Ratio and bidirectional control applied to distillation columns

Brage Bang

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Trondheim

17 June 2025
DYCOPS
Bratislava, Slovakia

“The goal of my research is to develop simple yet rigorous methods to solve problems of engineering significance”


https://folk.ntnu.no/skoge/

did v2 rockets use der... adapti skogestad KLM Royal Dutch Airli... All Bookmarks

Chemical Engineering

Sigurd Skogestad Professor

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Start here...

- About me - CV - Powerpoint presentations - How to reach me - Email: skoge@ntnu.no
- Teaching: Courses - Master students - Project students
- Research: My Group - Research - Ph.D. students - Academic tree

"The overall goal of my research is to develop simple yet rigorous methods to solve problems of engineering significance"

"We want to find a self-optimizing control structure where close-to-optimalo operation under varying conditions is achieved with constant (or slowly varying) setpoints for the controlled variables (CVs). The aim is to move more of the burden of economic optimization from the slower time scale of the real-time optimization (RTO) layer to the faster setpoint control layer. More generally, the idea is to use the model (or sometimes data) off-line to find properties of the optimal solution suited for (simple) on-line feedback implementation"

"News"...

- 27 Nov. 2023: Welcome to the SUBPRO Symposium at the Britannia Hotel in Trondheim
- Aug. 2023: Tutorial review paper on "Advanced control using decomposition and simple elements". Published in Annual reviews in Control (2023). [paper] [tutorial workshop] [slides from Advanced process control course at NTNU]
- 05 Jan. 2023: Tutorial paper on "Transformed inputs for linearization, decoupling and feedforward control" published in JPC. [paper]
- 13 June 2022: Plenary talk on "Putting optimization into the control layer using the magic of feedback control", at ESCAPE-32 conference, Toulouse, France [slides]
- 08 Dec. 2021: Plenary talk on "Nonlinear input transformations for disturbance rejection, decoupling and linearization" at Control Conference of Africa (CCA 2021), Magaliesburg, South Africa (virtual) [video and slides]
- 27 Oct. 2021: Plenary talk on "Advanced process control - A newe look at the old" at the Brazilian Chemical Engineering Conference, COBEQ 2021, Gramado, Brazil (virtual) [slides]
- 13 Oct. 2021: Plenary talk on "Advanced process control" at the Mexican Control Conference, CNCA 2021 (virtual) [video and slides]
- Nov. 2019: Sigurd receives the "Computing in chemical engineering award from the American Institute of Chemical Engineering (Orlando, 12 Nov. 2019)
- June 2019: Best paper award at ESCAPE 2019 conference in Eindhoven, The Netherlands
- July 2018: PID-paper in JPC that verifies SIMC PI-rules and gives "Improved" SIMC PID-rules for processes with time delay (taud=theta/3)
- June 2018: Video of Sigurd giving lecture at ESCAPE-2018 in Graz on how to use classical advanced control for switching between active constraints
- Feb. 2017: Youtube videos of Sigurd giving lectures on PID control and Plantwide control (at University of Salamanca, Spain)
- 06-08 June 2016: IFAC Symposium on Dynamics and Control of Process Systems, including Biosystems (DYCOPS-2016), Trondheim, Norway.
- Videos and proceedings from DYCOPS-2016
- Aug 2014: Sigurd recieves IFAC Fellow Award in Cape Town
- 2014: Overview papers on "control structure design and "economic plantwide control"
- OLD NEWS

Books...

- Book: S. Skogestad and I. Postlethwaite: MULTIVARIABLE FEEDBACK CONTROL-Analysis and design. Wiley (1996; 2005)
- Book: S. Skogestad: CHEMICAL AND ENERGY PROCESS ENGINEERING CRC Press (Taylor&Francis Group) (Aug. 2008)
- Bok: S. Skogestad: PROSESSTEKNIKK- Masse- og energibalanser Tapir (2000; 2003; 2009).

More information ...

- **Publications** from my Google scholar site
- **Download publications** from my official publication list .... or look HERE if you want to download our most recent and upublished work
- Proceedings from conferences - some of these may be difficult to obtain elsewhere
- Process control library - We have an extensive library for which Ivar has made a nice on-line search
- Photographs that I have collected from various events (maybe you are included...)
- International conferences - updated with irregular intervals
- SUBPRO (NTNU center on subsea production and processing) [Annual reports] [Internal]
- Nordic Process Control working group - in which we participate
- 5-year Master program in Chemical and Biochemical Engineering at NTNU (MTKJ) - Sigurd Skogestad is Program Leader 2019-2025.

