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CHAPTER 9

Ethane Recovery and Rejection Through a Sequenced Turboexpander Plant

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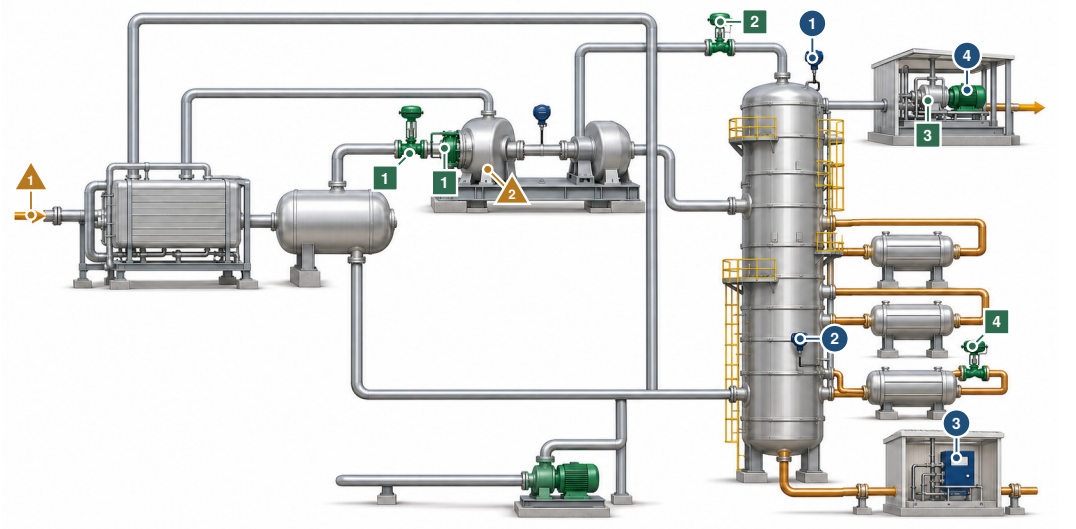
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CHAPTER 9

Ethane Recovery and Rejection Through a Sequenced Turboexpander Plant



The controller moves 1 expander guide vanes, with the valve beside them that takes the rest 2 reflux split valve 3 residue compressor, which holds the column pressure 4 trim heat to the bottom reboiler

The controller reads 1 column pressure 2 control tray temperature 3 methane in the liquid product, reported minutes late 4 compressor driver power and its limit

What disturbs it 1 a richer or larger inlet gas 2 an expander trip ▲ a fall in the ethane price

Notice the one shaft that joins the expander to the booster, and the analyzer shelter at the far end of the liquid product line, the last place the methane shows. Illustration, not to scale.

IN PLAIN TERMS Pulling Liquids Out of Natural Gas

The main idea. A gas plant chills natural gas until ethane and its heavier parts turn to a liquid it can sell. A spinning machine called an expander makes the deepest cold, and a tall column boils the last methane out of the liquid.

The challenge. When ethane is worth more in the sales gas, the plant warms the column and raises its pressure to let the ethane go. If the pressure rises first, methane spoils the liquid, and the instrument that checks it reports minutes late.

Why it matters. The same two moves can keep or ruin a day's product, depending only on their order.

THREE WAYS TO READ THIS CHAPTER

Explore	12 min	This page, the History boxes in Section 9.1 and Section 9.2, the Did you know? boxes in Sections 9.3 and 9.5, the Try it boxes in Section 9.5, and Key points at the end of the Conclusion. No equations.
Study	40 min	The whole chapter in order, stopping at the three pause points, then Exercises 1 to 12.
Practice	15 min	This page, Section 9.5 with Figure 9.2, Section 9.6 and Section 9.7 with its Open challenge box.

The control question When ethane is worth less as a liquid than as gas, a gas plant moves its demethanizer from ethane recovery to rejection and back, while a chromatograph reports the methane in its liquid product minutes late. In what order must the column pressure and the tray temperature move, and what does the move cost?

Learning objectives

- Explain how one expander shaft and the feed-heated reboilers tie refrigeration, recompression and column temperature together, and compare expander and valve cooling.
- Compute the frac spread and the margin of a mode from prices, and read the pressure map as recovery traded against recompression power.
- Explain why a tray temperature held on its setpoint stops meaning a methane specification, and how a delayed chromatograph trims it.
- Predict which setpoint must move first in a mode switch, and show from runs that the order and the return margin are both needed.
- Judge a controller by every limit of a run, not by its setpoint tracking or its margin alone.

Prerequisites Sections 1.4.7 (expansion cooling), 1.4.9 (late and inferred measurements), 1.6.2 (the framework's PI) and 1.6.4 (the apps). Section 1.6.3 is needed only for the predictive comparison. Chapter 8 introduces the analyzer trim.

The app https://apmonitor.com/apps/turboexpander-plant/

See also Volume 8, Chapter 13 for a supervisor that holds a mode as a latched state, Volume 8, Chapter 9 for late results matched to the product they describe, and Volume 8, Chapter 3 for a composition loop that holds its setpoint while the product misses. Volume 9, Chapter 8 for setpoints chosen from a cost model, and Volume 9, Chapter 10 for a slow quality trim on a setpoint loop, priced in energy.

9.1 Introduction

A single steady recovery number hides the hardest moment in a natural-gas-liquids (NGL) plant. When ethane's value as liquid falls below its value in the residue gas, the desired column pressure and bottoms temperature change together. The NGL must still meet its methane limit while cold metal, column inventory and a shaft continue moving. A static process diagram cannot show whether a pressure move arrives before the reboiler's effect or whether a chromatograph reports an earlier inventory. Dynamic studies of NGL recovery and demethanizer control establish why the process has coupled, delayed responses [1, 2].

The plant of this chapter joins the economic choice to its physical consequence. Its cold box preheats the column's reboilers, so an action that changes the column returns to the feed temperature. The expander's refrigeration simultaneously drives a residue booster. Column pressure trades ethane recovery against recompression power, and a warmer control tray strips methane but also rejects ethane. Pressure, tray temperature and reflux therefore cannot be assigned independent economic values during a transition. Published optimization studies have shown the pressure and heat-integration tradeoffs in turboexpander processes [3–5]. The simulator of this chapter adds a reproducible dynamic experiment in which the order of two ordinary setpoint moves is a measurable constraint, with the economics and the source of each measurement visible. Section 1.4.7 gives the expansion cooling on which the plant rests, and Table 1.2 places its limits beside those of the other plants of the volume.

This chapter explains how the synthetic turboexpander plant's sampled NGL specification depends on the mode-switch sequence. Section 9.2 places the process in its technical history. Sections 9.3 and 9.4 give the model and the controls: the thermodynamics, stages, exchangers and economics, then the two regulatory loops, the analyzer trim, the mode sequencer, the economic supervisor and the predictive comparison. Section 9.5 reports the nine measured cases, ending with the twelve-hour shift. Section 9.6 sets the hardware of a real gas plant beside the model, Section 9.7 states what the model cannot represent, and Section 9.8 summarizes the measured lessons.

> ⧖ HISTORY **1964: Expansion That Pays for Its Own Recompression**
>
> On 28 April 1964 an engineer in Houston applied for a patent on treating high-pressure natural gas that had to leave the plant at a lower pipeline pressure. Feed gas already cooled well below the freezing point of water was expanded through a turbine of a kind that tolerates the liquid it condenses, down to a stripping pressure below the line pressure, chosen for the best liquid recovery. The turbine's own work then drove the compressor that restored the gas to line pressure, and that compression load steadied the turbine. On its way to the compressor, the cold gas from the stripper precooled the incoming feed [6]. The expander, the booster on its shaft and the feed exchanger of this chapter's plant are those three pieces.

9.2 Process Background and Prior Work

Early cryogenic NGL recovery relied on cooling and expansion to condense heavy components, followed by a demethanizer that rejected methane overhead. The gas-subcooled process (GSP) added a subcooled reflux branch to increase ethane capture relative to a single expander arrangement, and the History box below tells of the patent that split the cold gas this way. Honeywell UOP describes GSP as a high-recovery standard solution and labels its ethane recovery band 80–92%, its recovery-mode propane band 92–100%, and its ethane-rejection option as partial [7]. These are licensor categories, not measurements on the present model. The brochure defines partial ethane rejection by the ethane share of the liquid product, so its category is not itself a target of 8.1% recovery.

