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CHAPTER 8

A Verified Browser Port of the Tennessee Eastman Plant and Its Shutdown

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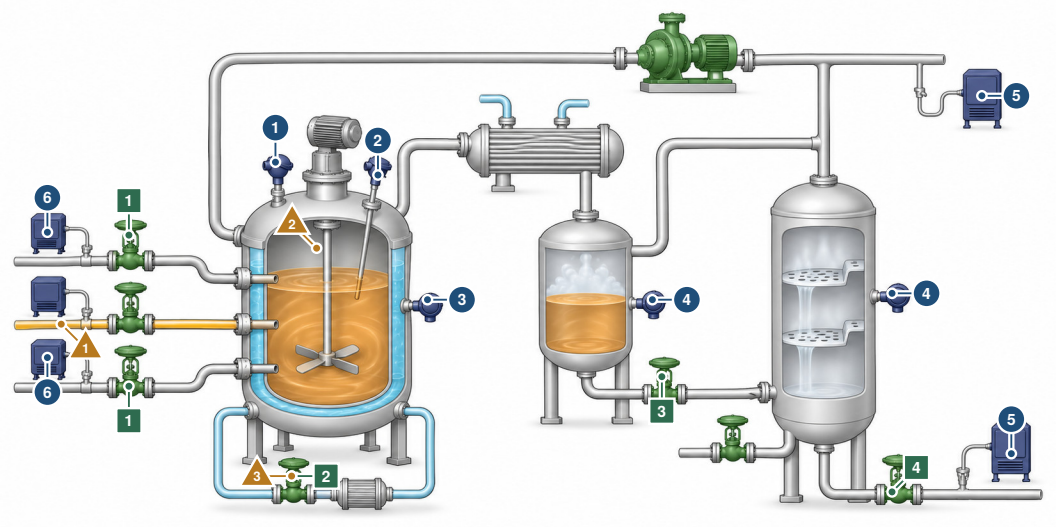
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CHAPTER 8

A Verified Browser Port of the Tennessee Eastman Plant and Its Shutdown



The controller moves 1 feed valves for A, D and E 2 reactor cooling water valve 3 separator liquid valve 4 product valve

The controller reads 1 reactor pressure 2 reactor temperature 3 reactor level 4 separator and stripper levels 5 purge and product analyzers 6 feed flow meters

What disturbs it 1 a feed lost or changing in makeup 2 slow drift of the reaction rates 3 a sticking cooling valve

Notice the gas that returns from the separator through the compressor to the reactor, which holds its pressure only while a loop opens the purge as the inert builds up. Illustration, not to scale.

IN PLAIN TERMS A Plant That Will Not Stand Still

The main idea. This is a model of a whole chemical plant. A reactor makes the product, a cooler and a tank split gas from liquid, and a column strips leftover gas out of the product. Gas that did not react is pumped back to the reactor, so each part feels the others.

The challenge. With every valve left fixed, gas builds up until the reactor pressure stops the plant within four hours. The controls must hold the gas and the liquid levels first, and only then the mix, the output and the cost.

Why it matters. The order in which the loops are closed decides whether such a plant keeps running.

THREE WAYS TO READ THIS CHAPTER

Explore	12 min	the first page, the fact on degrees of freedom in Section 8.1, the history in Section 8.2, the try-its and the fact on a lost feed in Section 8.6, and the key points at the end of Section 8.9
Study	45 min	the whole chapter with its three pause points and the twelve exercises at its end
Practice	15 min	the first page, the results (Section 8.6), Technology in Practice (Section 8.7), Limitations (Section 8.8) and the open challenge

The control question The published base case of the Tennessee Eastman plant is a steady state that is not stable, and with every valve fixed it trips on high reactor pressure within hours. Which loops must close first, what must a structure above them do for production, product mix and cost, and how far can a port of the public code be trusted?

Learning objectives

- Explain why a gas recycle with an inert and a purge on a fixed valve does not hold its own pressure.
- Order the layers of a plantwide structure, inventories before composition, production and cost, and support the order with the cases.
- Compute what the production index and its rate limit do to the flow setpoints and to the challenge's production score.
- Describe how the integrating pressure override trades production for margin, and tell a failure of a structure from a limit of the plant.
- Judge a reported trip time or cost as evidence about an implementation as well as about a plant.

Prerequisites Sections 1.4.2, 1.4.3 and 1.4.8 (degrees of freedom, inventories, levels as buffers), 1.6.2 and 1.6.4, with Section 1.6.3 for the predictive structure only. Chapter 7 treats recycle and the production handle on a simpler plant.

The app https://apmonitor.com/apps/tennessee-eastman/

See also Volume 4, Chapter 2 for a protective override that is idle in normal operation. Volume 1, Chapter 12 for step-test models of a multivariable plant and Volume 1, Chapter 9 for inventories judged by the time to a limit. Volume 13, Chapter 2 for an inert purged from a gas recycle under a pressure override, and Volume 13, Chapter 11 for optimization above the loops.

8.1 Introduction

Four gaseous feeds enter a stirred reactor, three reactions make the liquid products G and H, a condenser and a separator return the unreacted gas through a compressor, and a stripper recovers the light components from the product. At the base case of the public model this plant makes 22.95 m^3/h of product at a reactor pressure of 2705 kPa g and 120.4 °C, for 170.62 $/h by its own cost function. With all twelve valves frozen at that point and no disturbance, the reactor pressure still reads 2705 kPa g at 2 h, falls to 2690 at 3 h, and then climbs to the 3000 kPa g interlock at 3.59 h, which stops the plant.

That run corrects the belief that a plant left at a steady state stays there. A gas recycle with an inert, whose only exit is a purge through a fixed valve, does not hold its own pressure, and the liquid levels on pumped outlets are integrators. Section 1.4.3 treats such an inert as a component inventory, and the ammonia synthesis loop of Volume 13, Chapter 2 purges one in the same way. The plantwide procedures written since the benchmark appeared put these inventories first, before production rate and product quality, and they treat the class as general to plants with recycle, among them the reactor and separator flowsheets that the procedures were illustrated on [1–3]. A student who has tuned single loops has not met a plant in which the order of the loops decides whether it survives the afternoon.

DID YOU KNOW? Twelve Valves, Three Free Choices

The plant has 12 manipulated variables, yet a degree-of-freedom analysis finds only 8 degrees of freedom at steady state. At the economic optimum of the base operating mode 5 constraints are active, which leaves 3 unconstrained degrees of freedom for the control structure to settle by its choice of controlled variables [4].

The same plant corrects a second belief, that a port of a public benchmark reproduces the benchmark. The open-loop trip time above is set by the residue left when the constants are rounded. With the model's decimal constants read in double precision the trip comes at 3.59 h, with the single precision of a standard Fortran compiler at 4.06 h, and with an independent C translation's reading of five of them at 3.47 h, on the same one-second Euler step. However, the code itself was revised in

2015 because its results had stopped being repeatable in current simulation software [5], and much recent use of the plant is as a data set for fault diagnosis [6] rather than as a plant to hold.

This chapter carries the Tennessee Eastman model into a browser through a JavaScript port, states where the port came from and how closely it agrees with the code it was checked against, and measures what its control structures do with the plant. The model is Downs and Vogel's [7], and the decentralized structure follows Ricker's archived design [8]. The results are measurements, among them that this archived structure trips on the stripper level at 8.40 h after the A feed is lost, while Ricker's own recording of his 1996 strategy rides the same loss. The port is checked against three references, and its provenance, its licenses and the equations and settings from which it can be rebuilt are recorded here. Ten cases are provided, the tenth a two-day challenge redesigned so that its shipped structure fails and its suggested one passes at six seeds.

Section 8.2 traces the benchmark. Section 8.3 sets out the public model and the port, and Section 8.4 compares the port with the Python port, the C translation and the tabulated base case. Section 8.5 describes the inventory loops, the decentralized structure, the pressure override and the predictive controller. Section 8.6 measures the ten cases, Section 8.7 describes the real analyzers, valves and controllers, Section 8.8 states the limitations, and Section 8.9 concludes. The Exercises at the end of the chapter run the cases and the challenge.

8.2 A Challenge Problem From Kingsport

The problem came from industry. J. J. Downs and E. F. Vogel of the Process and Control Systems Engineering group of the Tennessee Eastman Company in Kingsport, Tennessee, wrote a simulation of an industrial process whose eight components are named only by the letters A to H, and distributed it as Fortran subroutines. The header of that code records a presentation at the AIChE Annual Meeting in Chicago on November 14, 1990, in a session on industrial challenge problems in process control, and a revision of April 4, 1991. The journal paper followed in 1993 [7]. The code defines 50 states, 12 manipulated variables (XMV), 41 measurements (XMEAS), of which 19 come from three sampled analyzers, 20 disturbances (IDV) and eight shutdown tests.

Control structures were proposed within two years. A base control layer was designed for the problem in 1994 [9], a plantwide structure and a control configuration design followed in 1995 [10, 11], and Ricker computed the plant's optimal steady operation [12] and, with Lee, applied nonlinear model predictive control to it [13]. His decentralized design of 1996 [8] is the one an implementer can rebuild without the paper, because he archived it as Simulink models with the code, his tabulated steady states and his closed-loop recordings at the University of Washington. The self-optimizing analysis of Larsson et al. recommends controlling the reactor pressure at its maximum and the reactor level at its minimum, closing the compressor recycle and stripper steam valves, and running the agitator at its maximum, and it finds that holding the inerts at a setpoint is a poor choice for this plant [14]. Ricker's tabulated optimized operating points all run at 2800 kPa g, against the base case's 2705.

> ⧖ HISTORY **1995 to 1996: Decentralized Loops Against Predictive Control**
>
> Ricker and Lee applied nonlinear model predictive control to the challenge plant and compared it with an earlier multiloop strategy. The decentralized loops performed adequately in many scenarios, but for the harder cases they would have needed multiple overrides, where the predictive controller might be preferred. In a second study Ricker designed decentralized control of his own. He controlled the variables that should optimally sit at their constraints, left out those whose best value varies a lot, and installed overrides [4]. He found little, if any, advantage in the nonlinear predictive controller: the decentralized strategy handled the constraints better, the very area where predictive control was reputed to excel. The problem, he wrote, had too many competing goals and special cases for a conventional predictive formulation [15].

The benchmark also shaped the general procedures. A nine-step heuristic procedure for plantwide control, ordered from energy management and production rate through inventories, component

balances and optimization, used the Eastman process as one of its three industrial examples [1], and control structure design for complete plants was later written as a systematic procedure [2]. One of the plant's authors has since described the industrial and academic views of plantwide control side by side [3]. The plant also serves as a source of process data, and comparisons of fault detection and diagnosis methods are run on it [6].