This talk

Simple data-based architectures applies to distillation control

- PID control (feedback)
- Ratio control
- Bidirectional control with selectors

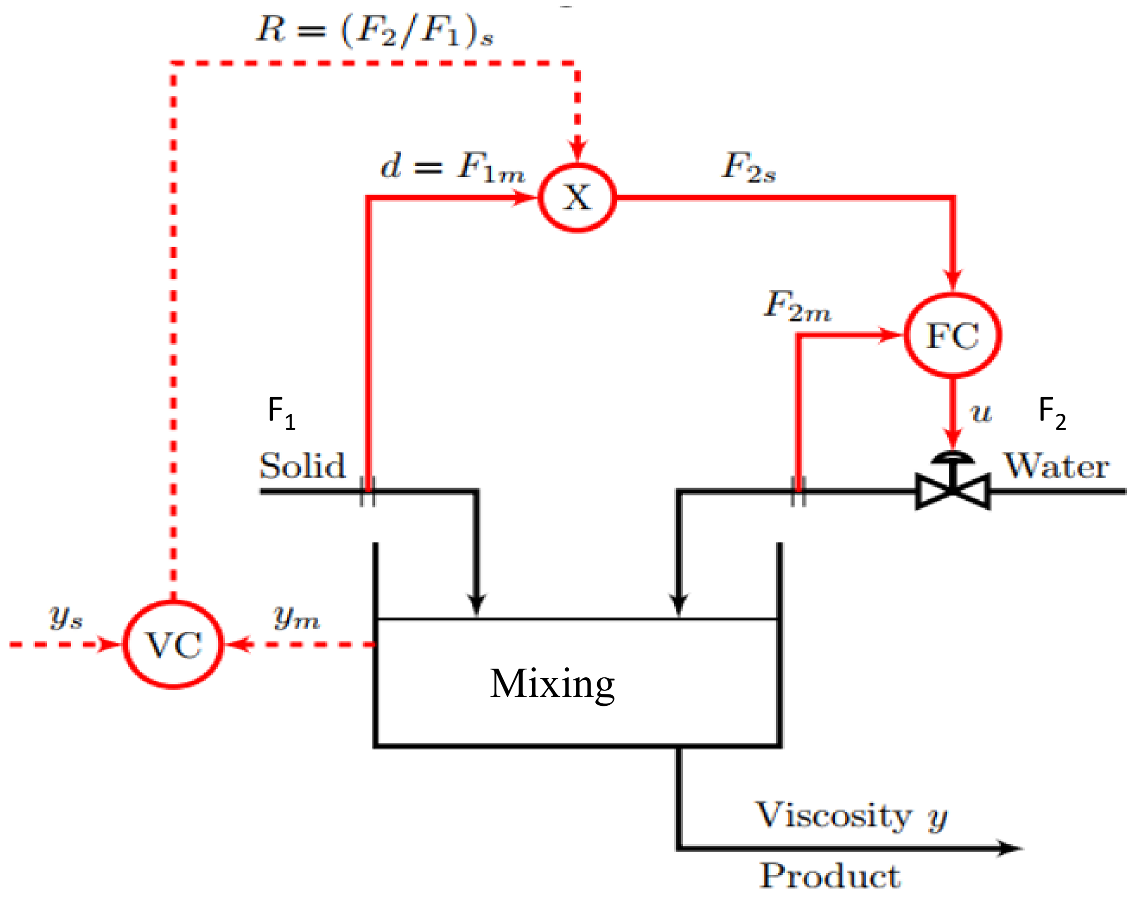
Ratio control

- Feedforward control without a model y=f(u,d)
- Just process insight
- Example: Food recipe
  - 1 part sugar
  - 3 parts milk
  - 3 parts coffee
  - MIX

Usually: Combine ratio (feedforward) with feedback

***Example cake baking**: Use recipe (ratio control = feedforward),*
but a good cook adjusts the ratio to get desired result (feedback)

Ratio control combined with feedback



- **The correct ratio setpoint is found by feedback control** (VC) based on keeping the measured controlled variable y at its desired setpoint y_s..
  - So also here **no model** is needed (just data)

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Distillation

At home doing moonshine distillation (1979)