> ⧖ HISTORY **1977: Splitting the Cold Gas for a Reflux**
>
> In a patent application of 14 July 1977, two engineers named the limit of the expander plants of their day: the demethanizer worked largely as a stripping column, so the ethane lost overhead depended on the column's top temperature at a given pressure. Their answer split the cold gas before expansion. One part was cooled until it had substantially condensed, then expanded and fed above the rest, while the rest went through the expander to a feed point lower in the column. The split also diluted the carbon dioxide in the rising vapor and lowered its risk of freezing, so a colder top, and more ethane, became practical. The patent was granted on 14 July 1981 [8]. The reflux split and subcooler of this chapter's plant follow the same arrangement.

The separation and its power draw remained objects of process optimization. Chebbi and coauthors varied demethanizer pressure and feed in a conventional turboexpander study in 2010 [3]. Later comparisons examined reflux and alternative flowsheets [4, 5], while exergy studies allocated loss among expansion, heat exchange and recompression [9, 10]. This work motivates showing the energy cost alongside recovery instead of asking for maximum recovery alone. The 2013 demethanizer control study and later dynamic NGL study addressed interacting column and plant controls [1, 2]. A 2022 separation-train study compared a cascade with boilup approximation and pressure compensation with conventional temperature control under feed disturbances [11]. Their models and aims differ from this teaching plant, but together they establish that an inferential tray target and a measured product composition need not respond on the same clock.

The simulator's four pure-component reference points use the methane, ethane, propane and normal-butane boiling and critical data compiled in the NIST Chemistry WebBook [12]. The model deliberately does not use a production equation of state or a calibrated binary-interaction matrix. Matching a licensor recovery band is only a scale check for its selected feed and equipment settings, and Section 9.3.2 reports the sensitivity to exchanger-cell count.

9.3 Plant Model

9.3.1 Flow Arrangement, States and Measurements

The feed is 100 MMscfd, represented as 4980 kmol/h at 80 bar absolute and 40 °C. Its mole fractions are $(0.900, 0.055, 0.025, 0.020)$ for methane, ethane, propane and a normal-butane surrogate for heavier hydrocarbons. A fraction $s = 0.27$ of the feed passes E-2, where it gives heat to two side reboilers and the feed-heated bottom reboiler. The rest passes the counter-current E-1 exchanger

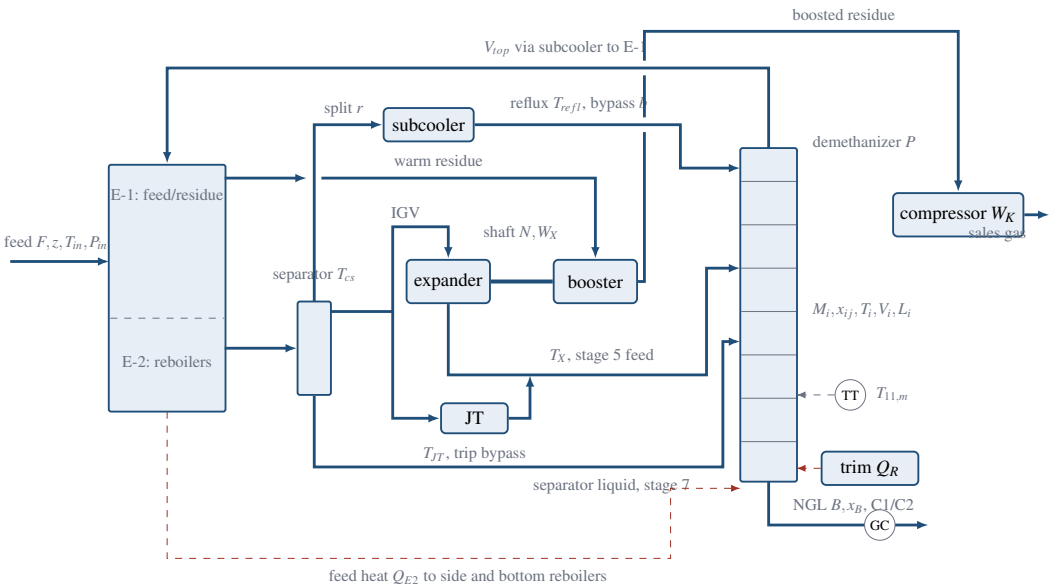


Figure 9.1 Teaching model of the gas-subcooled recovery plant. The expander and residue booster share a shaft. Solid arrows are material paths. Dashed links denote heat or work coupling and chromatograph information. E-2 returns feed heat to the side and bottom reboilers. The column has twelve equilibrium stages. The marked feeds and measured stage 11 are the key control locations. This is a model diagram, not an equipment piping specification.

against cold column overhead. The portions reunite in a cold separator. Separator vapor divides into subcooled reflux and a path through the expander or its parallel Joule–Thomson (JT) valve. Separator liquid enters the lower column. Column overhead returns through the subcooler and E-1 before a shaft-driven booster and a residue compressor deliver gas to the sales line. NGL leaves the sump, and Chapter 10 splits such a liquid into products. Figure 9.1 follows the application's Plant view.

The integrated state has 127 quantities: each of twelve stages stores total liquid and three independent mole fractions, the sump stores four quantities, E-1 has 48 metal temperatures, E-2 has three, and the reflux subcooler has 16. Shaft speed, column pressure, delivered trim duty and a flooding lag add four states, with four integrated component ledgers completing the vector. The engine of Section 1.6.1 integrates it with a two-second fourth-order Runge–Kutta physics step, a ten-second control and history interval, and ten-minute chromatograph cycles. It has no gas-side dynamic holdup beyond the one column-pressure inventory. The screens distinguish measured values from optional simulation truth.

9.3.2 Declared Thermodynamics and Stage Balances

For component j, a two-constant vapor-pressure line is fixed by its normal boiling point $T_{b,j}$ and critical point $(T_{c,j}, P_{c,j})$. Ideal-solution K values approach unity as pressure approaches the declared convergence pressure P_k:

$$B_j = \frac{\ln(P_{c,j}/P_b)}{1/T_{b,j} - 1/T_{c,j}}, \quad A_j = \ln P_{c,j} + \frac{B_j}{T_{c,j}},$$
$$\ln P_j^{sat} = A_j - \frac{B_j}{T}, \qquad \ln K_j = \left(1 - \frac{P}{P_k}\right) \ln \frac{P_j^{sat}}{P}. \tag{9.1}$$

Here $P_b = 1.013\,25\,\text{bar}$ and $P_k = 140\,\text{bar}$. A stage bubble point solves $\sum_j x_j K_j = 1$ in ten Newton iterations. The methane line above its critical temperature is a hypothetical volatility, not a physical

saturation curve. The four reference-point pairs reproduce P_b at each T_b and the corresponding P_c at each T_c to machine precision. This does not validate a mixture near its critical region.

The gas stream uses a common $c_{p0} = 0.036\,\mathrm{MJ/(kmol\,K)}$ and a declared nonideal volume $v = RT/P - A/T^2$. Its consistent enthalpy and entropy are

$$\begin{aligned} h_V &= c_{p0}(T - T_0) - 3A(P - P_0)/T^2, \\ s_V &= c_{p0}\ln T - R\ln P - 2A(P - P_0)/T^3. \end{aligned} \tag{9.2}$$

The enthalpy reference is $T_0 = 273.15\,\mathrm{K}$ at $P_0 = 1\,\mathrm{bar}$ in Eq. (9.2), distinct from the normal-boiling reference pressure in Eq. (9.1). The constant is $A = 420$ in the application's MJ, kmol, K and bar units. The liquid enthalpy subtracts $\sum_j x_j \lambda_j(T)$ from vapor enthalpy, and the two-phase enthalpy subtracts $(1 - \psi)$ times that sum. The latent heat at each normal boiling point is RB_j and follows a Watson exponent 0.38, with reduced temperature capped at 0.97. An isothermal Rachford–Rice flash supplies vapor fraction ψ, liquid and vapor compositions. These closures are small enough to inspect but do not include heats of mixing or composition-dependent gas heat capacities.