The code drifted as it moved between platforms. Ricker's archive carries a C translation of the Fortran produced by f2c and a MATLAB interface around it. In 2015 Bathelt, Ricker and Jelali reported that the original program flow no longer holds in current simulation software, so results were not repeatable, and gave a modified code with further measurements and a shorter run time [5]. Their revised code, in the same archive, gives the disturbances and the measurement noise separate random-number states and adds disturbances 21 to 28. The Fortran itself is redistributed in the Tennessee Eastman archive of R. D. Braatz's Large Scale Systems Research Laboratory at the University of Illinois, under that university's notice, and that archive is carried on GitHub in https://github.com/camaramm/tennessee-eastman-profBraatz, whose fork by J. Kitchin adds a pure-Python port.

8.3 The Public Model and Its Port

8.3.1 Provenance and Licenses

The application's model file, `teprob.js`, was translated routine by routine (TEFUNC, TEINIT and TESUB1 to TESUB8) from `tep/python_backend.py` in the fork https://github.com/jkitchin/tennessee-eastman-profbraatz at commit `9a6c8e5`, whose `pyproject.toml` and README declare the BSD-3-Clause license. Every block was then checked against `teprob.f` as the same repository distributes it from the Braatz archive, under the University of Illinois notice of 1998 to 2002. Where the Python and the Fortran differ, the Fortran was followed. Ricker's archive supplied the base-case table against which the initial state is checked, the decentralized structure in `MultiLoop_mode1.mdl`, the C translation `temexd.c` used as an independent check, and four of his closed-loop recordings. From the 2015 revision only the split random-number streams were taken, because the revision changes no plant equation, its solver corrections do nothing at a fixed Euler step, and its added disturbances lie outside this application. Both license notices are reproduced in full in the application's Model dialog, as their conditions ask of a redistribution in compiled form, and at the head of `teprob.js`. No Fortran, Python or WebAssembly compiled from either is shipped.

8.3.2 The Plant

Feeds of A (stream 1), D (2) and E (3) enter a header with the compressor discharge (8) and the stripper overhead (5), and the mixture (6) enters the reactor, as shown in Figure 8.1. The reactor effluent vapor (7) is cooled in a condenser and separated, the vapor returns through the compressor, whose recycle valve (XMV 5) spills discharge back to suction, and a purge (9) leaves from the compressor suction. The separator liquid (10) goes to the stripper, where the A and C feed (4) strips the light components back to the header and the product (11) leaves the foot, heated by steam. Table 8.1 classifies the variables. The code's internal units are pound-moles, cubic feet, millimeters of mercury and degrees Celsius, and the measurements leave it in the published units.

8.3.3 Balances

Each of the four vapor or liquid holdups carries eight component holdups and an internal energy. With $x_{k,i}$ the mole fraction of component i in stream k and F_k its molar flow, the component balances

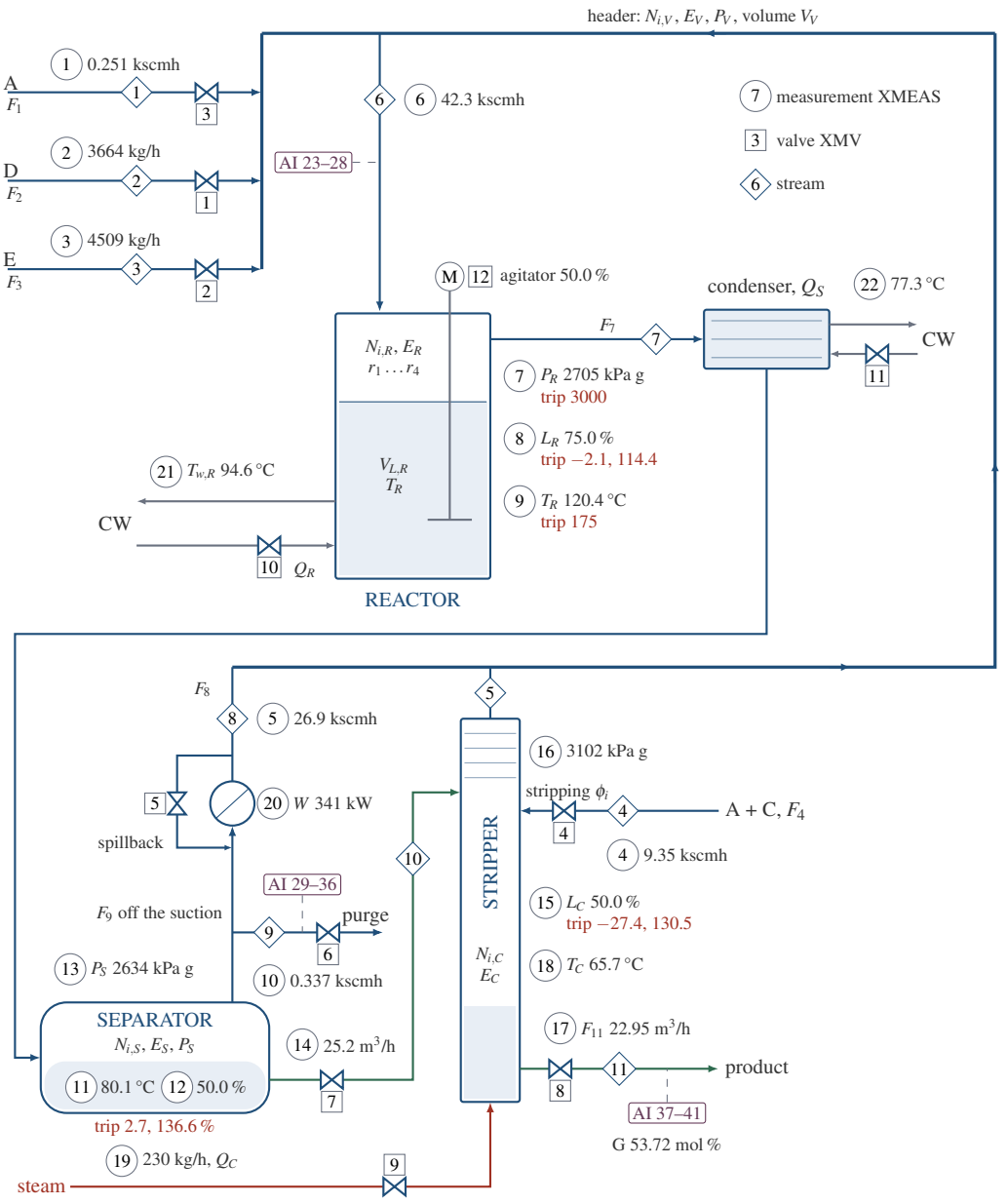


Figure 8.1 The Tennessee Eastman plant as modeled, after the application's Plant view of case 1 at $t = 0$, the published base case, drawn in two bands joined by the effluent from the condenser and by the header, which returns the compressor discharge (8) and the stripper overhead (5) up the right edge. Streams 1 to 11 carry their published numbers in diamonds, each circle is a continuous measurement (XMEAS 1 to 22) with its base-case value, each square a valve (XMV 1 to 12), and the three analyzer boxes give the XMEAS ranges they report. The lines marked trip give the interlock limits on the watched variables, the levels in their transmitters' percent. The holdups $N_{i,u}$ and energies E_u of the reactor (R), separator (S), stripper (C) and header (V) are the states of Equations (8.1) and (8.2). Process model only, not to scale.

Table 8.1 Variables by Role, With Base-Case Values

Variable	Role	Base case, range
F_{11}, XMEAS 17	CV, production	22.95 m^3/h, SP 15–32
x_G, XMEAS 40	CV, G in product (analyzer)	53.72 mol %, SP 5–95
P_R, XMEAS 7	CV, reactor pressure	2705 kPa g, trip 3000
L_R, L_S, L_C	CV, three levels (XMEAS 8, 12, 15)	75.0, 50.0, 50.0 %
T_R, XMEAS 9	CV, reactor temperature	120.4 °C, trip 175
y_A, y_{AC}	CV, feed analyzer (XMEAS 23, 25)	54.96 %, 58.57 mol %
$v_1 \dots v_{12}$	MV, XMV 1–12	Figure 8.1, 0–100 %
IDV 1–20	DV, steps, random walks, sticking	off, on or off
$N_{i,u}$, E_u	states, $u \in \{R,S,C,V\}$	36
$T_{w,R}$, $T_{w,S}$	states, cooling-water outlets	94.6, 77.3 °C
v_j	states, valve positions	12
F_p	controller, production index	100 %
J	output, operating cost	170.62 $/h

are Equation (8.1).

$$\begin{aligned}
\frac{dN_{i,V}}{dt} &= \textstyle\sum_{k\in\{1,2,3,5,8\}} F_k x_{k,i} - F_6 x_{6,i} \\
\frac{dN_{i,R}}{dt} &= F_6 x_{6,i} - F_7 x_{7,i} + \textstyle\sum_{j=1}^{4} \nu_{ij} r_j \\
\frac{dN_{i,S}}{dt} &= F_7 x_{7,i} - F_8 x_{8,i} - F_9 x_{9,i} - F_{10} x_{10,i} \\
\frac{dN_{i,C}}{dt} &= (1-\phi_i)(F_4 x_{4,i} + F_{10} x_{10,i}) - F_{11} x_{11,i}
\end{aligned} \tag{8.1}$$

where streams 6, 7, 8 and 9 carry the vapor compositions of the header, the reactor and the separator, 10 and 11 the liquid compositions of the separator and the stripper, and the stripper overhead is $F_5 x_{5,i} = \phi_i(F_4 x_{4,i} + F_{10} x_{10,i})$. The reactor and separator store A, B and C as vapor and D to H as liquid. The stripping factor is $\phi_i = \sigma_i \psi / (1 + \sigma_i \psi)$ with $\psi = (F_4/F_{10})\, g(T_C)$ and g a piecewise function of the stripper temperature, for D to H only, and A, B and C go overhead in fixed fractions. The energies follow the same streams, as shown in Equation (8.2).

$$\begin{aligned}
\frac{dE_V}{dt} &= \textstyle\sum_{k\in\{1,2,3,5,8\}} H_k F_k - H_6 F_6 \\
\frac{dE_R}{dt} &= H_6 F_6 - H_7 F_7 + \Delta H_1 r_1 + \Delta H_2 r_2 + Q_R \\
\frac{dE_S}{dt} &= H_7 F_7 - \textstyle\sum_{k\in\{8,9,10\}} H_k F_k + Q_S \\
\frac{dE_C}{dt} &= H_4 F_4 + H_{10} F_{10} - H_5 F_5 - H_{11} F_{11} + Q_C
\end{aligned} \tag{8.2}$$

where H_k is the molar enthalpy of stream k from cubic heat-capacity polynomials with a vaporization term for vapor, H_8 includes the compressor's work, Q_R and Q_S are the heats exchanged with the two cooling-water streams (negative in operation) and Q_C is the steam's. Only reactions 1 and 2 carry a heat of reaction. Each unit's temperature is found from its energy by Newton's method on the enthalpy polynomial, started from the previous temperature.

