h
Ethane in distillate (kmol/h)	48.5	48.5
Ethane in distillate (mol%)	16.7	14.0
Butane dragged overhead (kmol/h)	0.24	0.95

10.5 Results

Eleven scenarios run from an uncontrolled disturbance to a graded challenge. Scenario 1 follows one disturbance down the train in manual, 2 shows that the unit that measures the ethane cannot fix it, and 3 sets the cut against the boilup. Scenarios 4 to 6 build the local, product analyzer and cascade architectures, 7 puts a number on giveaway, 8 trips the steam header, 9 lands a heavier feed and then a rate step, 10 sets the MPC against the cascade and 11 is the graded challenge. Each is credited by a completion rule, as Section 1.6.4 describes, and the exercises of Section 10.10 are built on them. Every run below is the case as it loads, with any setting named in the text changed before the run starts. Reset restarts a case from $t = 0$ with the learner's settings kept (Section 1.6.1), so scenarios 5, 6 and 10 reach their comparison runs by Reset, with the architecture changed before the scripted disturbance arrives.

10.5.1 Following a Disturbance Down the Train

Stepping the raw NGL from 8 to 13 mol% ethane at 30 min with nothing controlling gives the transport structure directly. On the run clock the deethanizer inferential is 30 percent of the way to its new value at 60 min and half way at 90. The propane product is 10 percent at 60 min, 28 percent at 90 and half way at 129. With no loop closed the ethane in the propane finishes at 15.0 mol% against a limit of 2.0. The first hour belongs to the surge drum rather than to the deethanizer. The bottoms follow the feed by difference, so the inferential moves as soon as the drum composition does, and 1500 kmol of drum holdup at 1000 kmol/h is a 90 min mixing lag. With the drum capacity at its 500 kmol minimum, set before Run, the inferential is half way 8.9 min after the step instead of 60. The further 39 min the product takes is mostly the depropanizer's 30 min composition lag.

> **TRY IT IN THE APP** **1 · Follow one disturbance down the train (MANUAL)**
>
> Load the case and press Run. At 30 min the raw liquid turns richer in ethane and nothing on the plant moves. Watch the ethane trend: the deethanizer inferential moves first and is half way to its new value at 90 min, the propane product follows and is half way at 129, and the ethane in the propane finishes at 15.0 mol% against a limit of 2.0.

10.5.2 Where the Specification Is Made

Table 10.3 reports the depropanizer cut ratio stepped from 31.6 to 37.6 % of its feed on an already rich feed of 13 mol% ethane, as scenario 2 asks.

The values are the steady state at each cut. A cut chosen before Run starts the case at its steady state, so the scenario's first 90 min hold reads 16.7, and the second, after the step to 37.6 %, reads 14.2 mol%. The ethane flow reads 48.5 kmol/h at every step of the run. The percentage falls only by dilution, and reaching 2 mol% that way would require a distillate of about 2430 kmol/h against a column feed of 918. The valve that sets this specification is the deethanizer cut, one column and about an hour upstream.

Pause. Try Exercise 10.1 before reading on.

Table 10.4 Ethane in Propane With a Drifted Inferential, Target 1.80 mol%

Architecture	15 h (mol%)	30 h (mol%)
Local inferential alone	0.705	0.705
Cascade	1.595	1.751
Analyzer alone	1.856	1.800

10.5.3 Architecture Audit

Scenario 6 drifts the deethanizer inferential 30 percent low at 30 min and steps the feed to 12 mol% ethane a minute later. Table 10.4 reports the ethane in the propane product against a 1.80 mol% target for each architecture, selected before the run. The scenario's own route to the local run, Reset and then Local, gives the same 0.705.

The direction matters more than the magnitude. A soft sensor reading low makes the local loop let more ethane through than its setpoint intends, but the 0.50 setpoint sits so far inside the 1.80 mol% target that the product still settles below it. The failure appears as giveaway rather than as an alarm, and a plant can run that way for months. The cascade recovers most of the error because the analyzer above it corrects the inferential setpoint, moving it from the shipped 0.50 to 1.24 mol% by 30 h. The trim is deliberately slow, and the product reaches 1.80 ± 0.01 mol% only 47 h into the run. The analyzer alone is exact because nothing biased stands between it and the product. What the local loop buys in exchange is speed, and what it costs is a measurement nobody is checking, which is the standard soft sensor maintenance problem [13].

Pause. Try Exercise 10.3 before reading on.

> **DID YOU KNOW? The Slow Loop Was Closer**
>
> From the simulator (Table 10.4). At 15 h into the drifted-inferential case, the product analyzer alone, the slowest of the three structures, held the propane at 1.856 mol% against the 1.80 mol% target, while the cascade read 1.595 and the local loop 0.705. At 30 h the analyzer alone read exactly 1.800.

10.5.4 Giveaway as a Quantity

Running at 0.60 mol% ethane and 1.00 mol% pentane against limits of 2.0, then stepping both targets up at 240 and 245 min, adds 3.39 kmol/h of propane and 6.30 kmol/h of butane four hours later, which is about half a tonne an hour. By then the slow trim has taken ethane in the propane only to 1.11 mol%, and the propane gain grows to 6.46 kmol/h as ethane approaches its target over the next three days. The steam consumption moves from 4380 kmol/h to 4387 kmol/h, which is to say it does not change. Giveaway on this train is not paid for in energy. It is paid for in product that left in the wrong stream, and it is invisible on every trend that has an alarm on it. The low pentane target also held butane in the natural gasoline at 2.74 mol% at 240 min, above its 2.0 limit, so the careful-looking operation was already off specification on the stream whose specification has no loop.

> **TRY IT IN THE APP 7 · What running below the limit costs**
>
> Press Run. The cascade holds ethane in the propane at 0.60 and pentane in the butane at 1.00 mol%, and at 240 and 245 min both targets step up. Four hours later the propane rate is up 3.39 kmol/h and the butane 6.30 kmol/h, while the Steam badge moves only from 4380 to 4387. Butane in the natural gasoline already read 2.74 mol% at 240 min, over its 2.0 limit.

10.5.5 The Shared Header

Tripping a boiler at 60 min multiplies every boilup ratio by 0.912, and the header stays short for the rest of the run. Table 10.5 reports the state 15 h into the run.

Table 10.5 Steam Header Trip From 5200 kmol/h to 4000 kmol/h. The Specification Without a Controller Is the One That Fails.