The expander's isentropic reference outlet satisfies $s_V(T_{2s}, P_2) = s_V(T_1, P_1)$. With efficiency $\eta_X = 0.87$, the actual specific work is $\eta_X[h_V(T_1, P_1) - h_V(T_{2s}, P_2)]$. Equation (1.13) is its ideal-gas counterpart. The JT outlet instead solves $h(T_{JT}, P_2) = h_1$. For the same separated vapor and 23 bar absolute destination, the model gives $-70.9\,^\circ\mathrm{C}$ after expansion and $-49.2\,^\circ\mathrm{C}$ after throttling. The measured flash enthalpy difference on the JT path is below 10^{-6} MJ/kmol. Shaft speed N changes as expander power less booster work and bearing loss is stored in $J\omega_0^2 N\dot{N}$. Booster head follows $h_{B0}N^2$.

The twelve equilibrium stages use the shared stage-stack component balance, with a stage-specific energy correction to the vapor flow. The subcooled reflux enters stage 1, expanded gas stage 5, separator liquid stage 7, and the control tray is stage 11. For component j in stage i, the stored liquid balance is

$$\begin{aligned} \frac{d(M_i x_{ij})}{dt} &= L_{i-1}x_{i-1,j} + V_{i+1}y_{i+1,j} \\ &\quad + \sum_{k \in i} F_k z_{k,j} - L_i x_{ij} - V_i y_{ij}. \end{aligned} \tag{9.3}$$

where flows in kmol/h are divided by 60 on the minute clock. The stage energy relation adjusts V_i for incoming vapor and liquid sensible heat, feed quality and side heat. Setting that correction to zero would remove an important path by which expander refrigeration and reflux subcooling affect the column. The upper and lower side reboilers warm liquid leaving stages 8 and 10, while the bottom reboiler receives feed heat and a separately manipulated trim duty Q_R. Constant stage liquid holdup is 15 kmol and sump target 80 kmol. The dynamic sump level regulator changes NGL withdrawal, so instantaneous component recovery can exceed 100% while inventory drains.

For clarity, the stage heat correction used in Eq. (9.3) is

$$\begin{aligned} \varepsilon_i &= \frac{1}{\lambda_i}\big[V_{i+1}\{h_V(T_{i+1}) - h_V(T_i)\} \\ &\qquad + L_{i-1}(h_{L,i-1} - h_{L,i}) + Q_i\big], \\ V_i &= V_{i+1} + \sum_{k \in i}(1 - q_k)F_k + \varepsilon_i, \\ q_k &= \{h_V(T_i) - h_{F,k}\}/\lambda_i. \end{aligned} \tag{9.4}$$

The liquid and vapor cascades are recalculated in four passes. Side-reboiler duties are evaluated in the second pass, and the subsequent passes update sensible-heat correction and sump boilup. A floor of 5 kmol/h prevents a stage from condensing its entire upward vapor stream, while a nonnegative liquid-flow bound prevents evaporating more liquid than arrives. The sump boilup also has a dry-inventory bound proportional to its liquid holdup. These numerical and physical bounds are included in the declared parameter defaults.

Table 9.1 Plant Parameter Defaults. V is shown in the parameter panel. H is hidden. All values are teaching choices. The key identifies the corresponding source parameter.

Key	Meaning	Value	Unit	Panel
`Tin`	Inlet gas temperature	40	°C	V
`Pin`	Inlet and cold separator pressure	80	bar a	V
`split`	Feed share through the reboilers (E-2)	0.27		V
`UAh1`	E-1 conductance, feed side	5	MW/K	H
`UAc1`	E-1 conductance, residue side	5	MW/K	H
`Cm1`	E-1 metal heat capacity	60	MJ/K	H
`UAh2`	E-2 conductance, feed side	1.2	MW/K	H
`UAc2`	Side reboiler conductance, column side	0.3	MW/K	H
`UAcB`	Feed-heated bottom reboiler conductance	0.4	MW/K	H
`Cm2`	E-2 metal heat capacity per reboiler	10	MJ/K	H
`UAsc`	Reflux subcooler conductance	1	MW/K	H
`Csc`	Reflux subcooler metal heat capacity	12	MJ/K	H
`etaExp`	Expander isentropic efficiency	0.87		V
`hB0`	Booster isentropic head at design speed	0.833	MJ/kmol	H
`etaB`	Booster efficiency	0.78		H
`Jshaft`	Shaft inertia (J ω_0^2)	15	MJ	H
`fShaft`	Bearing and seal losses (share of design power)	0.02		H
`Wexp0`	Design expander power	1.5	MW	H
`Mtray`	Liquid held on each stage	15	kmol	H
`Msump`	Sump holdup at its level setpoint	80	kmol	H
`kLevel`	Sump level loop gain	60	1/h	H
`kDry`	Boilup limit per kmol in the sump	400	1/h	H
`Vmin`	Least vapor a stage sends up	5	kmol/h	H
`Kflood`	Tray capacity at flooding	14650	kmol/h sqrt (K/bar)	V
`kE`	Efficiency lost per unit of flooding	8		H
`Emin`	Least stage efficiency when flooded	0.4		H
`tauFl`	Flooding onset lag	1	min	H
`tauReb`	Trim reboiler response	2	min	H
`Qmax`	Trim reboiler capacity	3	MW	V
`specC1`	Bottoms C1/C2 limit	2	mol %	V
`CP`	Gas held per bar between the column and the compressor	15	kmol/bar	H
`dPres`	Pressure drop through the cold box	1	bar	H
`dPac`	Pressure drop through the booster aftercooler	0.5	bar	H
`dTac`	Aftercooler approach to the cooling water	10	K	H
`etaC`	Residue compressor efficiency	0.78		H
`Fc0`	Compressor flow at full capacity and reference suction	6000	kmol/h	H
`Psuc0`	Reference suction pressure	34	bar a	H
`Tsuc0`	Reference suction temperature	40	°C	H
`Wmax`	Residue compressor driver limit	4.2	MW	V
`Psales`	Sales line pressure	70	bar a	H
`cp0`	Ideal-gas heat capacity	0.036	MJ/(kmol K)	H
`Agas`	Real-gas constant A (v = RT/P − A/T^2)	420	MJ K^2/(kmol bar)	H
`Pk`	Convergence pressure of the K values	140	bar a	H

For a hot-stream cell, the implicit outlet temperature solves

$$F_h(h_{in} - h_{out}) = UA_h(T_{out} - T_m), \qquad C_m\dot{T}_m = Q_h - Q_c. \tag{9.5}$$

Flows are converted from hourly units before calculating MW, and the metal derivative is converted to the minute clock. The cold-side enthalpy calculation runs in the opposite direction through the metal nodes. E-2 uses a flash of each column-side liquid, so boiling changes its temperature rather than being represented by a constant liquid heat capacity. The feed enthalpy is interpolated on a flash-based temperature table that is regenerated when composition or pressure changes. Table 9.2 identifies the physical variables separately from the source keys in Table 9.1.

Each exchanger cell moves enthalpy between a stream and a metal state. E-1 has 48 cells, the reflux subcooler 16, and E-2 has three metal states. The conductances and heat capacities in Table 9.1 are generic. The cell count matters: the same design calculation gives 84.5% ethane recovery with 24/8 E-1/subcooler cells, 88.8% with 48/16 and 90.8% with 96/32. Thus the reported 88.8% is a result at a declared resolution, not a converged property prediction.