8.3.4 Pressure, Equilibrium and Reaction

The noncondensables obey the ideal gas law in each vessel's vapor space and the condensables Raoult's law with Antoine vapor pressures, as shown in Equation (8.3).

$$\begin{aligned} P_u &= \sum_{i\in\{A,B,C\}} \frac{N_{i,u} R T_u}{V_u - V_{L,u}} + \sum_{i\in\{D..H\}} x_{i,u}\, \mathrm{e}^{A_i + B_i/(T_u + C_i)} \\ P_V &= \frac{\left(\sum_i N_{i,V}\right) R T_V}{V_V}, \qquad V_{L,u} = \sum_{i\in\{D..H\}} \frac{N_{i,u}}{\rho_i(T_u)} \end{aligned} \tag{8.3}$$

where u is the reactor or the separator, T_u is in kelvin in the gas term, $R = 998.9\,\mathrm{mmHg\,ft^3/(lbmol\,K)}$ in the code's units, and the liquid molar densities ρ_i are quadratic in temperature. The four reactions run in the reactor's vapor volume, as shown in Equation (8.4).

$$\begin{aligned} r_1 &= \delta_1\,(V_R - V_{L,R})\,\mathrm{e}^{a_1 - E_1/(R_c T_R)}\, p_A^{n_A}\, p_C^{n_C}\, p_D \\ r_2 &= \delta_2\,(V_R - V_{L,R})\,\mathrm{e}^{a_2 - E_2/(R_c T_R)}\, p_A^{n_A}\, p_C^{n_C}\, p_E \\ r_3 &= (V_R - V_{L,R})\,\mathrm{e}^{a_3 - E_3/(R_c T_R)}\, p_A\, p_E \\ r_4 &= 0.767488334\,(V_R - V_{L,R})\,\mathrm{e}^{a_3 - E_3/(R_c T_R)}\, p_A\, p_D \end{aligned} \tag{8.4}$$

for A + C + D $\to$ G, A + C + E $\to$ H, A + E $\to$ F and 3D $\to$ 2F, with p_i the reactor partial pressures, $R_c = 1.987\,\mathrm{cal/(mol\,K)}$, and δ_1, δ_2 the kinetic factors that IDV 13 drifts. The stoichiometry gives $\nu_A = (-1,-1,-1,0)$, $\nu_C = (-1,-1,0,0)$, $\nu_D = (-1,0,0,-1.5)$, $\nu_E = (0,-1,-1,0)$, $\nu_F = (0,0,1,1)$, $\nu_G = (1,0,0,0)$ and $\nu_H = (0,1,0,0)$, and B is inert.

8.3.5 Flows, Heat Transfer and Valves

Pressure differences drive the gas between the vessels, and the compressor follows a curve in its pressure ratio less what its recycle valve returns, as shown in Equation (8.5).

$$\begin{aligned} M_6 F_6 &= 1937.6\sqrt{P_V - P_R} \\ M_7 F_7 &= 4574.21\sqrt{P_R - P_S}\,(1 - 0.25\,\delta_{20}) \\ M_8 F_8 &= F_{max}\left[1 + (1-\rho^3)/1.197\right] \\ &\quad - 53.349\,v_5\sqrt{P_V - P_S} \\ M_9 F_9 &= 0.151169\,v_6\sqrt{P_S - 760} \\ F_k &= v_{j(k)}\,F_{k,max}/100 \quad \text{(feeds, liquid draws)} \end{aligned} \tag{8.5}$$

where M_k is the stream's mean molar mass, pressures are absolute in mmHg, each square root is of a difference clipped at zero, $\rho = \min(1.3, \max(1, P_V/P_S))$ is the compressor's pressure ratio, δ_{20} is the flow walk of IDV 20, the A feed is cut to zero by IDV 6 and the A and C feed to 80 % by IDV 7. The heat removed from the reactor and the condenser and the steam's heat are

$$\begin{aligned} Q_R &= UA_R(V_{L,R}, a)\,(T_{w,R} - T_R) \\ Q_S &= UA_S(F_7)\,(T_{w,S} - T_R) \\ Q_C &= UA_C\,v_9\,\max(0,\, 100 - T_C) \end{aligned} \tag{8.6}$$

where UA_R rises linearly with the reactor liquid volume from 78 to 390 $\mathrm{ft^3}$ and is quadratic in the agitator speed $a = (v_{12} + 150)/100$, and UA_S saturates in the effluent flow. Each cooling-water outlet temperature is a state that balances the water's flow, set by v_{10} or v_{11}, against the heat it takes, and IDV 4, 5, 11, 12, 16, 17 and 18 move the inlet temperatures or the three heat-transfer coefficients. Every valve moves toward its command through a first-order lag, $\tau_j\,\mathrm{d}v_j/\mathrm{d}t = c_j - v_j$, with τ_j from 5 s to 9 s and 120 s for the steam valve, and under IDV 14, 15 and 19 a valve keeps its last command until the new one differs by more than 2 %.

Table 8.2 Model Constants. H: Hidden, Shown Read-Only in the Model Dialog. A: Adjustable in the Panel

Symbol	Meaning	Value	Mark
V_R, V_S, V_V	reactor, separator, header volumes (ft^3)	1300, 3500, 5000	H
V_C	stripper level span (ft^3)	156.5	H
a_1, a_2, a_3	pre-exponentials (ln)	31.5859536, 3.00094014, 53.4060443	H
E_1, E_2, E_3	activation energies (cal/mol)	40000, 20000, 60000	H
n_A, n_C	orders in p_A, p_C	1.1544, 0.3735	H
ΔH_1, ΔH_2	heats of reaction (code units)	0.06899381054, 0.05	H
F_{max}	compressor maximum flow (lb/h)	280275	H
ρ_{max}	compressor maximum ratio	1.3	H
$\tau_1 \ldots \tau_{12}$	valve lags (s)	8, 8, 6, 9, 7, 5, 5, 5, 120, 5, 5, 5	H
T_a	analyzer periods and dead times (h)	0.1, 0.1, 0.25	H
limits	P_R, T_R, liquid volumes	see text	H
v_j	valve stations (%)	base case	A
IDV	twenty disturbances	off	A
Δt	Euler step (s)	1	fixed

8.3.6 Measurements, Disturbances, Shutdown and Cost

The 22 continuous measurements carry additive noise of the published standard deviations σ_k, each draw the sum of twelve uniform deviates less six, times σ_k. The reactor feed and purge analyzers sample every 0.1 h and report what they sampled one period earlier, and the product analyzer does the same at 0.25 h, as shown in Equation (8.7).

$$\begin{aligned} y_k(t) &= h_k(\mathbf{x}(t)) + \sigma_k\Big(\textstyle\sum_{m=1}^{12} u_m - 6\Big), \quad k \le 22 \\ y_k(t) &= h_k\big(\mathbf{x}(t_s - T_a)\big) + \text{noise}, \quad t_s \le t < t_s + T_a \end{aligned} \tag{8.7}$$

where t_s is the last sample time and T_a the analyzer's period. The random disturbances are cubic splines whose knots are drawn at random intervals, and the step disturbances shift a feed composition, temperature or availability. The plant shuts itself down, freezing every state as the code does, when P_R exceeds 3000 kPa g, T_R exceeds 175 °C, or a liquid volume leaves its band: reactor 2 to 24 m^3, separator 1 to 12 m^3, stripper 1 to 8 m^3. The operating cost is the published function as Ricker's archived Simulink model codes it, as shown in Equation (8.8).

$$\begin{aligned} J = {} & 0.0536\,W + 0.0318\,S + 0.44791\,F_9 \textstyle\sum_i c_i\, y_{9,i} \\ & + \tfrac{211.3}{46.534}\, v_8\,(0.2206\,x_D + 0.1456\,x_E + 0.1789\,x_F) \end{aligned} \tag{8.8}$$

where W is the compressor work in kW (XMEAS 20), S the steam in kg/h (XMEAS 19), F_9 the purge in kscmh with its analyzed composition $y_{9,i}$ priced at c_i (2.209, 0, 6.177, 22.06, 14.56, 17.89, 30.44 and 22.94 for A to H), and the last term the D, E and F lost in the product, with the product rate taken from the stripper valve as 211.3 kmol/h at 46.534 % open. Ricker's Simulink block rounds that ratio to 4.541, which changes the cost of the tabulated base case by 0.0016 $/h.

8.3.7 What the Port Changes, and Its Numerics

Three deliberate changes lie outside the plant's arithmetic. The random numbers come from two seeded generators of the framework (Section 1.6.1), one for the disturbance walks and one for the measurement noise, as in the 2015 revision, instead of the single linear congruential generator of the code, so a run replays bit for bit after Reset while its noise realization differs from the original's. The analyzers fire on a count of steps (360 and 900 one-second steps) instead of comparing a floating-point time, and the state lives in one object per plant. The constants are the decimal values typed in the code, read in double precision as the Python port reads them, which is the subject of Section 8.4. The integrator is the model's own explicit Euler step of 1 s, as the driver distributed with the code uses, because TEFUNC advances the disturbance walks and the analyzers when it is called and therefore

cannot be placed inside a multistage integrator. A 48 h run of the challenge, controllers and logging included, took about 1.3 s headless in Node on the development host. In the application, at the default speed of 0.25 h per second the open-loop shutdown at 3.59 h arrives about 14 s after Run, and at the top speed of 4 h per second the two-day challenge plays in about 12 s. The constants with their marks are in Table 8.2.

8.3.8 Assumptions

The plant equations, constants, noise levels, analyzer periods and limits are those of the public code and are the benchmark's teaching values, not data of any operating plant, whose relief and interlock settings would come from its vessel design code and its safety instrumented system study under IEC 61511. The phases are in equilibrium at each vessel's temperature, the stripper separates by fixed factors rather than by stages, and the compressor and the flows between vessels are algebraic, so pressure waves and transport delays are missing, which makes fast transients the least trustworthy part of the model. The noise statistics are the published ones while the realizations are not. Disturbances 16 to 20, which the code implements although its own comments list them as unknown, are kept, and the revision's disturbances 21 to 28 and its extra measurements are not ported.