Product quality	Peak	15 h	Limit	Loop
Ethane in propane (mol%)	1.49	1.49	2.0	yes
Butane in propane (mol%)	2.56	2.29	2.5	yes
Pentane in butane (mol%)	2.36	1.90	2.0	yes
Butane in gasoline (mol%)	3.81	3.81	2.0	no
Propane recovery (%)	—	97.1	—	—

Propane recovery falls from 98.3 to 97.1 %. The three controlled impurities are inside their contracts at 15 h, but two of them leave on the way while the cut controllers move the material balances to make up for the lost separation. Butane in the propane spends 65 min above its 2.5 mol% limit and pentane in the butane 194 min above 2.0. Ethane is still climbing toward its target under the slow trim. Butane in the natural gasoline crosses 2.0 mol% 45 min after the trip, climbs to 3.81 mol% and stays there. The debutanizer defends the specification that has a controller on it, using the only cut it has, and the specification on the other side of the same cut is the casualty. Nothing in the control system chose that outcome. The header cut everybody by the same factor and each loop then did what it was told, and the result is the specification nobody would have picked to sacrifice. Resolving that properly requires a layer that knows the relative value of the four products, which on a real train is a multivariable controller [5] and on this one is deliberately absent.

10.5.6 The Graded Challenge

The challenge runs a 1400 min shift with four disturbances, with noise on the analyzers and the inferential. Raw NGL ethane steps from 8 to 12.5 mol% at 60 min, the deethanizer inferential drifts to 0.75 of its reading at 240 min, butanes step from 30 to 33.5 mol% at 460 min, and the rate rises from 1000 to 1150 kmol/h at 700 min. The case loads on the local inferential alone with both downstream loops in manual and a cautious tuning. It is met when ethane in the propane stays within 0.30 mol% of target and butane in the propane within 0.20 mol% from 924 min to the end, with all five product limits met at 1400 min. It is judged on the case as given, and changing a model parameter or a disturbance voids the attempt.

With the shipped line-up (Local, both downstream loops in Manual, cautious tuning) the worst errors over the scored window are 0.74 and 1.79 mol%, and butane in the natural gasoline ends at 22.9 mol%, so the case fails on all three counts. With the suggested line-up (Cascade, both loops in Auto, the tunings of Section 10.4, boilup ratios 1.95, 2.00 and 1.80) they are 0.22 and 0.09 mol%, inside the bands by 0.08 and 0.11, with every product inside its limit, whether the line-up is chosen before Run or carried through a Reset. Over eight other seeds of the analyzer and inferential noise the suggested ethane error ranges from 0.21 to 0.25 mol% and every run passes. The suggested boilups ask the header for about 5406 kmol/h after the rate step, so every reboiler is trimmed by 4 % and the case still passes.

10.5.7 Three Cuts Planned Together

Scenario 10 starts at the design cut ratios, 0.50 mol% ethane in the propane against a 1.80 mol% target, with analyzer and inferential noise of 0.04 and 0.03 mol%. Raw NGL ethane steps from 8 to 12 mol% at 180 min, the inferential drifts 25 percent low at 600 min and the butanes step from 30 to 33.5 mol% at 1020 min. The MPC runs as the case loads, and the cascade on the tunings of Section 10.4 with its inferential setpoint at 1.82 mol%, selected before the run as the scenario's Reset route does. Table 10.6 compares the two with three cascade variants and the MPC restricted to the deethanizer cut, and then the MPC with the cascade, or the challenge's shipped local loop, on scenarios 5, 6, 7 and 9 and the challenge, always on the settings of scenario 10 without retuning.

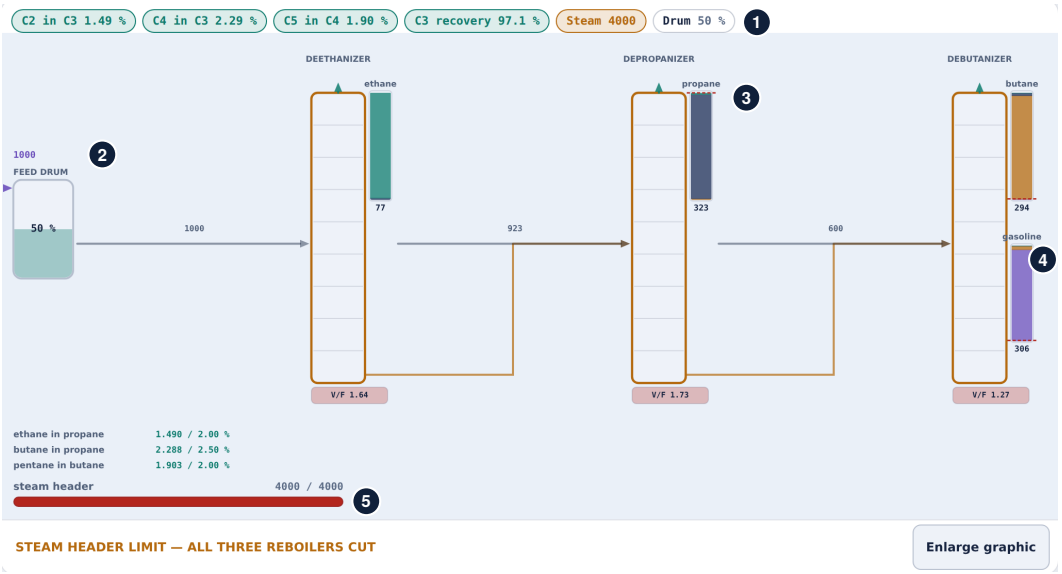


Figure 10.2 The NGL fractionation train application running scenario 8, the steam header trip, at $t = 15.00$ h with the ethane cascade and both downstream quality loops in automatic, cropped to the plant view. (1) The status chips for the three controlled impurities, the propane recovery, the steam header and the drum. (2) The feed surge drum, drawn with the three columns on one line in the order the material flows. (3) Beside each column is a stacked bar of its overhead composition, ethane at the top and then propane, butanes and gasoline, with the contract limit drawn as a red line on the band it applies to, short-dashed. (4) The debutanizer also shows its bottoms, because the gasoline vapor pressure specification lives there and has no loop of its own. (5) The bar under the scene is the steam header, here passing 4000 of 4000 kmol/h, and every column outline has turned from its dark default to amber because the reboilers ask for more than it can pass. The boilup ratios read 1.64, 1.73 and 1.27 against setpoints of 1.80, 1.90 and 1.40. Ethane, butane and pentane in their products read 1.490, 2.288 and 1.903 mol%.