Table 9.2 Variables Used in the Plant and Control Model

Symbol	Unit	Meaning
F, z_j	kmol/h, –	Inlet flow and composition
M_i, x_{ij}, y_{ij}	kmol, –	Stage liquid and phase compositions
L_i, V_i	kmol/h	Stage liquid and vapor flows
T_i, T_m	K	Fluid and exchanger metal temperatures
P	bar absolute	Column top pressure
$B, x_{B,j}$	kmol/h, –	NGL flow and composition
N	–	Shaft speed relative to design
W_X, W_K	MW	Expander and residue-compressor power
Q_R, Q_{E2}	MW	Trim and feed-supplied reboiler heat
r, b	–	Reflux split and subcooler bypass
c_m	mol %	Reported NGL methane/ethane ratio
m	\$/h	Incremental processing margin

Column pressure accumulates top vapor minus compressed residue with capacity $C_P = 15\,\text{kmol/bar}$. The residue compressor flow scales with its commanded capacity and suction density until the calculated power reaches its 4.2 MW driver limit. Vapor load is the maximum of $V_i\sqrt{T_i/P}/K_f$ across stages, with $K_f = 14650$ in the code's units. Exceeding one initiates a one-minute flooding lag and reduces effective stage efficiency. The 2% bottoms C1/C2 limit, power limit and flooding threshold are intentionally declared teaching values. Real thresholds come from a column, compressor and product specification for a particular installation.

9.3.3 Economics and Numerical Checks

For NGL component flows B_j, the processing margin is the incremental value of liquids relative to leaving those molecules in the gas:

$$m = \sum_{j=2}^{4} B_j(p_{L,j} - p_G H_j) - p_W W_K - p_Q Q_R. \tag{9.6}$$

Here $p_{L,j}$ is the liquid price converted from cents per gallon to dollars per kmol, p_G is dollars per MMBtu, H_j is gross heating value, and power and heat are charged in dollars per MWh. Methane in the NGL earns no separate liquid credit. Base prices are 25, 75 and 100 cents per gallon for ethane, propane and the butane lump, gas is 3 dollars per MMBtu, power 60 dollars per MWh and trim heat 12 dollars per MWh. These are scenario inputs. The ethane frac spread, liquid price less heating-value opportunity cost, is $+5.12$ cents per gallon at the base prices and -9.88 when ethane falls to 10 cents per gallon.

> ✦ **DID YOU KNOW? Ethane Leads the Liquids**
>
> In 2025 the natural gas processing plants of the United States produced an average of 7.5 million barrels a day of natural gas liquids, and 3.1 million of them were ethane, more than any other liquid [13]. Ethane is also the one liquid a plant like this chapter's can choose to leave in its sales gas.

The initial condition is a nonlinear steady solve at the selected mode's pressure and tray setpoints. Changing a starting control before Run solves a different initial condition, while changing it during a run applies a step. Reset and the reload button behave as Section 1.6.1 describes, and Reset solves the start again with the reader's choices. The load-path test confirms that each of the nine cases gives the same CSV loaded first, selected after another case, or reset with its original settings.

The disturbed-run component ledgers compare net inlet minus products with the accumulated column and sump component inventories. The simulator's regression checks close every component to below 10^{-6} kmol for a trip, restart and feed change. A separate design energy check includes feed enthalpy, warm residue, NGL, expander work and trim heat and closes within 0.02 MW. This is an internal conservation check, not evidence for the selected constitutive laws. A two-to-one-second refinement after a reboiler step changes the reported temperature and recovery by less than 0.02 in their displayed units. The larger cell-count sensitivity reported above is a separate spatial approximation.

9.4 Regulation, Sequencing and Economic Supervision

9.4.1 Two Regulatory Loops and a Delayed Composition Trim

The pressure PI increases residue-compressor capacity when pressure exceeds its target. Its gain is 6% of rated capacity per bar, integral time 8 min, and command bounds 0–120%. The tray-temperature PI increases trim duty when stage 11 is colder than its target, using 0.02 MW/K, 15 min integral time and 0–3 MW bounds. Both use the framework's velocity-form controller of Section 1.6.2, with back-calculation and a tracking time equal to the integral setting. The header's Manual mode hands these two commands to the reader, and the inactive controllers track them for a bumpless return to Auto. Guide vanes, reflux split and subcooler bypass remain operator stations in either header mode. Ordinary PI tuning ideas apply locally [14], but the large operating-mode move also changes the heat integration and the appropriate temperature target.

The product chromatograph samples every 10 min and reports after a 6 min analysis delay at the default cycle. At each fresh recovery-mode report, the analyzer changes the tray target by $K_A(c_m - c_{SP})$, clipped to ± 3 K per report and a total ± 25 K. The default K_A is 1.5 K per percentage point per report and $c_{SP} = 1.5\%$, below the 2% limit. The trim is held during a switch, an expander trip or a compressor-limit condition. A high C1/C2 therefore warms the recovery tray only after the report arrives. The tray sensor is an inferential measurement in the sense of Section 1.4.9, not an instantaneous assay of the bottoms. Chapter 8 trims a setpoint from a sampled, delayed analyzer in the same way.

9.4.2 A Mode Is a State and a Procedure

Recovery owns pressure and stage-11 targets of 23 bar absolute and $-1.3\,^{\circ}$C. Rejection owns 27 bar absolute and $45.5\,^{\circ}$C. The interface names RECOVERY, REJECTION or SWITCHING with a step count. A mode is reached when the effective setpoints finish moving, even though the column inventory may still be settling. The flow-diversion supervisor of Volume 8, Chapter 13 is another mode that is held as a state with its own procedure.

The Loops architecture steps both setpoints when the requested mode changes. The Sequencer instead ramps one target over a default 20 min, then ramps the other over 20 min. Its Declared order moves tray temperature first in both directions. Reversed moves pressure first. A return to recovery sets a positive analyzer-trim margin before the ramp, so the first temperature move stops warmer than

the ordinary recovery target. The delayed analyzer subsequently works that margin down after the mode is reached. The default margin is 12 K, increased to 20 K in the suggested challenge.

This procedure is specific to the implemented dynamics. On the way to rejection, preheating the lower column before raising pressure prevents the methane excursion observed with the reverse path. On the return, neither the order nor the larger margin alone is sufficient in the tested shift. The successful path retains the warm margin while pressure and the richer-feed inventory change. The result supports that tested procedure, not a universal rule for real GSP plants.

The Supervisor evaluates a stored steady-state table every 15 min. It interpolates the lean and rich feed cases using inlet-chromatograph ethane, scales product and energy flows with metered feed, and prices each mode using Eq. (9.6). It requests the other mode only if its estimated advantage exceeds 50 dollars per hour, at least 120 min have passed since the last switch, and no switch is in progress. It then uses the same sequencer. Its estimate prices the destination steady states and has no prediction of transition loss. The supervisor of Volume 9, Chapter 8 likewise chooses setpoints for a regulatory layer from a cost model, there by constrained search.

9.4.3 Predictive Control as a Comparison

An optional linear MPC replaces the pressure and temperature PIs. It is a receding-horizon quadratic program, consistent with the industrial MPC class surveyed by Qin and Badgwell [15], and Section 1.6.3 gives the general setup. Its four-state model was identified near recovery at the design feed. Compressor capacity has pressure gain -0.326 bar per percent with a six-minute lag, tray gain -0.156 K per percent with a forty-minute lag, and power gain 0.0587 MW per percent with a two-minute lag. Trim duty has tray gain 57 K per MW and a twenty-two-minute lag.

The MPC executes each minute with a ninety-minute prediction horizon. It corrects its predicted outputs by the measured current residual, holds that bias over the horizon and carries the internal state forward with the commands actually applied. Default pressure and temperature weights are 10 and 1, while compressor-move suppression is 0.01 and duty-move suppression is 10 in their respective units. Input bounds match the two actuators. A soft power ceiling is 0.05 MW below the driver limit, with quadratic slack penalty 10^4. The model has no flooding or methane state. A failed or nonfinite solution, or a first move outside the box, returns control to the tracked PIs. This small local model fails the mode-switch challenge and is presented as a comparison, not its solution. The software context of the apps follows APMonitor and GEKKO [16, 17], while this plant executes its own local model and controller in the browser.

9.5 Cases Measured on the Simulation Engine

Table 9.3 lists the reader's sequence, and Section 1.6.4 describes how a case is loaded, run and exported. All values below were regenerated on the built engine at the default seed, with operator actions applied during the run at their stated times. Measured recovery is a ten-minute filtered ratio of measured NGL ethane to measured inlet ethane. True recovery uses instantaneous component flows. The difference matters while liquid inventory changes. C1/C2 is 100 times bottoms methane mole fraction divided by bottoms ethane mole fraction, not methane percent of total NGL.