8.4 How Closely the Port Follows the Code

8.4.1 The Base Case

At $t = 0$ the 41 measurements agree with the base case tabulated in Ricker's archive to within 0.96 of half the last printed digit (worst at XMEAS 1), and the 12 valve positions to within 0.98 (XMV 11). Every derivative is at most 9.31×10^{-4} per hour with the constants in double precision, inside the 10^{-3} criterion that Ricker's archive states for a steady state, and 2.54×10^{-3} with the single-precision reading. This is why the port reads the constants in double precision.

8.4.2 Against the Python Port

For the point-by-point comparison the port runs in a validation mode that reproduces the Python port's choices: the original generator in one shared stream with exact integer arithmetic, its Newton starting temperatures, a floating-point analyzer clock and its clipping of the valve commands. The Python port was run once, in an isolated environment, with its trajectories recorded every minute, and every comparison below reads those records. Under IDV 1, 2, 4, 6, 8 and 13 in open loop, the analyzer readings differ by at most 1.76×10^{-11} mol % and the states by 9.57×10^{-12} relative, and with four PI loops closed for 6 h under nine disturbance sets the states differ by at most 1.59×10^{-11}, as shown in Figure 8.2b. The undisturbed open-loop run reaches 2.05×10^{-8}, because differences in the last bit grow along the unstable mode. Every shutdown comes at the same step, at 3.5856, 0.4783, 1.5992, 1.0564, 0.4422, 1.8878 and 2.0261 h in open loop (undisturbed, then IDV 1, 2, 4, 6, 8 and 13) and 2.8894 h closed under IDV 1.

8.4.3 Against an Independent C Translation

Ricker's f2c translation shares no code with the Python port. With the constants read as f2c reads them (the three ratios E_j/R_c folded in double precision and then rounded to single, the two exponents kept in double, the rest in single), the port agrees with it bit for bit at $t = 0$ in all 50 derivatives and 41 measurements, and to 3.98×10^{-10} relative over 6 h of open loop, with the shutdown at the same step, second 12485. Under IDV 1, 2, 4, 6 and 7, which are deterministic, the agreement over runs of 1 to 2 h is 3.95×10^{-12} or better. The shipped double-precision reading differs from the C code by 1.71×10^{-6} in the measurements at $t = 0$, which is the size of the rounding, not of an error.

8.4.4 An Open-Loop Trip Set by Rounding

The undisturbed open-loop shutdown is the plant's instability acting on that residue, so its time belongs to the reading of the constants, as shown in Figure 8.2a. On the same 1 s Euler step the trip comes at 3.47 h with the C translation's reading, 3.59 h in double precision and 4.06 h in single

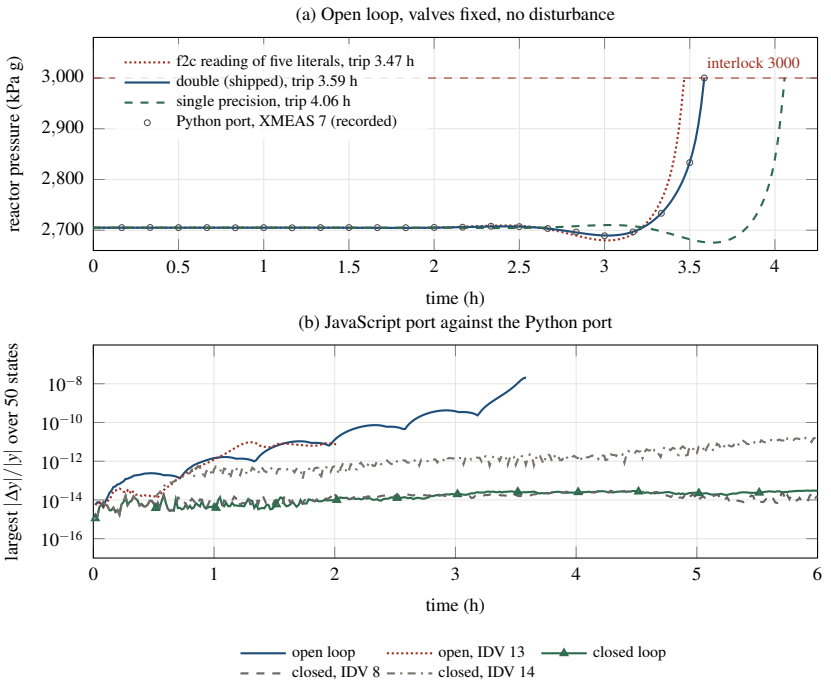


Figure 8.2 The port against its references. (a) Open-loop reactor pressure for three readings of the same decimal constants, the f2c reading dotted, the double-precision reading solid and the single-precision reading dashed, with the Python port's recorded measurement every 10 min as open circles and the interlock as a thin dashed line. (b) The largest relative state difference from the Python port, minute by minute, for two open-loop and three closed-loop runs: open loop solid, open loop under IDV 13 dotted, closed loop solid with triangles, and closed loop under IDV 8 dashed and under IDV 14 dash-dotted.

precision, always on high reactor pressure. The integration matters far less: half and a quarter of the step give 3.5863 and 3.5867 h, and the classical fourth-order method at 1 s gives 3.5872 h. The README of Ricker's Simulink example reports the same shutdown after about 2 h of simulated time. A trip time quoted for this plant therefore identifies an implementation as much as a dynamic.

🕮 **Pause.** Try Exercise 8.1 before reading on.

8.4.5 The Base-Case Cost

The port's cost at $t = 0$ is 170.62 \$/h, and the same function evaluated on Ricker's tabulated base case gives 170.83 \$/h. By term the table exceeds the port by -0.023 for power, -0.010 for steam, -0.045 for the purge and $+0.285$ \$/h for the product losses. Almost all of the gap is the table's two-decimal product composition (D, E and F printed as 0.02, 0.84 and 0.10 mol % against the port's 0.0179, 0.8357 and 0.0986), which alone accounts for 0.28 \$/h. Within its printed digits the table's cost can lie anywhere from 168.10 to 173.57 \$/h, and the port's value is inside. The 1993 paper was not available when the port was checked, so no printed base-case cost is quoted from it.

8.4.6 Python and Fortran Differences

Eight differences between the Python port and the Fortran were found while translating, none in the balances, rates, equilibrium, flows, cost or shutdown tests. They concern the generator's arithmetic, the precision of untyped literals, the Newton starting temperatures, where the valve commands are clipped, evaluation after a shutdown, guarded divisions and the driver's single-precision step. The

Table 8.3 The Seventeen PI Loops of the Archived Structure. K_c Acts on SP − PV, So a Negative Gain Raises the Output With the Measurement

Measured	Output	K_c	τ_I (min)	Limits
production, XMEAS 17	trim on F_p (%)	3.2	120	±30
P_R, XMEAS 7	purge ratio r_5	-10^{-4}	20	0, 100
L_R, XMEAS 8	separator T setpoint	0.8	60	0, 120
T_R, XMEAS 9	XMV 10	−8	7.5	0, 100
T_S, XMEAS 11	XMV 11	−4	15	0, 100
L_S, XMEAS 12	liquid ratio r_6	-10^{-3}	200	0, 100
L_C, XMEAS 15	product ratio r_7	-2×10^{-4}	200	0, 100
x_G, XMEAS 40	D/E shift E_{adj}	−0.4	100	none
y_A, XMEAS 23, 25	A trim u_A	2×10^{-4}	60	±1
y_{AC}, XMEAS 23, 25	A + C trim u_{AC}	3×10^{-4}	120	±1
seven flows	XMV 3, 1, 2, 4, 6, 7, 8	see text	0.001	0, 100

port follows the Fortran in each except the literals and the step, and its validation mode follows the Python.

8.5 Control Structures as Implemented

The controllers execute at every 1 s step and read only the 41 measurements, including the analyzers' late reports. The header's Manual gives all twelve valves to their stations. Auto runs the structure chosen in the panel, and a valve the structure does not drive stays on its station. Loops outside the running structure track the plant, so a change of structure or a transfer to Auto moves no valve. Every PI loop is the velocity-form PID of Section 1.6.2, with the tracking time of its back-calculation set to $T_t = \max(\tau_I, \Delta t)$ because Ricker's flow loops have integral times of 0.06 s, below the step. His discrete PI blocks are carried with his gains and integral times, his sign convention mapped to the loop's action, as listed in Table 8.3.

8.5.1 Inventory Loops

The inventory structure closes the reactor pressure on the purge, the reactor level on the separator temperature setpoint, the separator temperature on the condenser cooling water, the separator and stripper levels on their liquid draws, and the reactor temperature on its cooling water, with the four feeds, the recycle, the steam and the agitator on their stations. The reactor temperature loop belongs in this layer because without it the undisturbed plant trips on high pressure at 8.12 h.

8.5.2 The Decentralized Structure

The decentralized structure follows Ricker's archived Simulink model `MultiLoop_mode1.mdl`, run at the base-case setpoints. A production index sets every flow setpoint as a ratio times the index, as shown in Equation (8.9). The index is the plant's throughput manipulator in the sense of Section 1.4.3, and it moves every flow at once.

$$\begin{aligned} F_p &= \tfrac{100}{22.89}\,F_{SP} + \mathrm{PI}_F(F_{SP} - F_{11}) \\ F_{k,SP} &= r_k F_p, \qquad k = 1\ldots7 \\ r_2 &= p_D(x_{G,SP}) - 32\,E_{adj}/F_p + b_D \\ r_3 &= p_E(x_{G,SP}) + 46\,E_{adj}/F_p + b_E \\ r_1 &= \bar{r}_1 + u_A, \qquad r_4 = \bar{r}_4 - u_A + u_{AC} \end{aligned} \tag{8.9}$$

where F_{SP} is the production setpoint after a rate limit of 30 % of the base rate per day, p_D and p_E are Ricker's quadratic feedforward polynomials in the G setpoint (itself rate-limited to 50 mol % per day), E_{adj} is the output of the G loop, u_A and u_{AC} are the outputs of the two feed analyzer trims, and r_5 to r_7 are the outputs of the pressure, separator level and stripper level loops. The biases b_D,

b_E, $\bar{r}_1$ and $\bar{r}_4$ are set when a loop enters service so that each feed setpoint equals the feed the meter reads. The flow loops have gains of 0.01, 1.6×10^{-6}, 1.8×10^{-6}, 0.003, 0.01, 4×10^{-4} and 4×10^{-4} percent of valve per unit of their flows (kscmh for gases, kg/h for D and E, m^3/h for liquids).