Giveaway is the room left below the ethane, pentane and gasoline butane limits, in kmol as the application reports it and in tonnes at each component's molar mass.

Over the day the MPC spends 153 against 513 min over the ethane limit, but its excess is 86.7 against 65.8 mol% min and its peak 3.28 against 2.54 mol%, as shown in Figure 10.3(a). The two structures lose in different windows. In the ethane step the MPC is over for 153 against 103 min with 86.7 against 34.1 mol% min, and in the drift window the cascade's slow trim holds the propane over the limit for 394 min while the MPC, on this seed, stays under it. Both rankings hold at half the physics step and on six other noise seeds, where the MPC spends 124 to 374 min over with 84.6 to 92.5 mol% min against 498 to 544 min with 62.1 to 66.4 mol% min.

In moles the two give away the same, 63.4 against 63.5 kmol, and by mass the MPC gives away 3.59 against 4.42 t, because the room it leaves is lighter, 23.1 against 7.7 kmol below the ethane limit in the propane and 25.2 against 43.1 kmol below the pentane limit in the butane. The molar tie lies within the noise. On the six other seeds the MPC gives away 61.4 to 65.6 kmol against 63.5 to 63.8, while its tonnage stays lower, 3.52 to 3.66 against 4.42 to 4.44 t.

Three cascade variants locate the difference. With the MPC's mean purities as targets and the matching inferential setpoint, the cascade gives away more, 73.7 kmol, so the gain is not where the targets sit. Restricted to the deethanizer cut, the MPC gives away 78.5 kmol, so the pentane gain comes from moving the three cuts together. Tuned once on this case by the rule that tuned the MPC,

Table 10.6 MPC Against the Loop Structures, Each Case at Its Own Seed, Noise and Duration

Case	Structure	Over limit (min) C2 / C4 / C5	Excess C2 (mol% min)	Peak (mol%) C2 / C4 / C5	Giveaway kmol (t)	Plans, outcome
10, 1440 min	MPC	153 / 0 / 0	86.7	3.28 / 2.50 / 1.93	63.4 (3.59)	1440
three cuts	cascade	513 / 0 / 0	65.8	2.54 / 2.34 / 1.98	63.5 (4.42)	—
planned	cascade, giveaway by setpoint[a]	373 / 0 / 0	46.3	2.51 / 2.32 / 1.90	73.7 (5.27)	—
together	cascade, hindsight[b]	111 / 0 / 0	52.1	2.77 / 2.35 / 1.98	49.3 (2.81)	—
	cascade, tuned once by the MPC's rule[c]	256 / 0 / 0	80.5	2.90 / 2.34 / 1.97	54.5 (3.40)	—
	MPC, deethanizer cut only[d]	154 / 0 / 0	89.1	3.30 / 2.34 / 1.91	78.5 (4.88)	1440
5, 1200 min[e]	cascade	107 / 206 / 324	40.3	2.59 / 5.56 / 2.60	92.9 (4.84)	—
13 mol% ethane	MPC	122 / 0 / 0	111.3	3.55 / 2.41 / 1.90	50.0 (2.75)	1200
6, 1800 min	cascade	58 / 38 / 0	8.1	2.21 / 2.57 / 1.91	93.7 (4.73)	—
drift and step	MPC	109 / 0 / 0	86.7	3.34 / 2.41 / 1.90	64.8 (3.61)	1800
7, 960 min	cascade	0 / 0 / 0	0.0	1.42 / 2.32 / 1.90	80.5 (3.57)	—
target steps	MPC	0 / 0 / 16	0.0	1.68 / 2.32 / 2.08	77.1 (3.46)	960
9, 1440 min	cascade	0 / 124 / 50	0.0	1.68 / 3.07 / 2.02	112.9 (6.12)	—
heavier feed	MPC	52 / 140 / 197	7.7	2.23 / 2.73 / 2.43	101.4 (4.47)	1440
11, 1400 min	local, loops in manual, as shipped	783 / 0 / 0	1369.5	5.22 / 2.32 / 1.90	152.3 (10.72)	fails
challenge	MPC, loops in manual	198 / 0 / 0	118.8	3.38 / 2.32 / 1.90	174.6 (11.32)	1400, fails
	cascade, suggested	76 / 231 / 143	15.4	2.31 / 4.47 / 2.10	160.1 (9.31)	passes
	MPC, otherwise suggested	219 / 108 / 136	90.3	3.19 / 2.72 / 3.56	152.0 (7.99)	1400, fails

Minutes over and peaks are read from the true products at every physics step, and the excess integrates the ethane above 2.0 mol%. No plan was rejected. [a]Targets at the MPC's mean purities, 1.764, 2.247 and 1.656 mol%, and the inferential setpoint at 1.765. [b]The fewest minutes over among 216 fixed cascade tunings, $K_c = 1$ and $\tau_I = 120$ min on the trim, 0.6 on the local loop, 1 and 2.4 on the butane and pentane loops. [c]The same grid tuned once on scenario 10 by the rule of Section 10.4.3, 0.4, 120 min, 0.6, 0.5 and 1.2. [d]A copy of the controller that plans the deethanizer cut alone. [e]Scenario 5's ethane step run on the cascade instead of its product analyzer.

the cascade gives away 54.5 kmol with 256 min over, and the best of the 216 fixed tunings, chosen in hindsight, beats the MPC on every count, with 49.3 kmol, 111 min, 52.1 mol% min and a peak of 2.77 mol%. A firmer and faster trim, a gentler local loop and faster downstream loops buy more than the plan does. Figure 10.4 sets the six structures of scenario 10 side by side on minutes over the limit, peak and giveaway.

The MPC wins where its estimator or its coordination matters. After the drift its analyzer biases recalibrate the inferential within about 100 min, a correction the slow trim spreads over most of a shift, although on two of the six other seeds the MPC is also over the limit there, for 185 and 142 min at peaks of 2.05 and 2.06 mol%, against 386 to 396 min for the cascade. After the heavier feed it moves the debutanizer cut with the other two and gives away 16.2 against 25.9 kmol. Through scenario 5's ethane step the cascade's local loop drives the deethanizer cut to 21.85 % of feed and holds butane in the propane over its limit for 206 min, where the MPC keeps it inside, as do the hindsight and the once-tuned cascades of that case, whose local loop gain is 0.6 instead of 1.20.