9.5.1 Refrigeration and Pressure Tradeoffs

In case 1, the header is Manual. Closing guide vanes to 50% at 10 min reduces shaft work and leaves more separator vapor on the JT path. By 70 min, pressure is 28.07 bar absolute, shaft speed about 68% and cold-separator temperature $-18.4\,^{\circ}$C rather than the initial $-27.1\,^{\circ}$C. Measured ethane recovery is 41.3%, compared with instantaneous true recovery 32.6%. The measured value first falls below 60% at 39.7 min. Restoring full guide vanes at 70 min brings it back above 80% at 153.5 min. The warm separator is evidence of the reboiler/feed thermal loop, while the delayed recovery is evidence of both inventory and filtering.

Case 2 separates pressure from the expander station. At fixed NGL C1/C2 of 1.5%, the stored steady map has the values in Table 9.4. The maximum base-price margin within the driver capacity is

Table 9.3 Nine Cases and Their Main Operator Action

No.	Case	Action or comparison
1	Expander	Close guide vanes, then restore
2	Pressure map	Lower and raise column pressure
3	Specification	Enable delayed analyzer trim
4	Reflux	Change reflux split or bypass
5	Trip	Restart the expander after the trip
6	Flooding	Raise rejection pressure at high feed
7	Switching	Compare the two setpoint orders
8	Economics	Let the supervisor respond to prices
9	Challenge	Meet the twelve-hour target

Table 9.4 Steady Pressure Map at Base Prices and 1.5% C1/C2

P (bar absolute)	C2 recovery (%)	Power (MW)	Margin ($/h)
21.0	97.9	4.15	3147
21.5	96.5	4.03	3150
23.0	88.8	3.74	3138
25.0	76.4	3.39	3113

near 21.5 bar absolute, not at maximum ethane recovery. At 21 bar the compressor uses 4.15 MW, close to its 4.2 MW ceiling. A running pressure reduction to 21.5 bar at 2 min yields 96.1% measured recovery at 92 min. Raising pressure to 25 bar at that instant gives a chromatograph report of 3.17% C1/C2 at 122 min until analyzer trim compensates. A pressure map with a fixed composition target is therefore not the transient path of an untuned column.

9.5.2 Inferential Quality, Reflux and Equipment Limits

Case 3 holds tray temperature while the feed becomes richer at 20 min. With analyzer trim Off, the chromatograph exceeds 2% at 46 min, and true C1/C2 reaches 3.33% at 150 min. Switching trim On then brings the reading to 1.8% or less from 196 min. The warmer setpoint costs recovery: the measured value changes from about 88.9% to 66%. A controlled tray temperature has not guaranteed a controlled product composition.

Pause. Try Exercise 9.3 before reading on.

> **TRY IT IN THE APP** **3 · The methane specification: tray temperature and analyzer**
>
> Load the case and press Run. Watch the Tray 11 temperature and Bottoms C1/C2 trends on the Column view after the feed turns richer at 20 min. The tray holds its setpoint, yet the chromatograph reads above 2 % from 46 min. Near 150 min set Analyzer trim to On on the TIC + AIC card: the reading is 1.8 % or less from 196 min.

Case 4 changes the reflux branch after 10 min. At 150 min, reducing split to 15% gives 77.3% recovery, while increasing it to 32% gives 91.5%. A 50% subcooler bypass gives 65.7% recovery and 4.13 MW compression, with the top at $-88.5\,^{\circ}$C. The expander is still running in all three trials. The comparison makes the additional top reflux mechanism visible rather than assigning every recovery change to expander efficiency.

The expander trips at 15 min in case 5. Its gas moves to the isenthalpic bypass and its booster loses shaft drive. The residue compressor reaches its power limit at 15.87 min, column pressure is 29.53 bar

at 30 min and 30.15 bar at 60 min. The chromatograph exceeds 2% at 36 min, and measured recovery falls below 40% at 47.7 min. A restart at 60 min opens guide vanes at 20% per minute and restores measured recovery above 80% at 167 min. During the inventory recovery, instantaneous propane recovery exceeds 100%, reaching about 137% at 90 min. The numerator includes previously stored liquid, so that value does not violate component conservation.

Pause. Try Exercise 9.2 before reading on.

Case 6 starts in rejection. A 15% feed increase at 20 min causes flooding at 29.3 min, with maximum vapor load 101.3% of the declared capacity. Raising pressure to 29 bar at 60 min reduces load to 99.2% by 240 min, at a propane-recovery cost from about 92.2% to 87.0%. The case uses greater feed rate because making the feed richer alone reduced the top-stage load in the tested model. This is a physical distinction between vapor throughput and a verbal expectation that every heavier feed must flood.

9.5.3 Order, Prices and the Twelve-Hour Shift

Case 7 requests rejection at 20 min with twenty-minute ramps. The pressure-first path reaches true C1/C2 of 3.40% at 43.1 min, with the worst reported chromatograph value 3.17%. The tray-first path stays near 1.51%. Both finish moving their setpoints at 60.2 min. Slowing the reversed ramps from five through 120 minutes does not meet the sampled limit: the tested worst readings span 2.10–4.66%. The infeasibility claim is limited to that ramp sweep with the same reversed order and other settings.

Pause. Try Exercise 9.1 before reading on.

Case 8 changes ethane to ten cents per gallon at 60 min. The supervisor requests rejection at that evaluation and finishes its setpoint sequence at 100.4 min. The measured ethane recovery is still 39.6% at that instant, demonstrating why the mode label does not mean a settled inventory. Actual margin bottoms near 97 dollars per hour at 108.5 min during the move, then reaches 2984 dollars per hour at 150 min, compared with a contemporaneous recovery-mode table estimate of 2322. A profitable destination can have an expensive path.

> **TRY IT IN THE APP 8 · The frac spread flips**
>
> Press Run and open the Economics view. At 60 min the ethane price falls to 10 cents per gallon and the supervisor asks for rejection. Watch the Margin trend: the setpoints reach rejection at 100.4 min while the measured recovery still reads 39.6 %, the margin falls to near 97 dollars per hour at 108.5 min, and it reaches 2984 dollars per hour by 150 min.

The challenge schedules rejection at 180 min, a C2+ feed increase from 10 to 12 mol% at 360 min, and recovery at 540 min. Its whole-run target is C1/C2 at or below 2% at each chromatograph sample, no flooding, no time at the compressor limit, and integrated margin at least 95% of the better mode's lined-out margin at the current feed and prices. The quality score uses the true composition when each sample was drawn, not its noisy report. The controller sees only the delayed report. Changing model parameters, disturbance sliders or event buttons makes the challenge ungraded.

Table 9.5 separates the two requirements of the successful return. Declared order with the original 12 K margin fails at a sample near 600 min, and the 20 K margin with reversed order fails on the outward switch. The suggested combination gives a 1.511% worst sample at the default seed, peak vapor load 94.43%, no driver-limit time and 98.15% of the better-mode integral. A supervisor using the same sequence also passed a four-seed comparison when the app was built. The optional local MPC follows its pressure and tray targets but reaches 3.69% at a sample and 101.7% vapor load, quantities that its local prediction model does not constrain. Figure 9.2 sets the six configurations against the three limits of the target.

The default seed and seeds 0–29 were repeated at two-second and one-second physics steps with the control and instrument clocks held fixed. All 62 shipped trials fail and all 62 suggested trials pass. Suggested worst samples range from 1.5082 to 1.5369%, with peak vapor load no greater than

Table 9.5 Default-Seed Twelve-Hour Scorecard. Each row differs only in the settings named.