Chemical Engineering Research and Design

Trans IChemE, Part A, January 2007

THE DOS AND DON´TS OF DISTILLATION COLUMN CONTROL

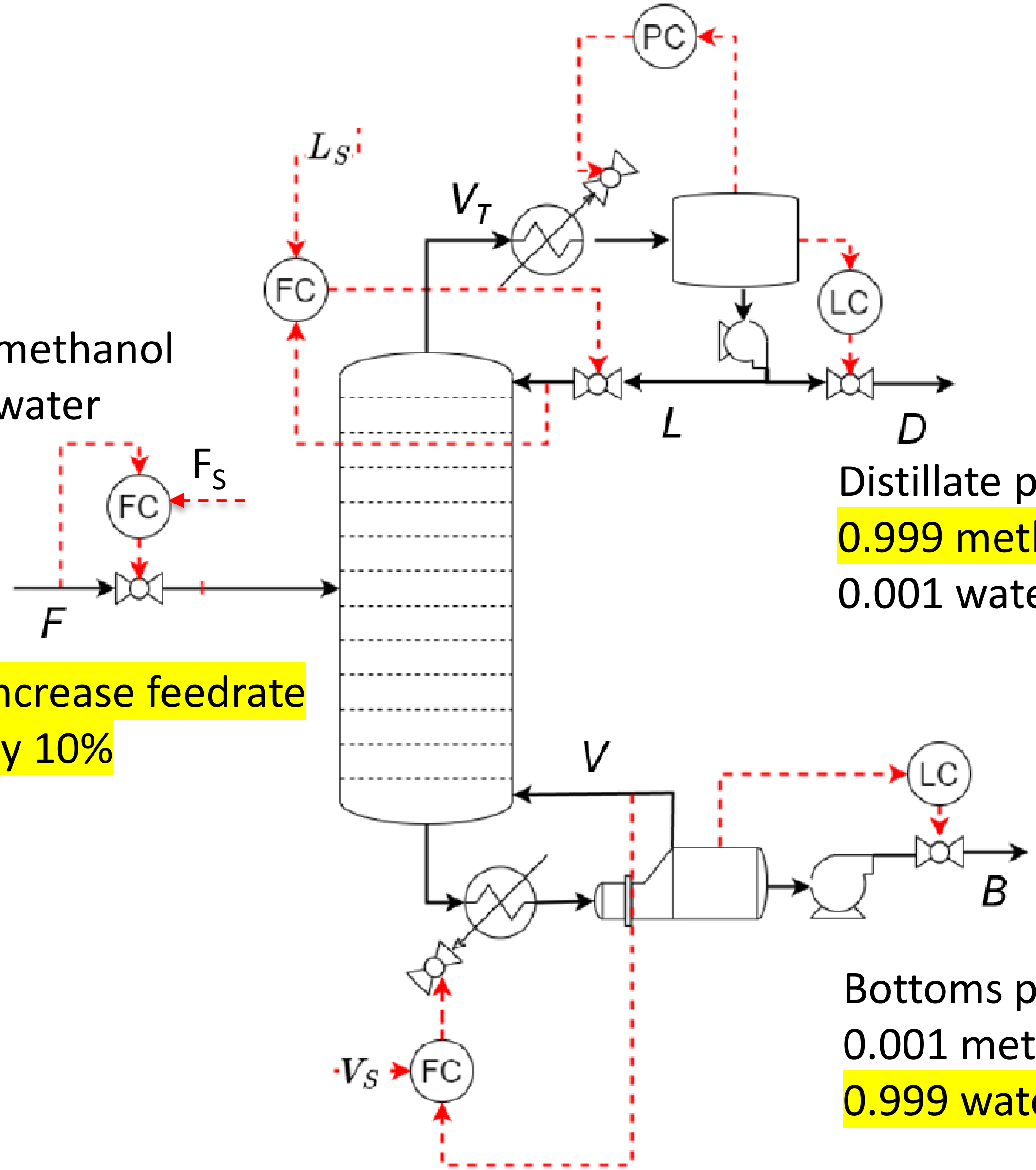
S. Skogestad*

Department of Chemical Engineering, Norwegian University of Science and Technology, Trondheim, Norway.

Abstract: The paper discusses distillation column control within the general framework of plant-wide control. In addition, it aims at providing simple recommendations to assist the engineer in designing control systems for distillation columns. The standard LV-configuration for level control combined with a fast temperature loop is recommended for most columns.

Feed:
0.50 methanol
0.50 water

Increase feedrate by 10%



Distillate product:
0.999 methanol
0.001 water

Bottoms product:
0.001 methanol
0.999 water

6 indpependent flows (valves):

1. Feed F (set F_S, disturbance)
2. Distillate D (top level control)
3. Bottom product B (bottom level control)
4. Cooling V_T (pressure control)
5. Reflux L (top composition control)
6. Boilup V (bottom composition control)

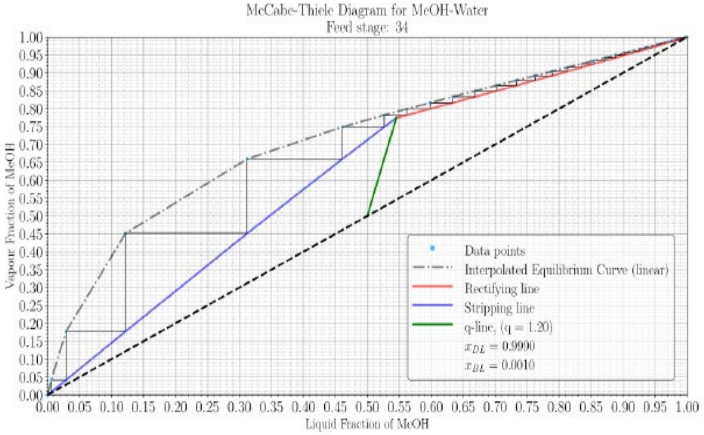


Figure 2.1 – McCabe-Thiele diagram for the water-methanol distillation column.