In this archived model the production index, with the production loop's trim, scales all seven flow loops, the product draw among them. Ricker's archived recordings of his 1996 strategy show a different formulation. In the run of IDV 6 at 50 h his product setpoint is 20.877, equal to the product ratio times the index less the production trim, and not 21.143, the ratio times the full index, so his trim moved only the four feeds, and the same record shows his pressure override taking 8.71 points off the index. The application keeps the archived formulation because it is the one his code states. A one-element change toward the 1996 formulation, tried during the review of the application, neither reproduced his IDV 6 recording nor kept the suggested challenge running (it tripped at 5.88 h).

8.5.3 A Pressure Override on the Production Index

The archived model carries no override, so the application adds one of its own design. Like the surge override of Volume 4, Chapter 2, it is idle in normal operation and acts only near a limit. The reactor pressure passes through a first-order filter of 0.02 h, and while the filtered pressure P_f exceeds the threshold P_{ovr}, each step takes a cut from the index and writes the index applied back into the production loop's output, as shown in Equation (8.10).

$$\begin{aligned} P_{f,k} &= P_{f,k-1} + (1-\mathrm{e}^{-\Delta t/0.02})(P_{R,k}-P_{f,k-1}) \\ \delta_k &= K_{ovr}\,(P_{f,k}-P_{ovr})\,\Delta t_s, \qquad F_p \leftarrow F_p - \delta_k \\ u_{F,k} &\leftarrow \max(1,F_p) - \tfrac{100}{22.89}\,F_{SP} \end{aligned} \tag{8.10}$$

where Δt_s is the control period in seconds (1 at the shipped step), $K_{ovr}=2$ percent of F_p per kPa per second and $P_{ovr}=2950\,\mathrm{kPa}\,g$. Because each cut stays in the production loop's output, the override integrates until that output reaches its lower limit of -30 points, and beyond it the cut is in proportion to the present excess. With the production loop frozen at the override's first step in case 8 at 17 m^3/h, a filtered excess of at most 1.11 kPa took F_p from 83.8 to 42.1 % in 3 min. Over the last 4 h of that case the production loop sat at its limit 98 % of the time and the override acted 58 % of the time, with the filtered pressure on average 9.85 kPa above its threshold while it did. The measured pressure noise is 0.298 kPa rms against the code's 0.3, and the filter is there because an unfiltered override, tried while the application was built, cut production on that noise alone. In all ten cases as shipped (case 1 in Auto, the challenge on its suggested settings) at three seeds, the override acts only in case 8.

8.5.4 Model Predictive Control Above the Loops

In the predictive structure the production and mix loops stand down and a receding-horizon controller sets their two handles, the production target u_1 (m^3/h) that sets $F_p=(100/22.89)u_1$ and the G target u_2 (mol %) that feeds the polynomials of Equation (8.9). The program is the one of Section 1.6.3 with this plant's model and weights. Its model, identified by steps on this plant as Volume 1, Chapter 12 teaches, is two first-order responses, as shown in Equation (8.11).

$$\begin{aligned} 0.1\,\dot{y}_1 &= u_1 - y_1, \qquad 1.4\,\dot{z} = u_2 - z \\ y_2(t) &= z(t-0.65) \\ \min\ & \textstyle\sum_{k=1}^{80}\left[w_F(y_{1,k}-r_1)^2 + w_G(y_{2,k}-r_2)^2\right] \\ & \quad + \textstyle\sum_j\left[w_{\Delta u}\Delta u_{1,j}^2 + 0.1\,w_{\Delta u}\Delta u_{2,j}^2\right] \end{aligned} \tag{8.11}$$

in hours, discretized for a zero-order hold every 0.05 h over 80 steps (4 h) in 12 blocks, with a terminal weight five times the stage weights, $w_F=10$, $w_G=1$, $w_{\Delta u}=5$, bounds of 15 to 32 m^3/h and 5 to 95 mol %, and moves limited by the two ramp limits. The cross effects are left to output biases corrected toward the measurements with a gain of 0.3 at every decision. Each program is handed

Table 8.4 The Ten Cases

#	Case and events	Teaching point
1	Manual, every valve fixed, no disturbance	the open-loop shutdown
2	Inventory loops only, IDV 1 at 2 h	inventories first
3	Production −15 % over 2 h from 2 h	one production handle
4	IDV 1 at 1 h	analyzers that report late
5	IDV 8 throughout	the cost of noise and of pressure
6	IDV 13 at 1 h, 48 h	slow drift
7	IDV 14 at 1 h	a sticking valve
8	IDV 6 at 1 h	the override, a structure's limit
9	G setpoint to 11.7 mol % at 1 h	mix change, 50/50 to 10/90
10	Challenge: IDV 8, IDV 1 at 2, production +10 % at 24 h, inventory loops at 2705 kPa g as shipped	the whole problem

to the dual active-set QP solver of Goldfarb and Idnani [16] that the series shares, the first move is applied as industrial practice does [17], and a plan the solver rejects, or one with a non-finite value or outside its bounds, returns both handles to the PI loops without a bump.

8.5.5 Three Models and the Truth Boundary

The plant is the ported code. The identified model is the pair of responses in Equation (8.11), and the controllers carry no other model. The scorecard, the challenge verdict and the comma-separated export columns prefixed `truth_` use the noise-free, delay-free values, which no controller reads. The export also carries the 41 measurements, the twelve valve commands and positions and the controller states.

8.6 Ten Cases, Measured

The ten cases are listed in Table 8.4. Each starts from the base case on the decentralized structure with its override unless its row says otherwise. Every number below was measured on the engine headless at the default seed, with settings entered before Run, which the engine treats as starting conditions (Section 1.6.1). Each case logs a bit-identical run whether it is opened first, after another case, or again through Reset.

8.6.1 Inventories Before Everything

Case 1 is the run of the Introduction. The plant holds its published state for two hours with every valve fixed, then the pressure reaches the interlock at 3.59 h with the reactor at 96.6 °C. In Auto the same plant holds the pressure between 2702 and 2708 kPa g for 48 h. Case 2 runs the inventory loops alone and steps the A and C feed's composition at 2 h. The analyzer's A share of A plus C in the reactor feed falls from 55.0 to 37.0 %, the unreacted C accumulates in the recycle, and the pressure loop opens the purge as far as it helps. The pressure passes 2900 kPa g at 4.84 h and the interlock at 5.84 h. With the decentralized structure the A share never falls below 45.4 %, the pressure peaks at 2751 kPa g, and the plant runs on. The inventory loops make the plant stable, and only the composition loops above them make it survive a feed change.

Table 8.5 Measured Outcomes at the Default Seed

#	Setting	Outcome
1	as shipped (Manual)	trip, pressure, 3.59 h
	Auto	2702–2708 kPa g, 48 h
2	inventory loops	trip, pressure, 5.84 h
	decentralized	peak 2751 kPa g, no trip
3	as shipped	19.29 m^3/h at 12 h, stripper 28.7–81.1 %
	stripper level gain 0.08	stripper 42.3–63.6 %, production off by 0.55
4	as shipped	A share low 45.29 %, peak 2752 kPa g
	trims at zero gain	A share low 22.94 %, peak 2981 kPa g
5	at 2705 kPa g, 24 h	mean 176.1 $/h (170.7 undisturbed)
	at 2800 kPa g, 24 h	mean 146.3 $/h, peak 2827 kPa g
6	decentralized, 48 h	2566–2854 kPa g, mean 209.1 $/h
	inventory loops, 48 h	2543–2847 kPa g, mean 212.3 $/h
7	$K_c = 8$ (Ricker's)	±0.31 °C, period 4.8 min
	$K_c = 4$	±0.57 °C, period 8.1 min
	$K_c = 16$	±0.23 °C, period 3.2 min
8	as shipped	trip, stripper level low, 8.40 h
	17 m^3/h, override	no trip in 24 h, 5.88 m^3/h last 4 h
	17 m^3/h, no override	trip, pressure, 13.34 h
	inventory loops	no trip, 21.94 m^3/h, 408 $/h
9	ramp 2.08 mol %/h	within 2 points at 19.8 h, low 7.6 %
	ramp 10 mol %/h	within 2 points at 5.7 h, low 4.6 %
	MPC	within 2 points at 21.0 h, low 9.2 %

TRY IT IN THE APP 1 · Every Valve Fixed: The Open-Loop Shutdown

Load the scenario, leave the header on Manual, press Run and watch the reactor pressure. For two hours nothing seems to happen. Then the pressure climbs to the 3000 kPa g interlock at 3.59 h, and the plant stops. Switch the header to Auto and press Reset: the inventory loops now hold the pressure between 2702 and 2708 kPa g.

8.6.2 One Handle for Production

Case 3 lowers the production setpoint 15 %, to 19.51 m^3/h, ramped at 1.720 m^3/h per hour from 2 h. Because the index multiplies every flow setpoint, the ramp reaches all seven flow loops at once, and the production comes within 0.3 m^3/h of the new setpoint at 3.95 h. By 12 h the index reads 84.6 % and the production 19.29 m^3/h. The stripper level takes up the mismatch between what the reactor still makes and what the product valve draws, swinging from 28.7 to 81.1 %, because its loop is slow on purpose. At four times Ricker's gain the level stays between 42.3 and 63.6 %, but the production strays up to 0.55 m^3/h from its setpoint in force instead of 0.32, so a buffer held tight passes the disturbance on. This is the trade between a level and the flow it passes that Section 1.4.8 describes.

Pause. Try Exercise 8.3 before reading on.

8.6.3 Late Analyzers

In case 4 the A and C feed changes composition at 1 h, and the only measurement of the change is the reactor feed analyzer, reporting every 0.1 h what it sampled 0.1 h earlier. As shipped the A share bottoms at 45.29 % at 3.30 h and returns to 56.89 % by 12 h, with the A feed raised from 0.251 to 0.595 kscmh and the pressure peaking at 2752 kPa g. With both analyzer trims at zero gain the share falls to 22.94 % and the pressure reaches 2981 kPa g within 12 h. In case 9 the G setpoint moves from the 50/50 mode to the 10/90 mode of Ricker's table, 11.66 mol % there and 11.7 mol % in the application. At Ricker's ramp of 2.08 mol % per hour, G reaches within two points of 10 % by mass at 19.8 h and dips to 7.6 %. At 10 mol % per hour it arrives at 5.7 h and undershoots to 4.6 %, because the product analyzer reports a quarter of an hour late. The predictive controller, limited to the same ramp, arrives at 21.0 h with a lowest G of 9.2 %.