Every ethane surge goes deeper under the MPC, with an excess of 86.7 against 8.1 mol% min in scenario 6 and 111.3 against 40.3 through scenario 5's step. The plan holds the bottoms disturbance it sees constant while the 1500 kmol surge drum keeps delivering the richer feed for about 90 min. In the ethane step of scenario 10 its present estimate of the ethane is within 0.004 in the logarithm on average, but its predictions 30 to 90 min ahead run low by 0.11 to 0.26 on average and by up to 1.1. No change of the move price, the band price, either estimator gain, the analyzer alignment or the coupling cures the depth, because across them the peak in scenario 10 stays between 3.03 and 3.60 mol%. A deethanizer move priced at 300 instead of 20000 still peaks at 3.08 mol% and puts butane over its limit for 124 min. Told the drum's eventual disturbance and its pace, which no

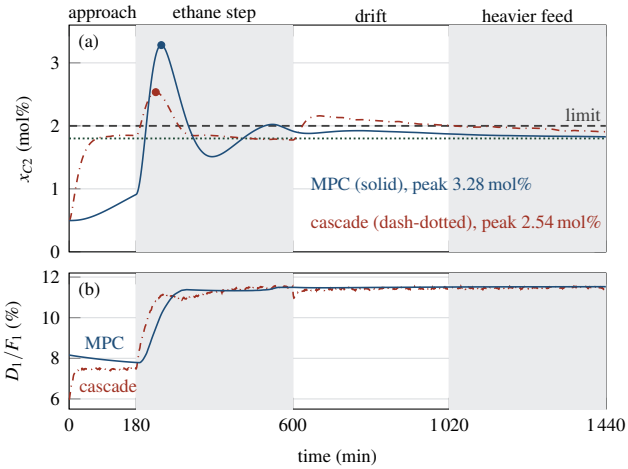


Figure 10.3 Scenario 10 under the MPC as it loads and under the cascade selected before the run, at the scenario's own noise seed, with the ethane step and the heavier feed windows shaded. The MPC is the solid line and the cascade the dash-dotted line in both panels. (a) True ethane in the propane product against the 2.0 mol% limit (densely dashed) and the 1.80 mol% target (densely dotted). The MPC peaks at 3.28 mol% at 247 min and the cascade at 2.54 mol% at 232 min. (b) The deethanizer cut ratio each structure applies.

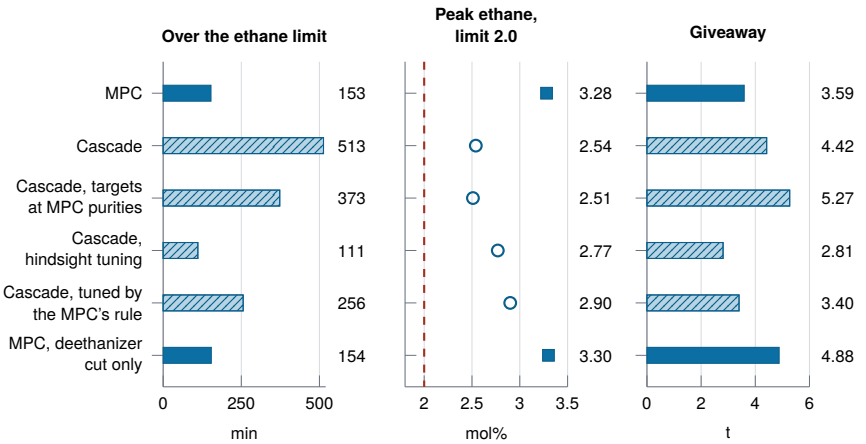


Figure 10.4 The scenario 10 rows of Table 10.6 as a chart: minutes over the 2.0 mol% ethane limit, the peak ethane in the propane against that limit (dashed line), and the giveaway in tonnes. MPC rows are solid bars and filled squares, cascade rows hatched bars and open circles. Notice that the two MPC rows spend far fewer minutes over the limit than the shipped cascade but reach the highest peaks, and that the cascade tuned in hindsight is the lowest on minutes and on giveaway.

instrument here measures, the same plan peaks at 2.24, 2.25 and 2.52 mol% in scenarios 10, 6 and 5, a bound and not a design, and with that information revealed 10 or 30 min late it peaks at 2.45 and 3.02 mol% in scenario 10.

The approach is slow as well, as shown in Figure 10.3(b). The MPC lowers the deethanizer cut from 8.16 to 7.79 % of feed over the first 180 min, and the ethane reaches 0.56, 0.72 and 0.91 mol% at 60, 120 and 180 min, where the cascade, whose local loop drops the cut to 5.96 % at once, is at 1.74, 1.84 and 1.84. The move price sets that pace. At 10000, 5000 and 2000 the plan reaches 1.24,

1.57 and 1.74 mol% by 180 min, but at 5000 and below butane goes over its limit in the surge, for 42 and 81 min.

Pentane is the other loss, over its limit for 16 min in scenario 7 where the cascade has none and for 197 against 50 min in scenario 9, with an excess of 52.7 against 0.5 mol% min, although the MPC gives away less in both. In scenario 7 the cause is the downstream move price, because priced at 10 or more instead of 1 the same plan keeps pentane inside. In scenario 9 it is the estimate, because priced at 1000 the plan still spends 190 min over, and while the pentane is over its limit the plan's estimate of it is low by 0.12 in the logarithm on average.

Judged by its own rule, the challenge is not met under the MPC. With the suggested line-up and the MPC selected, the worst errors from 924 min are 1.36 and 0.30 mol% against bands of 0.30 and 0.20, with every product inside its limit at the end, and it breaks limits that the rule leaves unscored, with 219, 108 and 136 min over the ethane, butane and pentane limits, a pentane peak of 3.56 mol% and butane in the natural gasoline over its limit for 369 min. As shipped, with the downstream loops in manual so that the plan moves the deethanizer cut alone, it fails at 1.53 and 1.77 mol% with a product outside its limit and gives away 174.6 against 152.3 kmol, the only case of Table 10.6 in which it gives away more than the structure it is set against.

10.6 Technology in Practice

Figure 10.5 draws the loops of Section 10.4 with the hardware of a fractionation train in place of the schematic's symbols.

What Measures It

Four products leave under five limits, and three of the limits have an instrument a loop can use. Ethane and butane in the propane are read by the chromatographs AI-C2 and AI-C4 on the depropanizer overhead and pentane in the butane by AI-C5 on the debutanizer overhead, product analyzers of the kind Chapter 9 describes, each on a 10 min cycle. Butane in the natural gasoline and propane in the ethane have limits and no loop. The one fast composition measurement is the deethanizer's local inferential, TI-101, a lagged estimate of the ethane left in its bottoms.