Configuration	Worst sample C1/C2 (%)	Peak vapor load (%)	Driver-limit time (min)	Margin share (%)	Verdict
Shipped: reversed, 12 K margin	3.65	95.2	0	97.7	Fail
Declared order only	3.20	94.4	0	98.2	Fail
20 K margin only	3.14	95.2	0	97.6	Fail
Suggested: declared, 20 K margin	1.51	94.4	0	98.2	Pass
Both setpoints stepped	11.36	93.6	0	98.8	Fail
Suggested sequence with MPC	3.69	101.7	0	98.7	Fail

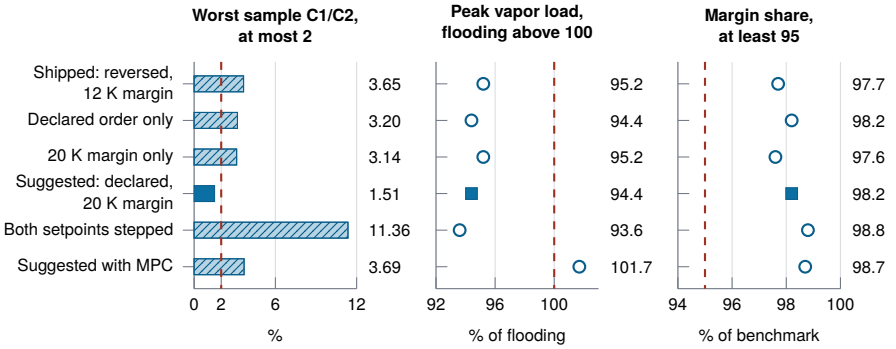


Figure 9.2 The twelve-hour scorecard of Table 9.5 against the target. The solid bar and the filled squares are the one configuration that passes, hatched bars and open circles the five that fail, and dashed lines the limits. Notice that the configuration with the largest margin share has by far the worst sample, and that only the predictive controller crosses into flooding. No row spends time at the driver limit.

94.437% and margin from 98.147 to 98.155% of the benchmark. This tests the declared noise and numerical step, rather than the reliability of a real plant.

> ✦ DID YOU KNOW? **The Richest Run Failed Worst**
>
> From the simulator (Table 9.5). Stepping both setpoints at once earned the largest margin share of the six configurations, 98.8 % of the better-mode benchmark, and also the worst sample by far, 11.36 % C1/C2 against the 2 % limit. The one configuration that passed earned 98.2 %.

9.6 Technology in Practice

Figure 9.5 draws the loops of Section 9.4 with the hardware of a gas plant in place of the schematic's symbols.

What Measures It

The NGL specification is read by one instrument, the slowest of the plant: the product chromatograph AT on the column's bottoms. The two regulatory loops read fast instruments instead: a pressure transmitter at the column top and a temperature element on stage 11, which infers the bottoms composition only as long as pressure and feed hold still. The supervisor reads a third pair, the inlet gas meter and an inlet chromatograph that gives the feed's ethane.

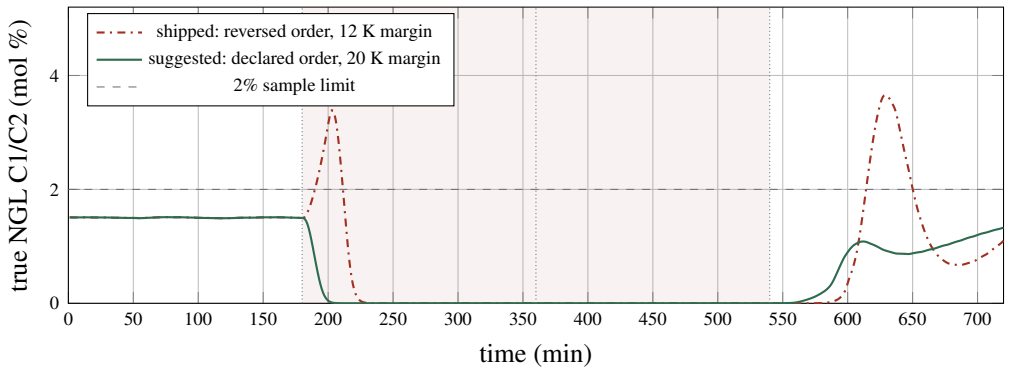


Figure 9.3 True NGL C1/C2 through the challenge, regenerated from the engine at one-minute spacing. The shaded interval has low ethane price. Vertical dotted lines mark the rejection request, richer feed and recovery request at 180, 360 and 540 min. The dash-dotted path combines reversed order with the original 12 K return margin. The solid path uses the declared order and 20 K margin. The dashed line is the 2% limit. The scorecard checks sample instants on these trajectories rather than every plotted instant.

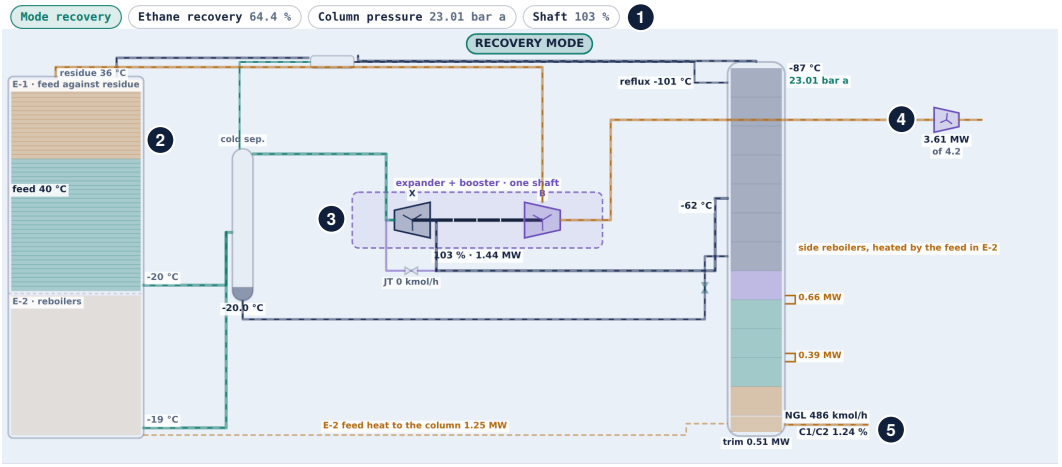


Figure 9.4 Suggested challenge at 720 min in the Plant view. Callout 1 marks the status chips: mode, measured ethane recovery, column pressure and shaft speed. Callout 2 is the cold box, E-1 above E-2, and callout 3 the expander and booster on one shaft with the JT valve below. Callout 4 is the residue compressor and callout 5 the NGL leaving the sump with its last chromatograph report. The final instruments read 23.01 bar absolute, 64.4% ethane recovery and 1.24% C1/C2, with 3.61 MW residue compression. These are the current values, not the whole-run worst sample or margin integral. The mode is again Recovery while analyzer trim is still reducing its warm return target.

TECHNOLOGY **Sensor: The NGL Chromatograph**

A heater at the inlet of the product chromatograph vaporizes its liquid sample, and the valve and column are kept hot so that it does not condense again, since liquid in a gas analyzer can cause anything from a delay to a gross error. Peak areas are converted with response factors measured on a calibration gas, often injected on a schedule. Separating species in time makes cycle times of minutes and a dead time inherent [18]. The chapter's samples every 10 min and reports 6 min later.