8.6.4 Noise, Pressure and Cost

Case 5 runs the random feed composition of IDV 8 for a day. The mean cost is 176.1 $/h against 170.7 undisturbed, so the noise costs about 5 $/h, while raising the pressure setpoint to 2800 kPa g lowers the mean to 146.3 $/h, because a higher pressure needs less purge to carry out the same inerts and loses less A, C, D and E with it. The price is margin, a peak of 2827 kPa g against 2756. The cost from minute to minute swings far wider than these differences, from 57 to 340 $/h between its 5th and 95th percentiles at 2705 kPa g, which is why the scorecard shows the mean. Moving such a setpoint toward its constraint is the work of an optimizer above the loops, as in Volume 13, Chapter 11. In case 6 the slow kinetic drift of IDV 13 runs 48 h without a trip, the pressure wandering over 287 kPa (2566 to 2854 kPa g) with the purge valve between fully shut and fully open, at a mean of 209.1 $/h. The inventory loops alone also hold the two days, at 212.3 $/h, because this drift is the pressure loop's to carry. In case 7 the reactor cooling valve sticks within a 2 % band, and with Ricker's gain the temperature cycles between 120.09 and 120.71 °C with a period of 4.8 min and the stem jumping between 38.7 and 43.4 %, against 120.39 to 120.41 °C without stiction. Halving the gain lengthens and enlarges the cycle and doubling it shortens and shrinks it, and none removes it, while the pressure stays between 2702 and 2708 kPa g, far from the override. That is why the override has its own case.

Pause. Try Exercise 8.2 before reading on.

> **TRY IT IN THE APP** **7 · Sticking Reactor Cooling Valve (IDV 14)**
>
> Run the scenario and watch the reactor temperature after the cooling water valve starts to stick at 1 h. With Ricker's gain of 8 it cycles by ±0.31 °C with a period of 4.8 min. Set the Reactor temperature gain to 4, press Reset and run again: the cycle grows to ±0.57 °C and slows to 8.1 min, and no gain removes it.

8.6.5 The A Feed Lost, and Ricker's Recording

Case 8 loses the A feed at 1 h. A still arrives with the A and C feed, whose 4.534 kscmh of A is 94.8 % of the A fed at the base case, but what is lost is the trim, since A now comes only in a fixed ratio to C. As shipped the structure trips on the stripper level at 8.40 h, and the override acts for only 2.0 min before it. The production loop chasing a rate that the reactor can no longer quite make scales the product draw with every feed and pulls the stripper empty. With the production setpoint cut to 17 m^3/h the override takes production off the index each time the filtered pressure passes 2950 kPa g, the pressure peaks at 2974 kPa g, and the plant settles at 5.88 m^3/h over its last four hours, a quarter of the base rate, alive. Without the override the same plant trips on high pressure at 13.34 h.

> **DID YOU KNOW?** **Fewer Loops, No Trip**
>
> From the simulator (Table 8.5). When the A feed is lost, the full decentralized structure trips on a low stripper level at 8.40 h. The inventory loops alone, with no production or composition loop above them, ride the same loss without a trip, at 21.94 m^3/h and 408 $/h.

The loss is not that deep for the plant. The inventory loops alone ride it at 21.94 m^3/h and 408 $/h, and Ricker's archive holds a recording of his 1996 strategy riding it at the base setpoint, ending at 20.82 m^3/h after 50 h with the pressure held near 2950 kPa g by his own override, as listed in Table 8.6. Under IDV 1, 8 and 13 the two structures agree within 68 kPa on the pressure extremes, 0.30 m^3/h on production and 0.81 mol % on G, and the stripper level is where they part, swinging two to three times as far here. Ricker's 1996 paper, whose own production-rate responses could not be consulted when the port was checked, is cited for the strategy [8], and the comparison rests on the recordings alone. Case 8 therefore teaches a property of this structure and not of the plant: a production loop that scales the product draw can empty a vessel that a gentler formulation would protect.

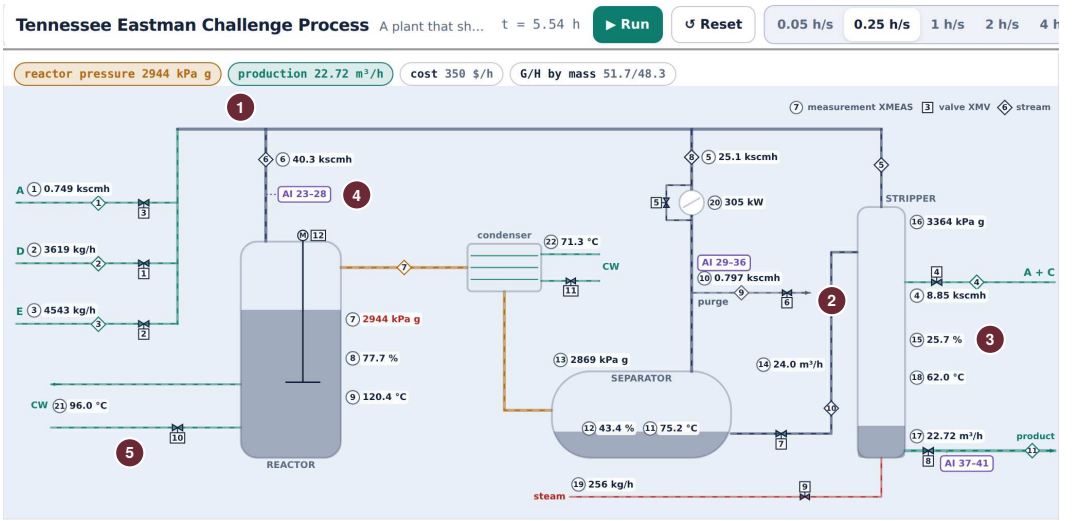


Figure 8.3 The suggested challenge at $t = 5.54$ h on the Plant view, decentralized structure with the override, reactor pressure setpoint 2800 kPa g, IDV 8 on and IDV 1 stepped at 2 h. The picture is a composite of the application's header above its Plant view, with the control panel, the trends and the Learn panel cut. The reactor pressure is at its peak of 2944 kPa g, 56 kPa below the interlock. The flowsheet carries every continuous measurement at its instrument, among them the A feed at 0.749 kscmh, the separator at 2869 kPa g, the stripper level at 25.7 % and the production at 22.72 m^3/h. Callout 1 marks the badge row of reactor pressure, production, operating cost and G/H split, 2 the purge valve (XMV 6) that the pressure loop moves, 3 the stripper level that case 8 pulls empty, 4 the reactor feed analyzer that reports late, and 5 the reactor cooling water valve (XMV 10) that sticks in case 7.

Table 8.6 Extremes Over 50 h, Disturbance at 1 h, Base Case: Ricker's Archived Recordings of His 1996 Strategy (R) and This Structure (A)

IDV		P_R (kPa g)	F_{11} (m^3/h)	L_C (%)	x_G (mol %)
1	R	2634–2747	22.25–23.37	36.0–65.8	51.97–55.78
	A	2599–2751	22.47–23.66	22.3–90.2	51.82–55.74
8	R	2617–2745	22.31–23.46	31.7–66.9	51.38–55.63
	A	2549–2783	22.09–23.77	−3.4–94.9	51.59–56.12
13	R	2532–2787	22.18–23.45	31.5–65.2	51.79–55.82
	A	2567–2854	22.41–23.63	19.9–91.8	50.98–55.82
6	R	2704–2961	20.55–23.21	40.3–56.4	52.32–55.11
	A	trip, stripper level low, at 8.40 h			

8.6.6 Infeasible at Maximum Production

An honestly infeasible case was also sought, the A feed lost at maximum production. At the base rate it is not infeasible, as the inventory loops show, so the claim is made at Ricker's tabulated maximum-production point, 36.04 m^3/h at 50/50 G/H. There every setting tried trips within a day: the decentralized structure with its override at four pairs of threshold and gain, without it, the inventory loops, the predictive controller, and production cuts to 8, 12, 17 or 18 m^3/h applied at the loss or earlier, sixteen settings in all, with trips between 5.08 and 18.92 h. Ricker's own strategy was not tested at that point, so the case says infeasible here, not infeasible for every structure.

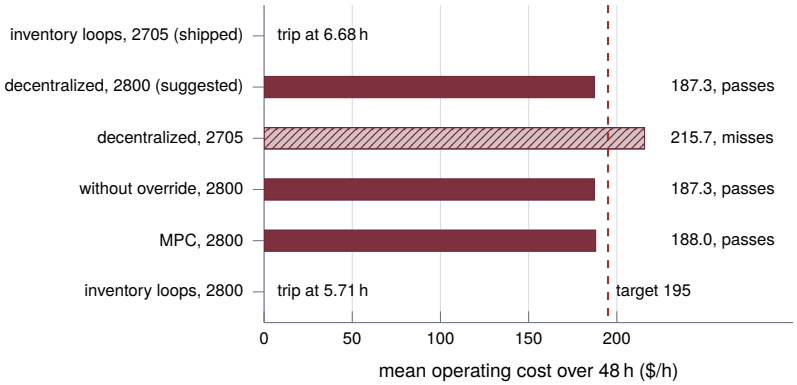


Figure 8.4 The challenge of Table 8.7 at the default seed, settings by reactor pressure setpoint in kPa g: mean cost of the runs that last 48 h (dark bars pass, the hatched bar misses), the target as a dashed line, and the trip times of the two runs on the inventory loops alone. Notice that the inventory loops trip at either pressure, and that the decentralized structure meets the cost target only when its pressure setpoint moves toward the interlock.

Table 8.7 The Challenge Over 48 h, Default Seed and Seeds 1 to 5

Settings	Default seed	Seeds 1–5
inventory loops, 2705 kPa g (shipped)	trip 6.68 h	trip 4.93–5.63 h
decentralized, 2800 kPa g (suggested)	passes, 187.3 $/h, G 2.33, 99.1 %	all pass, 156.7–179.0 $/h
decentralized, 2705 kPa g	misses, 215.7 $/h	4 miss (205.8–211.2), seed 1 passes at 186.9
without override, 2800 kPa g	passes, 187.3 $/h (identical)	—
MPC, 2800 kPa g	passes, 188.0 $/h	—
inventory loops, 2800 kPa g	trip 5.71 h	—

8.6.7 The Challenge, Redesigned

A first design of the challenge shipped the decentralized structure with its override removed and a detuned production loop. On the decentralized structure at the base pressure its disturbances raise the pressure no higher than 2899 kPa g, 51 kPa short of the override's threshold, and the override never acts, so the override was not what it tested. The challenge as built runs 48 h with IDV 8 throughout, IDV 1 at 2 h and production 10 % higher at 24 h, judged on true values: no shutdown, G within five points of 50 % by mass from 4 h, mean production at least 98 % of its setpoint, and a mean cost of at most 195 $/h. An attempt counts only with the model and the disturbances as given. It ships on the inventory loops at 2705 kPa g, and the suggested settings are the decentralized structure at 2800 kPa g. The results at six seeds are in Table 8.7, and Figure 8.4 draws the default seed.