> **TECHNOLOGY Sensor: The Tray Temperature Behind the Inferential**
>
> Product compositions in a distillation column are rarely measured directly for control, because analyzers such as gas chromatographs bring long delays and high investment and maintenance costs. Tray temperature control has widely taken their place, on the assumption that the product meets its specification while a well-chosen tray holds its temperature. The assumption weakens when the feed composition or rate changes in a multicomponent column, and pressure changes move the temperature too. Remedies include placing the measurement where the nonkey components hardly change and using a temperature difference to compensate for pressure [19]. TI-101 of this chapter is such an inference, with its drift written as an explicit bias.

What Moves It

Seven final elements carry the chapter's handles. Three distillate valves set the cuts, each under a ratio station that multiplies its column's feed, so the quality controllers move ratios rather than flows. Three steam valves set the boilups, which stay economic setpoints of the operator, and the train feed valve holds the surge drum level under the slow averaging loop LC-101. When the header passes less than the reboilers ask, every boilup falls by the same factor, as Equation (10.3) writes.

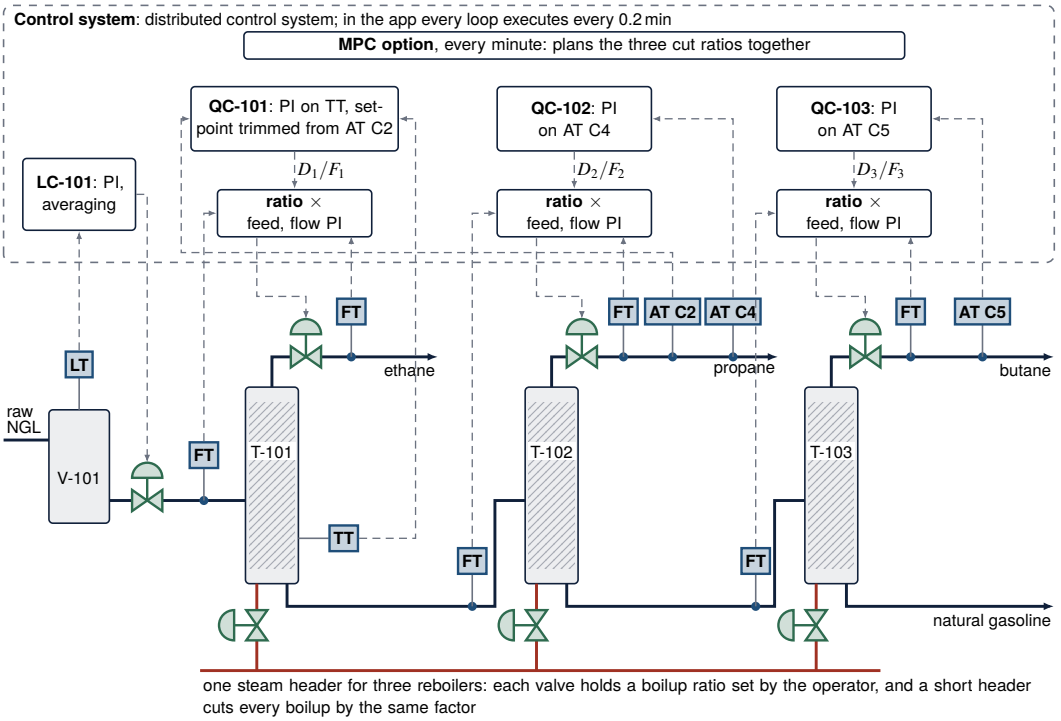


Figure 10.5 The loops of the NGL train with their hardware. Lettered boxes are sensing devices: LT on the surge drum, TT for the deethanizer inferential, AT C2 and AT C4 on the propane, AT C5 on the butane, and FT for the flow meters that the ratio stations need. Green symbols are final elements: the train feed valve, the three distillate valves and the three reboiler steam valves. Dashed arrows are signals. Notice the long arrow from AT C2 on the depropanizer overhead back to QC-101 on the deethanizer, and that every cut reaches its valve through a ratio station that reads the column's feed. Condensers, reflux drums and reboilers are left out, and a steam valve stands for each reboiler. Not to scale, and not a piping diagram.

> **TECHNOLOGY** **Final Element: The Distillate Valve Under a Ratio Station**
>
> A plant builds a cut ratio from a flow loop. The flow transmitter on the column's feed, the wild flow, is multiplied by the ratio in a ratio station, and the product becomes the setpoint of a flow controller that moves the distillate valve, the captive flow. The ratio holds only while the two transmitters' calibrations agree. Pressure-based flowmeters read well over a turndown often no more than three to one, perhaps ten to one with better transmitters, and a dulled orifice edge or uneven liquid in the impulse lines gives large errors at low flow [20]. The chapter's stations multiply exactly, so at a fixed feed composition a rate change is removed before any controller sees it.

Where the Controller Runs

A train like this runs on a distributed control system, which controls production within one location, and its supervisory and lower loops run with cycle times from milliseconds to minutes [21]. The control structure itself is layered by time scale: scheduling over weeks, site-wide optimization over a day, local optimization over an hour, supervisory and predictive control over minutes and regulatory control over seconds [2]. The app's loops execute every 0.2 min and its MPC every minute, while the chromatographs cycle every 10 min. The chapter's structure maps onto standard blocks: PI blocks for the quality and level loops, three ratio stations, a cascade in which the analyzer trim writes the

remote setpoint of the inferential loop and, in a real plant, a load shedding scheme or a multivariable controller where the app has its proportional header rule.

> ⚠ **IN PRACTICE** **Commission the Inner Loop First**
>
> A cascade is commissioned from the inside out. The secondary loop is tuned first, because it does not depend on the primary, while the primary can do nothing at all if the secondary is in manual. The secondary has three modes: manual, automatic on a local setpoint, and cascade on the remote setpoint the primary writes, and leaving it in automatic quietly defeats the cascade. The secondary must also respond faster than the primary [20]. Here the inner loop is the deethanizer's inferential loop with an integral time of 28 min, and the analyzer trim above it has 240 min.

What Is Modernizing It

Two developments touch this chapter. The first is the inference itself. Soft sensors are used to stabilize quality, to run closer to specifications and to validate online analyzers by comparison, and recursive models and model bias updates keep them current as a process changes [19]. The analyzer bias that the MPC of Section 10.4.3 updates at every result is of this kind. The second is predictive control, which sits in the supervisory layer above the regulatory loops [2]. The chapter's comparison shows what it still lacks on this train: a model of the surge drum's mixing, or a feed analyzer, so that it does not hold a disturbance constant while the drum is still delivering it.