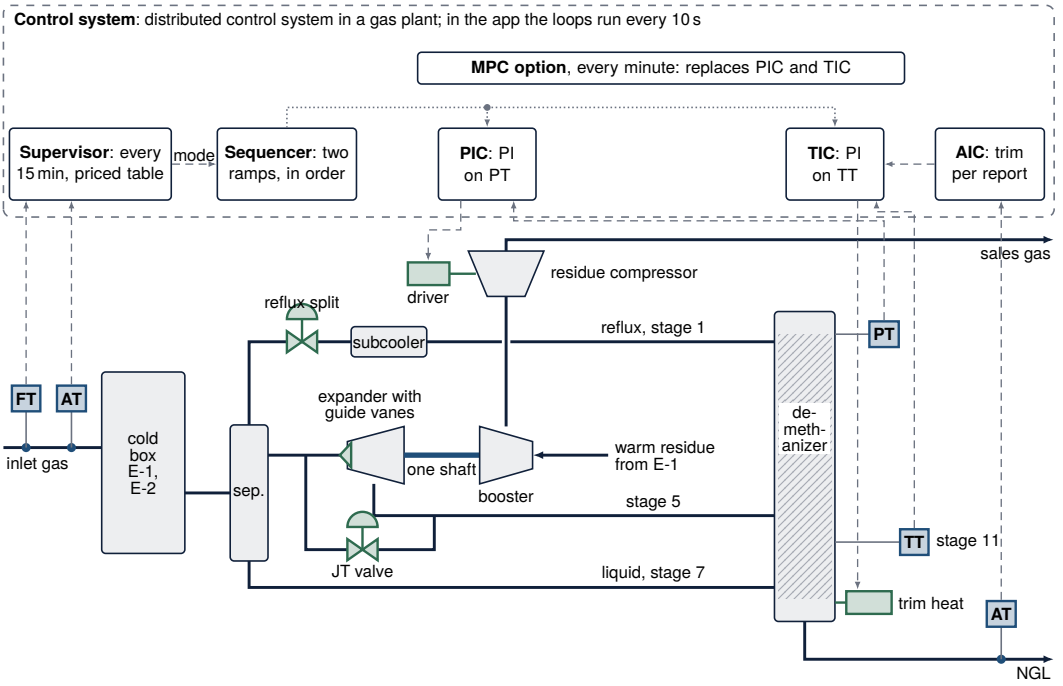


Figure 9.5 The loops of the turboexpander plant with their hardware. Lettered boxes are sensing devices with the chapter's tags: PT at the column top, TT on stage 11, AT on the NGL and FT with AT on the inlet gas. Green symbols are final elements: the expander's guide vanes with the Joule–Thomson valve beside them, the reflux split valve, the residue compressor and the trim reboiler's heat. Dashed arrows are signals, dotted arrows the setpoints the sequencer writes. Notice that the analyzer reaches the reboiler only through the tray setpoint, and that the supervisor reads the inlet meters, not the product. The subcooler bypass, the side reboilers and E-2 are left out. Not to scale, and not a piping diagram.

What Moves It

Five final elements carry the chapter's handles. The residue compressor's capacity holds the column pressure, between 0 and 120 % of its rating and below its 4.2 MW driver. The trim reboiler's heat, from 0 to 3 MW, holds the tray. The expander's guide vanes, the reflux split and the subcooler bypass are operator stations in either header mode, and of these only the guide vanes move refrigeration and recompression together.

> **TECHNOLOGY** **Final Element: The Turboexpander and Its Guide Vanes**
>
> A turboexpander is a turbine wheel driven by the expanding gas, here with the booster's wheel on the same shaft. In gas plants it often gives the coldest refrigeration on the site, and much liquid can condense inside it. The seventh edition of API 617, of 2002, covered cryogenic turboexpanders in its Chapter 4, and magnetic bearings are making inroads in gas processing [19]. The inlet guide vanes set how much gas the expander takes, and the Joule–Thomson valve beside it takes the rest. On a trip the vanes shut and the booster loses its drive, so the column pressure climbs and, in case 5, the residue compressor reaches its limit at 15.87 min.

Where the Controller Runs

Inside a gas plant these loops would run on a distributed control system, which controls production within one location under a supervisory level and nests loops so that one loop's setpoint comes from

another's measurement. Such loops run with cycle times from milliseconds to minutes [20]. The app's controllers execute every 10 s, its MPC every minute and its supervisor every 15 min, while the chromatograph cycles every 10 min, so no scan rate repairs the analyzer's delay. The chapter's structure maps onto standard blocks: two PI blocks with back-calculation for pressure and tray temperature, a sampled trim that writes the tray setpoint, a sequence of two setpoint ramps for a mode change, and a supervisory layer that chooses the mode from a priced table.

> **IN PRACTICE** **The Cold Box Has a Speed Limit**
>
> Brazed aluminum plate-fin exchangers serve natural gas processing down to about −130 °C [21], and their makers' association sets integrity operating windows, with API 668 as a controlled document. Near the nozzles a stream should change by less than 60 °C an hour and 5 °C a minute in a start-up, shutdown or trip, and by less than 1 °C a minute in steady operation. Stream-to-stream and stream-to-metal differences should stay under about 30 °C. Outside a window the operator stops opening or throttles the valve, and steady swings call for tuning the controllers, because the damage is fatigue that ends in leaks [22]. The 48 metal temperatures of the chapter's E-1 are where such a limit would apply.

What Is Modernizing It

Two developments touch this chapter. The first removes the analyzer from the loop: a 2022 study controlled a demethanizer without composition analyzers, with a cascade on an approximation of the boilup and a pressure compensator, and found it gave the best performance and the lowest energy use among the schemes compared [11]. A pressure compensator speaks to the failure of case 7, a tray temperature that stops meaning a composition when the pressure moves. The second is model predictive control, established in industrial practice [15]. The chapter's comparison shows what it still needs here: a model that carries the methane and the flooding.

9.7 Limitations and Interpretation

The simulated gas is dry and has four pseudo-components. Water freezing, hydrate formation, carbon-dioxide solids, contaminants and heavier-end characterization are absent. The ideal-solution volatility with a convergence correction is not an equation-of-state calculation, and methane's extrapolated vapor-pressure line cannot reproduce a mixture critical locus. The shared heat capacity and gas correction omit composition effects. Agreement with the GSP brochure's broad recovery category at one selected design therefore cannot validate the physical correlations over a plant envelope.

The finite exchanger grid has a visible numerical effect on recovery. Doubling cells from the default changes design ethane recovery from 88.8% to 90.8%. The time-step study and challenge half-step sweep test temporal integration, while that spatial dependence remains. A future model could reduce this cell sensitivity or replace the teaching closure with measured mixture-property data, but the present reported results must retain the stated 48/16 resolution.

Liquid stage holdups are fixed, separator level is assumed controlled, and the flooding-efficiency law is prescribed rather than a hydraulic design correlation. There is one pressure inventory instead of distributed vapor holdup. The machine models omit surge, choking, a detailed expander efficiency map, mechanical protection and compressor-control hardware. All limits are educational boundaries. These omissions make a fast interactive lesson possible but exclude equipment sizing and operating-procedure certification.

The switch study uses one feed family, specified prices and a fixed delay structure. The better-mode benchmark assumes instantaneous lined-out operation and therefore intentionally ignores the cost of getting there. Integrated margin can also temporarily exceed both table modes when the column releases inventory. A more complete economic optimization would price inventories and transition costs explicitly. The supervisor's interpolation and the local MPC are separate approximations, each visible in the source and each limited to the information its instruments provide.

Finally, deterministic seed repetition measures sensitivity to the declared instrument noise, not industrial reliability. The physics step, sampling cycle, noise generator and scripted disturbances define those trials. Reported pass counts have meaning only alongside the actual quality, flooding, power and margin limits. The application provides those quantities rather than replacing them with a single unexplained success label.

OPEN CHALLENGE **A Colder Top Against Frozen Carbon Dioxide**

A colder demethanizer top keeps more ethane, but carbon dioxide in the gas can freeze on the column's internals, and its concentration can rise in the colder stages above its level in the feed [8]. Solid carbon dioxide blocks and plugs equipment and can rupture it, and freezing calculations enter the design of the exchangers and turboexpanders that serve demethanizers. Empirical correlations have not proved trustworthy for industrial use, and process simulators have not reliably matched data even for methane with carbon dioxide [23]. Progress would be a freezing margin predicted well enough to serve, beside flooding and the driver limit, as a constraint of the supervisor.

9.8 Conclusion

This chapter's teaching plant connects ethane economics to a timed physical procedure. At its 23 bar absolute recovery design it captures 88.8% of ethane with 3.74 MW residue compression, while the JT comparison loses about 22 K of refrigeration relative to the expander. A guide-vane closure warms the separator through the integrated reboilers, a richer feed defeats a fixed tray inference, and a profitable mode change briefly reduces actual margin to 97 dollars per hour. These outcomes require the cold box, column inventory and measurement delay to be represented together.

The twelve-hour experiment demonstrates why its successful switch needs both the declared tray-first order and the 20 K return margin. Each change alone fails the sampled 2% C1/C2 target. Their combination keeps the default-seed worst sample at 1.511%, peak vapor load at 94.43% and integrated margin at 98.15% of the better-mode benchmark. The optional predictive controller provides a useful counterexample because its local model omits quality and flooding. The plant is a transparent, locally executed process experiment, and its measured lesson is the distinction between a profitable destination, a feasible path and the delayed evidence available to its controller.