An extended check separates instrument noise from the random feed path. With the challenge's physical IDV 8 walk fixed, the default realization and 30 additional measurement-noise streams all meet the target at both the one-second and half-second integration steps. Mean cost spans 187.08–187.88 $/h at one second and 168.42–169.94 $/h at half a second, while production remains above 99.06 % of its setpoint. The refined test keeps analyzer and disturbance clocks on physical time. Varying the physical walk as well is a different experiment: 30 of 31 days pass at each step, while seed 27 shuts down on low stripper inventory at 33.91 h with the native step and 33.82 h with the refined step. Thus the suggestion is reproducible for the fixed teaching day, including its instrument noise, but is not a robust operating policy for every feed trajectory. These wider results do not relax a shutdown or product target.

The inventory loops cannot follow the A/C step, as in case 2, and trip at 6.68 h. The decentralized structure keeps the plant running and the mix inside its band at either pressure, but at the base pressure

its mean cost of 215.7 $/h exceeds the target, while at 2800 kPa g the purge losses fall to a mean of 187.3 $/h with the worst G 2.33 points from 50 and the production at 99.1 % of its setpoint. The price is margin. After the A/C step the pressure peaks at 2944 kPa g at 5.54 h, as shown in Figure 8.3, and the straight-line estimate on the status strip reaches the interlock within the hour for 43.6 min, from 4.72 to 5.45 h. The override never acts in this run, so the structure without it scores identically. With half the integration step the shipped run still trips at 6.68 h and the suggested one still passes, at 169.3 $/h, the change coming from twice as many noise samples reaching the loops.

8.7 Technology in Practice

What Measures It

The inventory loops of this plant read pressures, levels, temperatures and flows, and the composition loops read three sampled analyzers. In the model the reactor feed and purge analyzers report every 0.1 h what they sampled one period earlier, and the product analyzer does the same every 0.25 h. In a plant such analyzers are process chromatographs, whose separation by time makes some measurement dead time inevitable, and whose sample line would delay the measurement grossly without a continuous flow to waste [18]. The 22 continuous measurements of the model carry only additive noise of the published standard deviations, so no measurement of the model ever fails or drifts.

> **TECHNOLOGY Sensor: The Process Chromatograph and Its Sample Line**
>
> The compositions of a plant like this one come from process chromatographs, which separate the species of a sample by the time each takes to pass through a column. That separation is why cycle times of minutes are common even on line, and some dead time is inevitable. Enough of it makes a feedback loop oscillate on its own. The sample must also reach the analyzer: a continuous flow from the process to waste keeps a fresh sample at the instrument, and without it the gas would take a very long time to travel the line. The sample conditioning, with its filtering, heating or cooling and drying, is one of many points of failure, and keeping analyzers working is treated as an advanced skill [18].

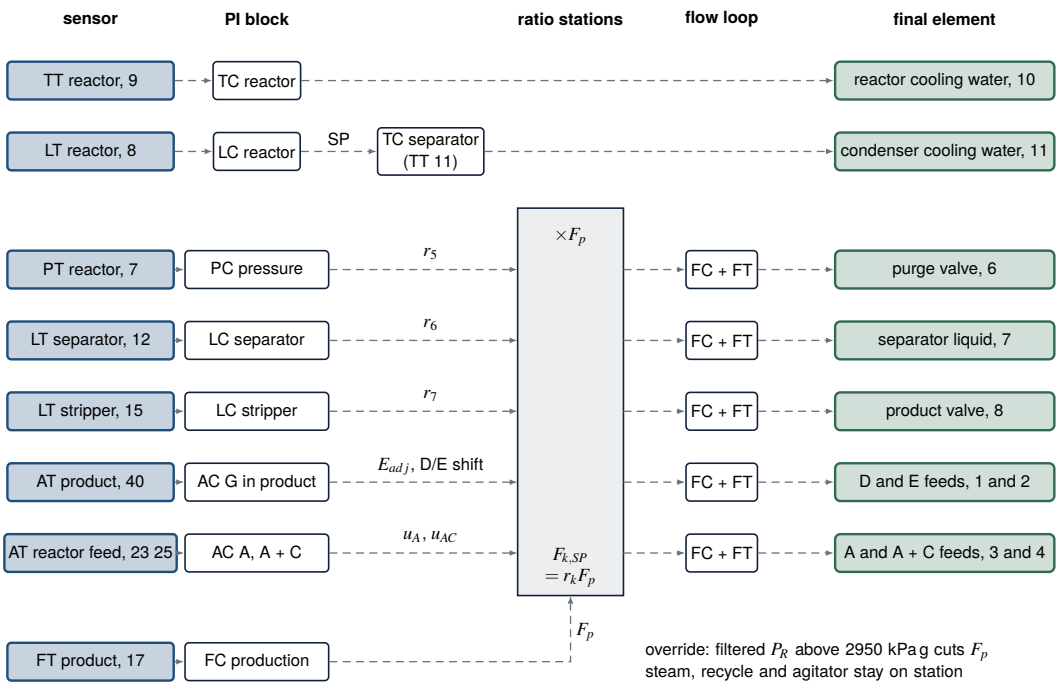
What Moves It

Each of the twelve valves follows its command through a first-order lag of 5 to 9 s, and 120 s for the steam valve. Under IDV 14, 15 and 19 a valve keeps its last command until the new one differs by more than 2 %. That rule is a dead band of the kind that stem friction gives a real valve [18], and case 7 shows the limit cycle it drives in the reactor temperature loop. Figure 8.5 sets the decentralized structure on such hardware, from each sensor to the valve it moves.

> **TECHNOLOGY Final Element: A Control Valve That Sticks**
>
> Friction between a valve stem and its packing is inevitable, and static friction is larger than the friction of a moving stem. A pneumatic actuator therefore holds the stem still until enough pressure builds to break it free. The stem then jumps until friction grabs it again, so a smooth signal turns into stair steps. This is stiction. To reverse direction the actuator must overcome the friction twice, which acts as a dead band in the loop. Over-tightened packing nuts are a common cause, and a test of small reversals of the actuator pressure measures the total friction. Many digital positioners record the actuator force against stem travel, and the gap between the opening and closing traces grows with the friction [18].

Where the Controller Runs

In a plant the seventeen loops of Table 8.3 would run in a distributed control system, whose supervisory level oversees localized controllers. Loops there are often nested or cascaded, with cycle times from milliseconds to minutes [19]. The application executes every loop at each 1 s step, and its predictive controller decides every 0.05 h above them. Supervisory control in the process industry is advanced regulatory control, a setpoint-based predictive controller on top of the PID layer, or a combination,



Numbers are XMEAS for sensors and XMV for valves. In a plant the blocks run in the distributed control system, and the application executes all of them at every 1 s step.

Figure 8.5 The decentralized structure on plant hardware, one row per loop of Table 8.3. The two temperature rows act on cooling water directly, and the reactor level sets the separator temperature. The five rows below send ratios to the ratio stations, which multiply each by the production index F_p set by the production loop at the bottom, and every product of a ratio and the index is a flow setpoint for a flow loop on its valve. Notice that one handle, F_p, moves every flow at once.

often implemented in the same distributed control system [15]. The production index with its ratio stations is such an advanced regulatory layer, and the chapter's predictive structure is the setpoint-based kind. An override is built with a select function that lets only one controller output reach the process at a time, and the de-selected controller must not wind up while it has no control [18]. The chapter's override meets that need by writing the index it applies back into the production loop's output.

> **IN PRACTICE** **Nameplate First, Optimization Later**
>
> From startup the primary objective of a new plant is to reach its nameplate capacity in a reliable and predictable way. Optimization often comes only after the plant has run for a few years, and a top-down strategy can then be added if the plant has a good regulatory control system. A strategy that drives the plant to the right economics in a counterintuitive way is hard to keep in service, because operator understanding suffers [3]. A dynamic model is usually not available at startup either, so a new plant first gets a simpler system of advanced regulatory elements, and predictive control is considered only if that system performs poorly [15].

What Is Modernizing It

Two developments bear on this plant. The first is the choice between advanced regulatory control and model predictive control in the supervisory layer. Constraints, often claimed as the special

advantage of predictive control, can in most cases also be handled by regulatory elements such as selectors, split-range control and anti-windup. Yet such solutions can grow complex, with many layers of cascades and selectors, and predictive control may then be the simpler one to implement and understand [15]. On the challenge the chapter's predictive structure passes at 188.0 $/h against 187.3 for the decentralized loops (Table 8.7). The second is valve diagnostics. Digital positioners that monitor the actuator's force against the stem's motion turn friction into a measured signature [18]. Case 7 shows why that matters: no gain of the loop removes the cycle of a sticking valve.

8.8 Limitations

The model is the public code, so it carries that code's simplifications: equilibrium vessels, a stripper of fixed separation factors, algebraic flows and a chemistry named by letters only, which maps onto no named process. Its limits and prices are the benchmark's teaching values. The port has been compared point by point with two translations, but the Fortran was checked by reading, not by compiling and running it, and the 1993 and 1996 papers were not available when the port was checked, so nothing printed in them is quoted or compared. The open-loop trip time depends on the rounding of the constants, by 0.59 h across the three readings measured. The noise realizations differ from the original generator's, so a trajectory published with the original code is matched in its statistics and not point by point.

The decentralized structure is the archived Simulink one, not the 1996 strategy, and it differs from Ricker's recordings in the stripper level and under the A feed loss. The override and the predictive controller are this application's designs, the predictive model is linear and identified at the base case, and the controllers meet no model error beyond what the plant itself presents. The individual case results use the application's seed, with the additional realizations stated beside each result. The challenge also separates 31 instrument-noise streams from 31 physical feed paths, as reported above. Half-step comparisons preserve the physical analyzer and disturbance clocks.

> **OPEN CHALLENGE Coordinating the Layers**
>
> The plantwide procedures written around plants like this one say what each layer of control must do, but many research issues remain in filling in the detail of each step, above all in the bottom-up part. More work is needed to understand how to decompose the layers of a control system and coordinate them [2]. A review of 2000 judged the field to be at a relatively early stage and called for a common ground on which case studies could be compared. It expected that, once alternative structures could be generated and analyzed automatically, the main problem would be building process models efficiently [4]. A predictive layer shares that need, because obtaining and maintaining its full dynamic model is costly [15].