10.7 Limitations

The default debutanizer cut of 49.80 % starts both products inside specification: pentane in butane is 1.899 mol% and butane in gasoline is 1.559 mol%.

The columns are shortcut splits, so nothing hydraulic can be studied. Flooding, weeping and efficiency loss at turndown are all absent, and a real train is often limited by one of them rather than by steam.

The composition dynamics are first-order lags fitted to the dominant time constant. The shape of the approach on a real column is not first order, and a student fitting a model from this application would find it easier than the same exercise on a rigorous simulation.

The four pseudo-components remove the isobutane and normal butane separation that a real debutanizer manages, and the natural gasoline vapor pressure specification is a mole fraction limit rather than a computed vapor pressure.

The steam header allocation has no priority scheme. That is the point of scenario 8, but it also means the application cannot demonstrate a well designed load shedding scheme working correctly, only the consequence of not having one.

The optional MPC has limits of its own. Its model is linear in the logarithms at one operating point, and at the design point the deethanizer cut moves the ethane 1.41 times as hard as it believes. It holds the disturbance at the deethanizer bottoms constant over the plan, with no model of the surge drum's mixing and no feed analyzer, which is why its surges go deeper. After large deethanizer moves its estimates of the downstream products run high, by up to 0.24 in the logarithm for butane and 0.37 for pentane in the ethane step of scenario 10, until analyzer results correct them, and its biases correct an offset, not a gain. It leaves the boilups, the steam header and the natural gasoline to the operator and the plant, so the arbitration missing in scenario 8 is still missing under it. Its limits are soft, with one slack per impurity for the whole horizon, and its prices were chosen once, on scenario 10.

Finally, there is no recycle anywhere in the train. One pass down it is the steady state, which is what makes the plantwide reasoning tractable and also what removes the snowball effect that plantwide control literature spends most of its attention on [1]. Volume 6, Chapter 7 studies that effect on a reactor and column joined by a recycle.

OPEN CHALLENGE Keeping a Soft Sensor Honest

Engineers surveyed in industry named model maintenance the most critical issue for soft sensors, and the problem cited most often was accuracy lost as process characteristics change. An estimate deteriorates as equipment changes and as feed properties vary, and rebuilding the models by hand each time is too heavy a workload. Recursive methods and bias updates follow slow change but cannot cope with abrupt change [19]. Progress would be the high-performance, maintenance-free virtual sensing that the review calls for, an inference whose drift shows before the giveaway does.

10.8 Discussion

Three results from this train generalize.

The first is that a nonzero process gain does not establish a pairing. The depropanizer cut moves ethane in the propane product by 2.7 mol% across the swept range, which is a large gain and a useless one, because the mechanism is dilution and the mechanism runs out. Before pairing a loop it is worth asking not whether the manipulated variable moves the controlled variable but by what physical route, and whether that route has an end.

The second is that a soft sensor fails in the direction that does not alarm. The drifted inferential in Table 10.4 settles the plant at 0.705 mol% against a 1.80 mol% target, which is inside every contract and costs money continuously. A failure that produces giveaway rather than an off-specification product is a failure a plant can run with for a long time, and the only defenses are a periodic comparison against a reference measurement and a supervisory layer above it.

The third is that a shared constraint resolved by arithmetic will be resolved against the specification that is least defended. The header cut all three reboilers by 8.8 % and each controller defended what it was given, and the product that left its contract and stayed out was the one with no controller on it. That is not a failure of any loop. It is a statement that the decision of which product to sacrifice was never made, and a control system that does not make it will make it implicitly.

10.9 Conclusion

A fractionation train is where loop pairing stops being a two by two exercise. The simulated train of this chapter measures ethane in the propane product on the depropanizer overhead and cannot change it there: the ethane flow in that distillate is 48.5 kmol/h across the entire cut range and the percentage falls only by dilution, which would require a distillate of 2430 kmol/h against a column feed of 918 to reach the contract. The valve is the deethanizer cut, one column upstream. Measured on the same disturbance, a deethanizer inferential drifted 30 percent low settles the propane at 0.705 mol% against a 1.80 mol% target while, at 30 h, the cascade above it reaches 1.751 and the analyzer alone 1.800, so the soft sensor failure costs giveaway rather than an alarm. A steam header trip from 5200 kmol/h to 4000 kmol/h multiplies every boilup by 0.912, leaves the three controlled impurities at 1.49, 2.29 and 1.90 mol% inside their limits at 15 h after excursions of up to 194 min, and takes butane in the natural gasoline to 3.81 mol% against a limit of 2.0, because that is the specification with no loop on it. The predictive controller offered as an option, planning the three cuts together, spends 153 against 513 minutes over the ethane limit in scenario 10 but peaks at 3.28 against 2.54 mol%, because it holds constant a disturbance that the surge drum is still delivering, and a cascade retuned in hindsight beats it on minutes, depth and giveaway. Further work would add a recycle so that the snowball effect becomes reachable, a load shedding scheme with an explicit priority, and a rigorous stage model on one column so that the shortcut split can be checked against it. On the train as it stands, a reader can find by experiment where each specification is made.

KEY POINTS

- The ethane in the propane is measured on the second column but made on the first, so the valve that sets it is one column upstream.
- A nonzero process gain does not establish a pairing: the depropanizer cut moves the ethane only by dilution, and the mechanism runs out.
- A soft sensor fails in the direction that does not alarm, and the drifted inferential settles the propane at 0.705 mol% against a 1.80 mol% target.
- A shared constraint resolved by arithmetic is resolved against the specification that is least defended, here butane in the natural gasoline.

Where else this applies. Volume 5, Chapter 3 closes a composition loop through a slow analyzer on one column, and Volume 5, Chapter 10 corrects a soft sensor's bias with a late measurement.

References

[1] M. L. Luyben, B. D. Tyreus, and W. L. Luyben, "Plantwide control design procedure," *AIChE Journal*, vol. 43, no. 12, pp. 3161–3174, 1997.

[2] S. Skogestad, "Plantwide control: the search for the self-optimizing control structure," *Journal of Process Control*, vol. 10, no. 5, pp. 487–507, 2000.

[3] ——, "Economic plantwide control," in *Plantwide Control*. John Wiley & Sons, 2012, pp. 229–251.