KEY POINTS

- Ethane can be sold as a liquid or left in the gas, and a profitable switch can have an expensive path.
- A controlled tray temperature has not guaranteed a controlled product composition.
- The successful switch needs both the tray-first order and the 20 K return margin, and each alone fails.
- The predictive controller follows its targets but reaches 3.69 % at a sample and 101.7 % vapor load, which its local model does not constrain.

Where else this applies. Chapter 8 trims a setpoint from a sampled, delayed analyzer in the same way, and Volume 8, Chapter 13 holds a mode as a state with its own procedure.

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9.9 Exercises

9.1 ○ Concept check · 2 min

In case 7 with the Reversed order, the sequencer raises the column pressure from 23 to 27 bar absolute first, while the tray loop still holds stage 11 at its recovery setpoint. The true C1/C2 of the bottoms reaches 3.40 %. Which statement explains the rise?

(a) The higher pressure floods the top stage, and the flooded stages send methane down the column with the liquid.

(b) The residue compressor reaches its driver limit, so the pressure overshoots its new setpoint.

(c) Every bubble point rises with the pressure, so liquid at the same tray temperature holds more methane until the tray setpoint is raised.

(d) The analyzer trim cools the recovery tray setpoint during the switch, which lets methane into the bottoms.

9.2 Concept check · 2 min

In case 5 the expander restarts at 60 min, and at 90 min the simulator's instantaneous propane recovery reads about 137 %. How can more propane leave in the NGL than enters with the feed?

(a) Chromatograph noise on the NGL composition inflates the recovery while the reading is near its range.

(b) The JT valve condenses more propane than the feed carries while the expander spins up.

(c) The column's component ledger stops closing during the restart, so propane appears from the integration.

(d) The sump level regulator draws down liquid that the column stored earlier, so the NGL carries propane that entered before.

9.3 Concept check · 1 min

In case 3 the feed turns richer at 20 min. The tray loop holds stage 11 on its −1.3 °C setpoint, yet the chromatograph reads above 2 % C1/C2 from 46 min. Why?

(a) A richer liquid boils warmer at the same methane content, so the held temperature now allows more methane.

(b) The trim reboiler reaches its 3 MW capacity and can no longer hold the tray temperature.

(c) The richer feed floods the column, and the flooded stages pass methane to the bottoms.

(d) The chromatograph's 6 min delay makes it report methane that the bottoms no longer contain.

9.4 Theory and calculation · 8 min

The cold separator vapor leaves at −27.1 °C and 80 bar and expands to 23 bar. The chapter's model gives −70.9 °C after the expander ($\eta_X = 0.87$) and −49.2 °C after the JT valve. Take $R = 8.314\,\mathrm{kJ/(kmol\,K)}$.

(a) Treat the vapor as an ideal gas with $c_{p0} = 36\,\mathrm{kJ/(kmol\,K)}$. Find γ and the expander outlet from Equation (1.13), isentropic and at $\eta = 0.87$. *Checkpoint:* $\gamma = 1.300$, −88.65 °C and −80.65 °C.

(b) An ideal gas does not cool through a valve. Use the chapter's vapor enthalpy, Equation (9.2) with $A = 420$ in MJ, kmol, K and bar units and $P_0 = 1$ bar, to find the vapor-only JT outlet. Start from $T = 217\,\mathrm{K}$. *Checkpoint:* about 216.8 K, −56.4 °C.

(c) Both estimates are colder than the model. The expander outlet in the model is 6.6 % liquid. Explain the difference.

9.5 Theory and calculation · 6 min

Ethane in the NGL is worth its liquid price, and ethane left in the residue gas is worth its heating value, Equation (9.6). The app uses 22.31 gal/kmol of liquid ethane and a gross heating value of 1.4786 MMBtu/kmol, with gas at 3 dollars per MMBtu. The design feed carries 273.9 kmol/h of ethane.

(a) Find ethane's value as gas in cents per gallon, and the frac spread at 25 and at 10 cents per gallon. *Checkpoint:* 19.88 cents/gal, +5.12 and −9.88 cents per gallon.

(b) Recovery keeps 88.8 % of the ethane and rejection 8.1 %. Considering the ethane term alone, what does rejection gain per hour at 10 cents per gallon?

(c) The supervisor requests the other mode only for an advantage above 50 dollars per hour. On the ethane term alone, how far below the breakeven price must ethane fall before rejection clears that hysteresis?

9.6 Theory and calculation · 6 min

The product chromatograph draws a sample every 10 min and reports it 6 min later. In recovery each fresh report moves the tray setpoint by $K_A(c_m - c_{SP})$ with $K_A = 1.5$ K per percentage point and $c_{SP} = 1.5$ %, clipped to 3 K per report.

(a) A methane excursion starts at the bottoms. What are the shortest and the longest times before a report can show it? *Checkpoint:* 6 min and 16 min.

(b) Compute the setpoint move for reports of 1.80, 3.17 and 3.65 %.

(c) After a return to recovery the tray setpoint starts the return margin above its recovery value. If every report then read 0.50 %, how many reports would remove a 12 K and a 20 K margin?

9.7 App challenge · 9 min

Load scenario 5, "The expander trips", default seed. Run at speed 5 (five simulated minutes per second). The expander trips at 15 min.

(a) Without pressing Restart, run to 60 min and pause (anywhere from 60 to 65 min is fine). Read from the exported CSV the first row with `comp_at_limit` equal to 1, `p_meas_bara` at 30 and 60 min, the first row with `c1c2_meas_pct` above 2 and the first row with `recovery_meas_pct` below 40. *Checkpoint:* 16.0 min, 29.52 and 30.17 bar, 36.2 min, 47.8 min.

(b) The Compressor power tile reads 4.20 MW. Explain why the pressure loop cannot hold 23 bar after the trip.

(c) Press Restart the expander (Manual actuator card), run to 240 min and read when the Ethane recovery tile is back above 80 %.

9.8 App challenge · 8 min

Load scenario 7, "The switch to rejection: in what order", default seed. It ships with the Sequencer order Reversed. Run at speed 5 and, between 15 and 25 min, set Mode to Rejection on the Mode card. Run to 120 min and export the CSV.

(a) Read the largest `truth_c1c2_pct` after the request and its time, and the largest chromatograph report, `c1c2_meas_pct`.

(b) Reload the case (↻), set Sequencer order to Declared before Run, and repeat the request at the same time. Read the same two maxima.

(c) Readers who requested at different times report true peaks within 0.02 % of each other but chromatograph peaks that differ by up to 0.5 %. Why?

9.9 App challenge · 9 min

Load the challenge, "a day with two switches", default seed. Run four times to 720 min, where it stops, at speed 100: as shipped, with only the Sequencer order set to Declared, with only the Return margin set to 20 K, and with both. Reload the case (↻) before each run and make the named settings on the Mode card before Run. Read the closing note, or the So far tile's tooltip, after each run.

(a) Tabulate the worst sample C1/C2, the most flooding, the minutes at the compressor limit and the margin share for each run. *Checkpoint:* 3.65, 3.20, 3.14 and 1.51 % worst samples.

(b) From the Bottoms C1/C2 trend, say in which switch each failing run breaks the limit.

(c) Explain why the return to recovery needs both changes.

9.10 Application area · 4 min

In case 6 a 15 % larger feed floods the top stage of the rejecting column at 27 bar, with a peak vapor load of 101.3 %. Raising the rejection pressure to 29 bar at 60 min brings the load to 99.2 % by 240 min, while propane recovery falls from about 92.2 to 87.0 %.

(a) The load is $V\sqrt{T/P}/K_f$. What would the pressure change alone do to the load? Why is the measured relief smaller?

(b) The feed carries 2.5 % propane. How much propane per hour moves to the residue gas, and what other handle could clear the flooding?

9.11 Application area · 3 min

A manager reads Table 9.4 and asks to run the column at 21.5 bar permanently, since the margin there is the highest. Use the table to advise.

9.12 Review the agent · 5 min

An AI assistant compares the rows of Table 9.5 and writes: “With the suggested sequence, the MPC earns 98.7 % of the better mode’s margin against 98.2 % for the PI loops, and it follows its pressure and tray-temperature targets. It is the better controller for the twelve-hour day, so the two PI loops should be replaced by it.” Find the flaw, and say how you would check it in the app.