8.9 Conclusion

The Tennessee Eastman plant runs here as a JavaScript port of the public Fortran model, translated through a BSD-3-Clause Python port, checked against the Fortran by reading, against the Python port to 1.76×10^{-11} and against Ricker's C translation bit for bit at the base case and to 4×10^{-10} over six hours, with identical shutdown steps. Its base case is a steady state that is not stable, and its open-loop trip at 3.47, 3.59 or 4.06 h identifies the rounding of the constants more than the plant. The inventory loops hold it until a feed changes, the archived decentralized structure rides the feed change and cuts the mean cost from 215.7 to 187.3 $/h when the pressure setpoint moves from 2705 to 2800 kPa g, and the same structure trips at 8.40 h on an A feed loss that Ricker's recorded strategy rode. Only an integrating override and a production cut keep it alive. The port omits the 2015 revision's added disturbances, and its noise matches the original only in distribution. A fuller port would add the 1996 formulation with its overrides, the revision's disturbances and measurements, and the remaining operating modes as challenges. The simulator puts the plantwide benchmark of three decades where a student can watch it shut down.

KEY POINTS

- Left with every valve fixed and no disturbance, the published steady state does not hold: the reactor pressure reaches its interlock at 3.59 h.
- The inventory loops make the plant stable, and only the composition loops above them make it survive a feed change.
- Moving the pressure setpoint from 2705 to 2800 kPa g cuts the challenge's mean cost from 215.7 to 187.3 $/h, and the price is margin.
- The archived structure trips at 8.40 h on an A feed loss that Ricker's recorded strategy rode. Only an integrating override and a production cut keep it alive.
- A trip time quoted for this plant identifies an implementation as much as a dynamic.

Where else this applies. An override that sits idle in normal operation also guards the compressor of Volume 4, Chapter 2, and an inert purged from a gas recycle under a pressure override returns in Volume 13, Chapter 2.

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8.10 Exercises

8.1 Concept check · 2 min

Three translations of the same public code are run open loop, with every valve fixed and no disturbance, on the same 1 s Euler step. They trip on high reactor pressure at 3.47, 3.59 and 4.06 h. What does the spread show?

(a) The Euler step is too coarse, so each translation integrates the unstable mode with its own error

(b) One translation carries an error in its balances that the check of the base case did not catch

(c) The measurement noise differs between the translations, and the noise moves the time of the trip

(d) The unstable mode grows from the rounding residue of the decimal constants, which each reading leaves

8.2 Concept check · 1 min

In case 5 the feed composition wanders at random for a day, and raising the reactor pressure setpoint from 2705 to 2800 kPa g lowers the mean operating cost from 176.1 to 146.3 \$/h. Which mechanism does the chapter give?

(a) A higher pressure carries the same inerts out in less purge gas, so less A, C, D and E leave with it

(b) A higher pressure speeds up the reactions, so the plant makes more product from the same feeds

(c) A higher pressure lowers the compressor work, which is the largest term of the operating cost

(d) A higher pressure damps the random feed walk, so the composition loops move the feeds less

8.3 Concept check · 2 min

Case 3 ramps the production setpoint down by 15 %. The run is repeated with the stripper level gain raised to four times Ricker's value. What does the chapter measure?

(a) The stripper level swings less, and the production also stays closer to its setpoint in force

(b) The stripper level swings less, and the production strays further from its setpoint in force

(c) The stripper level swings as before, because its 200 min integral time sets the swing

(d) The stripper level begins to cycle, and the plant trips on the stripper level interlock

8.4 Theory and calculation · 6 min

The decentralized structure sets the production index from Equation (8.9), and its production setpoint passes a rate limit of 30 % of the base rate of 22.95 m^3/h per day before it is in force.

(a) With the production loop's trim at zero, find the index F_p for a setpoint of 22.95 m^3/h and for the challenge's raised setpoint of 25.25 m^3/h. *Checkpoint:* 100.26 % and 110.31 %.

(b) The challenge raises the setpoint from 22.95 to 25.25 m^3/h at 24 h. Find the rate limit and the time the setpoint in force needs to reach the new value. *Checkpoint:* 0.287 m^3/h per hour and 8.0 h.

(c) The challenge scores the mean true production over 48 h against the mean of the setpoint the reader enters, not the one in force. Find the score a plant would get if its production followed the setpoint in force exactly, and compare it with the 99.1 % of the suggested run. *Checkpoint:* 99.2 %.

8.5 Theory and calculation · 6 min

The pressure override of Equation (8.10) filters the measured reactor pressure over 0.02 h at the 1 s step, and then takes $K_{ovr} = 2$ points of F_p per kPa of filtered excess every second.

(a) Find the filter coefficient $1 - \mathrm{e}^{-\Delta t/0.02}$ per step and the filter's time constant in steps. *Checkpoint:* 0.0138 and 72 steps.

(b) The measurement noise is 0.298 kPa rms and independent from step to step. Driven by such noise, this filter's output has a variance $\alpha/(2-\alpha)$ times that of its input, where α is the coefficient of part (a). Find the rms noise left after the filter. *Checkpoint:* 0.025 kPa, twelve times less.

(c) With the production loop frozen at the override's first step in case 8, F_p fell from 83.8 to 42.1 % in 3 min. Find the mean filtered excess this needs, and compare it with the largest excess of 1.11 kPa. *Checkpoint:* 0.116 kPa.

8.6 Theory and calculation · 6 min

At the base case the port reads a compressor work $W = 341.43$ kW, a steam flow $S = 230.31$ kg/h and a purge $F_9 = 0.3371$ kscmh of 32.96, 13.82, 23.98, 1.257, 18.58, 2.263, 4.844 and 2.299 mol % A to H. The product holds 0.0179, 0.8357 and 0.0986 mol % D, E and F, with the stripper valve at its base opening of 46.534 %. Use Equation (8.8) and its prices.

(a) Find the power and steam terms. *Checkpoint:* 18.30 $/h and 7.32 $/h.

(b) Find the purge term and its share of the total of 170.62 $/h. *Checkpoint:* 114.73 $/h, 67 %.

(c) Find the product-loss term, and the same term with the composition printed in Ricker's table, 0.02, 0.84 and 0.10 mol %. *Checkpoint:* 30.27 $/h and 30.56 $/h, 0.28 $/h apart.

8.7 App challenge · 8 min

Use the default seed and change nothing that the parts do not name.

(a) Load scenario 1, Every valve fixed: the open-loop shutdown, and run it in Manual at the default speed of 0.25 h per second until it stops. Read the time and the cause on the status strip, and in the exported file the true reactor pressure, column `truth_xmeas07_reactor_press_kpag`, at 2 and 3 h. *Checkpoint:* 3.59 h, reactor pressure high. 2705.0 and 2689.6 kPa g, within 0.5 kPa.

(b) Load scenario 2, Inventory loops only, and run it as shipped at the same speed until it stops. Find in the exported file the first row with the true reactor pressure above 2900 kPa g, the shutdown time, and the lowest A share of A + C in the reactor feed analyzer, column `yA_pct`. *Checkpoint:* 4.85 h within 0.02 h, 5.84 h, 37.0 % within 0.3.

(c) Choose Decentralized with pressure override in Structure (in Auto), press Reset and run 8 h at 1 h per second. Read the highest true reactor pressure and the lowest A share in the new file. *Checkpoint:* 2751 kPa g within 2 kPa, 45.4 % within 0.3.

Explain the three results in terms of the layers of the structure.

8.8 App challenge · 8 min

Load scenario 8, A feed lost (IDV 6) and the pressure override. The A feed is lost at 1 h. Use the default seed and a speed of 4 h per second.

(a) Run the scenario as shipped until it stops, and read the time and the cause on the status strip. *Checkpoint:* 8.40 h, stripper level low.

(b) Press the reload button, set the Production setpoint to 17 m^3/h before Run, and run 24 h. In the exported file find the highest true reactor pressure and the mean true production, column `truth_xmeas17_strip_underflow_m3h`, over the rows after 20 h. *Checkpoint:* No shutdown. 2974 kPa g within 2 kPa, and 5.88 m^3/h within 0.02.

(c) Choose Decentralized without override in Structure (in Auto), press Reset and run again. *Checkpoint:* 13.34 h, reactor pressure high.

Say which part of the structure each result belongs to.

8.9 App challenge · 8 min

Load the Challenge, Two days of upsets at lower cost. Use the default seed and a speed of 4 h per second, and leave the model and the disturbances as given.

(a) Run it as shipped and read the time and the cause of the stop. *Checkpoint:* 6.68 h, reactor pressure high.

(b) Choose Decentralized with pressure override in Structure (in Auto), press Reset and run the 48 h. Read the Mean cost tile and the note at the end of the run. *Checkpoint:* 215.7 $/h, missed on cost (worst G 2.42 points, production 99.1 %).

(c) Set the Reactor pressure setpoint to 2800 kPa g, press Reset and run again. Read the same note, and find in the exported file the highest true reactor pressure and its time. *Checkpoint:* Target met, 187.3 $/h, worst G 2.33 points, production 99.1 %. 2944 kPa g near 5.53 h, within 2 kPa and 0.05 h.

Accept a mean cost within 0.1 $/h. Explain what the structure and the pressure setpoint each contribute, and what the second change costs.

8.10 Application area · 4 min

An operations engineer wants to move a plant like this one from a reactor pressure setpoint of 2705 to 2800 kPa g, on the evidence of case 5 and the challenge.

(a) Case 5 gives mean costs of 176.1 and 146.3 $/h at the two setpoints. Find the saving over a year of 8760 h. *Checkpoint:* About $261 000 per year.

(b) The peak pressures are 2756 and 2827 kPa g in case 5 and 2944 kPa g in the challenge at 2800 kPa g. Find each margin to the 3000 kPa g interlock, and name what must be established before the move on a real plant. *Checkpoint:* 244, 173 and 56 kPa.

8.11 Application area · 4 min

The two composition loops and the mix loop of the decentralized structure read sampled analyzers, as in Equation (8.7).

(a) Find the youngest, oldest and mean age of a reading of the product analyzer, period 0.25 h, and of the reactor feed analyzer, period 0.1 h. *Checkpoint:* 0.25 to 0.5 h, mean 0.375 h. 0.1 to 0.2 h, mean 0.15 h.

(b) Case 9 moves the G setpoint from 53.72 to 11.7 mol %. Find how long the setpoint ramp lasts at Ricker's 50 mol % per day and at 10 mol % per hour, and how far the setpoint moves during the age of the oldest product reading in each case. *Checkpoint:* 20.2 h and 4.2 h. 1.04 and 5 mol %.

8.12 Review the agent · 5 min

An AI assistant is asked why scenario 8, A feed lost (IDV 6) and the pressure override, ends at 8.40 h and replies as follows. "The loss of the A feed raises the reactor pressure, and the override, at 2 points of F_p per kPa per second, is too gentle to pull production down in time. Raise the Override gain to 5 and the decentralized structure will ride the loss at the full production setpoint." Find the flaw in the reply and say how it could be checked.