[4] W. L. Luyben, "Heuristics for plantwide control," in *Plantwide Control*. John Wiley & Sons, 2012, pp. 97–119.

[5] S. J. Qin and T. A. Badgwell, "A survey of industrial model predictive control technology," *Control Engineering Practice*, vol. 11, no. 7, pp. 733–764, 2003.

[6] T. Larsson and S. Skogestad, "Plantwide control – a review and a new design procedure," *Modeling, Identification and Control*, vol. 21, no. 4, pp. 209–240, 2000.

[7] B. F. Greek, "Natural gas liquids remain strong petrochemical feedstock," *Chemical & Engineering News Archive*, vol. 62, no. 11, pp. 17–38, 1984.

[8] L. Li, "Development of natural gas industry in China: Review and prospect," *Natural Gas Industry B*, vol. 9, no. 2, pp. 187–196, 2022.

[9] M. R. Fenske, "Fractionation of straight-run Pennsylvania gasoline," *Industrial & Engineering Chemistry*, vol. 24, no. 5, pp. 482–485, 1932.

[10] A. J. V. Underwood, "Fractional distillation of multicomponent mixtures," *Industrial & Engineering Chemistry*, vol. 41, no. 12, pp. 2844–2847, 1949.

[11] E. R. Gilliland, "Multicomponent rectification: Estimation of the number of theoretical plates as a function of the reflux ratio," *Industrial & Engineering Chemistry*, vol. 32, no. 9, pp. 1220–1223, 1940.

[12] F. Duanmu, D. N. Chia, and E. Sorensen, "A shortcut design method for complex distillation structures," *Chemical Engineering Research and Design*, vol. 180, pp. 346–368, 2022.

[13] P. Kadlec, B. Gabrys, and S. Strandt, "Data-driven soft sensors in the process industry," *Computers & Chemical Engineering*, vol. 33, no. 4, pp. 795–814, 2009.

[14] U.S. Energy Information Administration, "U.s. natural gas plant field production, annual, thousand barrels per day," Petroleum and Other Liquids data series, release of October 30, 2026, 2026, accessed November 18, 2026. [Online]. Available: https://www.eia.gov/dnav/pet/pet_pnp_gp_dc_nus_mbblpd_a.htm

[15] K. J. Åström and T. Hägglund, *Advanced PID Control*. Research Triangle Park, NC: ISA – The Instrumentation, Systems, and Automation Society, 2006.

[16] C. A. Larsson, C. R. Rojas, X. Bombois, and H. Hjalmarsson, "Experimental evaluation of model predictive control with excitation (MPC-X) on an industrial depropanizer," *Journal of Process Control*, vol. 31, pp. 1–16, 2015.

[17] W. L. Luyben, "Control of a train of distillation columns for the separation of natural gas liquid," *Industrial & Engineering Chemistry Research*, vol. 52, no. 31, pp. 10 741–10 753, 2013.

[18] G. Pannocchia and J. B. Rawlings, "Disturbance models for offset-free model-predictive control," *AIChE Journal*, vol. 49, no. 2, pp. 426–437, 2003.

[19] M. Kano and K. Fujiwara, "Virtual sensing technology in process industries: Trends and challenges revealed by recent industrial applications," *Journal of Chemical Engineering of Japan*, vol. 46, no. 1, pp. 1–17, 2013.

[20] T. R. Kuphaldt, *Lessons in Industrial Instrumentation*, version 2.33 ed. Open textbook, Creative Commons Attribution 4.0 license, ibiblio.org/kuphaldt/socratic/sinst, 2024.

[21] K. Stouffer, M. Pease, C. Tang, T. Zimmerman, V. Pillitteri, S. Lightman, A. Hahn, S. Saravia, A. Sherule, and M. Thompson, "Guide to operational technology (OT) security," National Institute of Standards and Technology, Gaithersburg, MD, Special Publication 800-82r3, 2023.

10.10 Exercises

10.1 Concept check · 2 min

In scenario 2 the raw NGL carries 13 mol% ethane. Taking the depropanizer cut ratio from 31.6 to 37.6 % of its feed lowers the ethane in the propane product from 16.7 to 14.0 mol%. Why is the depropanizer cut not a usable handle for the ethane specification?

(a) The gain is small, and a cut range a little wider than the slider's would reach the limit

(b) Ethane is lighter than the column's light key, so all of it goes overhead and the cut only dilutes it

(c) The analyzer cycle of 10 min is too slow for a loop closed on the depropanizer cut

(d) The cut also drags butane overhead, so an ethane loop would fight the butane loop

10.2 Concept check · 1 min

The deethanizer cut is held fixed and its boilup ratio is raised from 1.6 to 2.2. What happens to the ethane left in its bottoms and the propane lost overhead?

(a) The ethane falls and the propane rises, along the curve a larger cut also moves along

(b) Both rise, because more vapor carries more of each key component up the column

(c) Both fall, because a larger separation factor sharpens the split at the same cut

(d) Neither moves, because the cut alone fixes how the material balance is placed

10.3 Concept check · 2 min

In scenario 6 the local loop alone holds the deethanizer inferential at its 0.50 mol% setpoint when the inferential drifts to read 30 % low. The product target is 1.80 mol%. Where does the ethane in the propane settle, and what does the control room see?

(a) Above the 2.0 mol% limit, and the propane analyzer alarm sounds within the hour

(b) On the 1.80 mol% target, because the loop's integral action removes the bias

(c) Near 0.35 mol%, because a low reading makes the loop cut harder than needed

(d) Near 0.7 mol%, above what the setpoint intends but below target, with no alarm

10.4 Theory and calculation · 5 min

In scenario 2 the deethanizer bottoms carry 48.51 kmol/h of ethane into the depropanizer, whose feed is 918.4 kmol/h. The cut ratio is the distillate as a percentage of that feed.

(a) Find the distillate and the ethane in the propane at cut ratios of 31.6 and 37.6 %. *Checkpoint:* 290.2 and 345.3 kmol/h, 16.72 and 14.05 mol%.

(b) Find the distillate that would bring the ethane to 2.0 mol% by dilution alone, as a multiple of the column feed. *Checkpoint:* About 2425 kmol/h, 2.6 times the feed.

(c) Find the slope of the ethane percentage against the distillate at each cut, and explain why a nonzero gain is not enough to pair the loop. *Checkpoint:* -0.058 and -0.041 mol% per kmol/h.

10.5 Theory and calculation · 7 min

The deethanizer has $\alpha_1 = 2.10$, $N = 25$, $k_1 = 0.4703$ and $(V/F)_0 = 0.50$, and at design it is fed 80 kmol/h of ethane and 320 kmol/h of propane.

(a) Use Equation (10.2) to find η and S at boilup ratios 1.6, 1.8, 2.0 and 2.2, and the factor by which each step of 0.2 multiplies S. *Checkpoint:* $S \approx$ 1790, 4840, 11900 and 27200, factors 2.70, 2.47 and 2.28.

(b) At a boilup ratio of 1.8 the chapter reports 1.60 kmol/h of ethane left in the bottoms and 3.20 kmol/h of propane overhead, and 0.62 and 8.22 kmol/h with six more kmol/h of cut. Find the key recoveries and S from Equation (10.1) in each case. *Checkpoint:* About 4850 and 4860.

(c) At a boilup ratio of 2.2 and the design cut the chapter reports 0.45 and 2.05 kmol/h. Find S the same way and compare it with (a). *Checkpoint:* About 27400.

10.6 Theory and calculation · 7 min

In scenario 8 the three reboilers ask for 4387 kmol/h of boilup when the header capacity drops to 4000 kmol/h. The boilup setpoints are 1.80, 1.90 and 1.40.

(a) Find ϕ of Equation (10.3) and the boilup ratios the three columns receive. *Checkpoint:* $\phi = 0.912$, ratios 1.64, 1.73 and 1.28.

(b) The debutanizer has $\alpha_3 = 2.60$, $N = 25$, $k_3 = 0.4587$ and $(V/F)_0 = 0.50$. Find its S before the trip and after it. *Checkpoint:* About 3230 and 1290.

(c) The pentane loop holds pentane in the butane on target with the debutanizer cut. Use Equation (10.1) to say which way the cut must move once S has fallen, and what that does to the butane in the natural gasoline.

10.7 App challenge · 8 min

Run scenario 2, *The unit that measures it cannot fix it*, at the default seed. It loads in Manual with the raw NGL at 13 mol% ethane and the Depropanizer cut ratio at 35.14 % of feed.

(a) Set the Depropanizer cut ratio to 31.6 % before Run and run 90 min at speed 20 (20 min per second, about 5 s). Read the C2 in C3 and C4 in C3 badges, and in the exported CSV the columns `d2_kmolh` and `b1_kmolh` at 90 min. Compute the ethane flow in the distillate. *Checkpoint:* 16.72 % and 0.08 %, 290.2 and 918.4 kmol/h, 48.5 kmol/h of ethane.

(b) Set the cut ratio to 37.6 % and run another 90 min. Read the same four values and compute the ethane flow again. *Checkpoint:* 14.16 % (within 0.02) and 0.27 %, about 342.6 and 918.4 kmol/h, 48.5 kmol/h of ethane.

(c) From your readings, find the distillate that would put the ethane on 2.0 mol%, and say which valve sets this specification.

10.8 App challenge · 8 min

Run scenario 6, *The cascade, and an inferential that has drifted*, at the default seed. At 30 min the deethanizer inferential drifts to 0.70 of its reading and at 31 min the raw NGL steps to 12 mol% ethane. Run each case 1800 min at speed 120 (120 min per second, 15 s) and export the CSV.

(a) Run as shipped, on Cascade. Read `c2_in_c3_pct` at 900 and 1800 min, `c2_local_sp_pct` at 1800 min, and the Giveaway and Off-spec propane tiles at the end. *Checkpoint:* 1.595 and 1.751 %, 1.244 %, 93.7 kmol and 96 min.

(b) Reload, select Local in Ethane-in-propane control before Run, and read the same values. *Checkpoint:* 0.705 and 0.705 %, 0.50 %, 168.9 kmol and 39 min.

(c) Reload, select Product analyzer before Run, and read the same values. Then rank the three structures for a plant that sells propane against the 2.0 mol% limit. *Checkpoint:* 1.856 and 1.800 %, 0.50 % (unused), 62.5 kmol and 622 min.

10.9 App challenge · 8 min

Run scenario 8, *One header, three reboilers*, at the default seed. It loads on Cascade with both downstream loops in Auto, and at 60 min the header capacity falls from 5200 to 4000 kmol/h. Run 900 min at speed 60 (60 min per second, 15 s) and export the CSV.

(a) Read `steam_cut` in the first rows after 60 min, and `vf1`, `vf2` and `vf3` at 900 min. *Checkpoint:* 0.912, then 1.639, 1.730 and 1.275 (within 0.005).

(b) At 900 min, read the C2 in C3, C4 in C3, C5 in C4 and C3 recovery badges and the gasoline C4 on the debutanizer card. *Checkpoint:* 1.49, 2.29 and 1.90 %, 97.1 %, 3.81 %.

(c) Read the three Off-spec tiles at 900 min, and the time at which `c4_in_c5_pct` first exceeds 2.0. Say which product the trip sacrificed and why. *Checkpoint:* Propane 64 (within 1), butane 194 and gasoline 795 min, and a crossing at 105 min.

10.10 Application area · 3 min

After the boiler trip of scenario 8 the header shares 4000 kmol/h of boilup in proportion to what each reboiler asks for.

(a) Name the two kinds of layer the chapter says a real train uses instead, and what such a layer must know that the proportional rule does not.

(b) At 15 h ethane in the propane reads 1.49 % against 2.0, butane in the propane 2.29 against 2.5 and pentane in the butane 1.90 against 2.0. Which column would you ask to give up steam first to protect the gasoline, and what else would you check?

(c) Which simplifications of the model would make you distrust the 3.81 mol% figure on a real plant?

10.11 Application area · 4 min

A gas plant runs its deethanizer on a local loop that holds a pressure-compensated tray temperature inferential.

(a) Why does a 30 % low drift of that inferential raise no alarm in scenario 6?

(b) Name the two defenses the chapter gives, and say how the trim output of a cascade can itself serve as a drift monitor.

(c) In scenario 10 the predictive controller meets a 25 % drift. How quickly do its analyzer biases take it up, and how quickly does the cascade's trim?

10.12 Review the agent · 5 min

An AI assistant summarizes scenario 10: "Over the day the MPC spends 153 min over the ethane limit against the cascade's 513, and the Giveaway tiles read 63.4 against 63.5 kmol. The MPC is better on ethane at no cost in giveaway, so it should replace the cascade." Find the flaw, and say how you would check the claim in the app and in the chapter.