Parameter	Value
Number of theoretical stages	40
Feed stage (numbered from top)	34
Feed flow F	100 kmol/h
Feed mole fraction (methanol)	0.50
Feed state	Liquid
Column pressure	2 bar
Reflux ratio L/D	1.013
Top product, x_D (water)	0.001
Bottom product, x_B (methanol)	0.001
Reboiler type	Kettle
Vapor-liquid equilibrium model	NRTL

Table 1. Nominal operating data for the methanol-water distillation column

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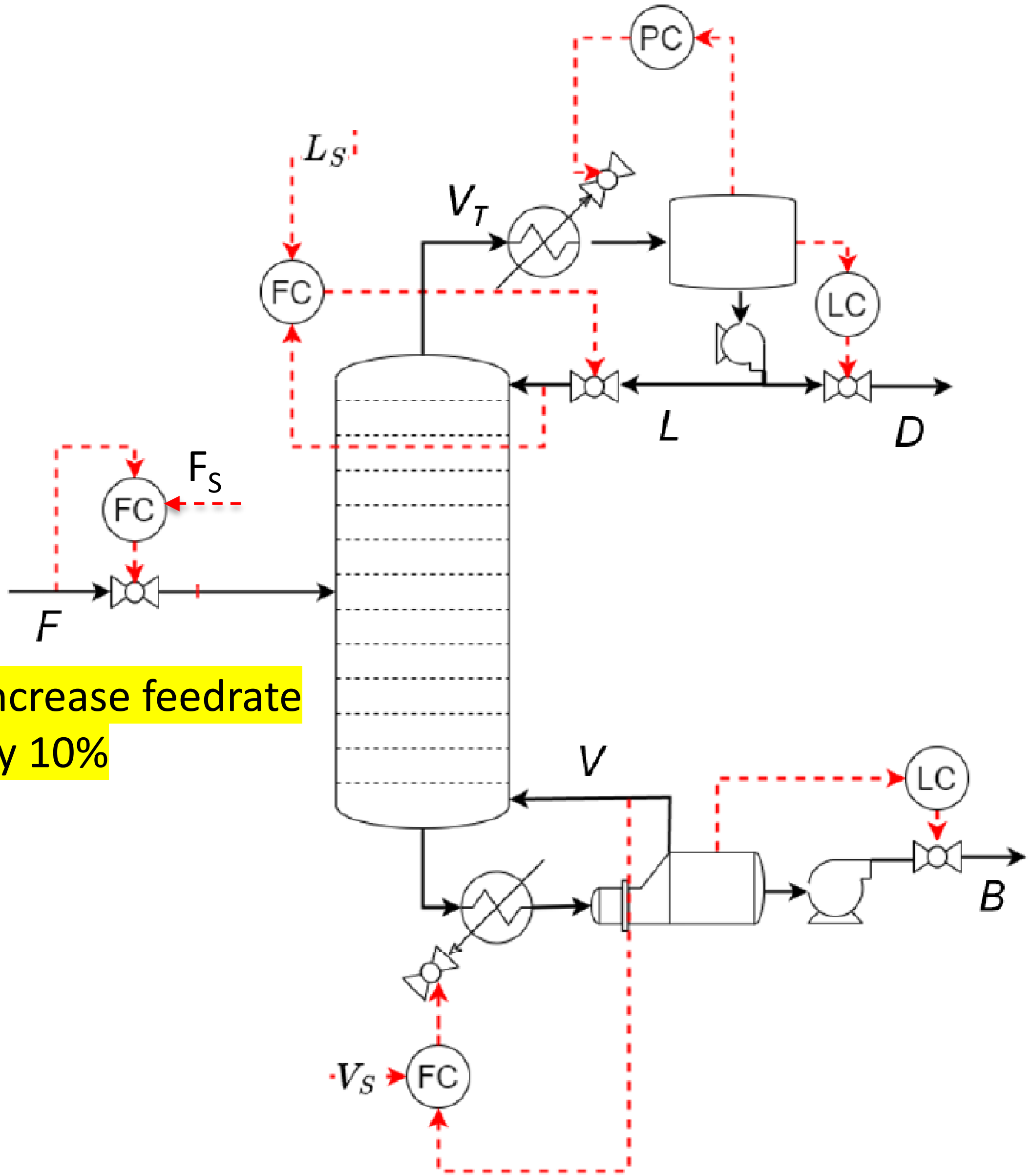
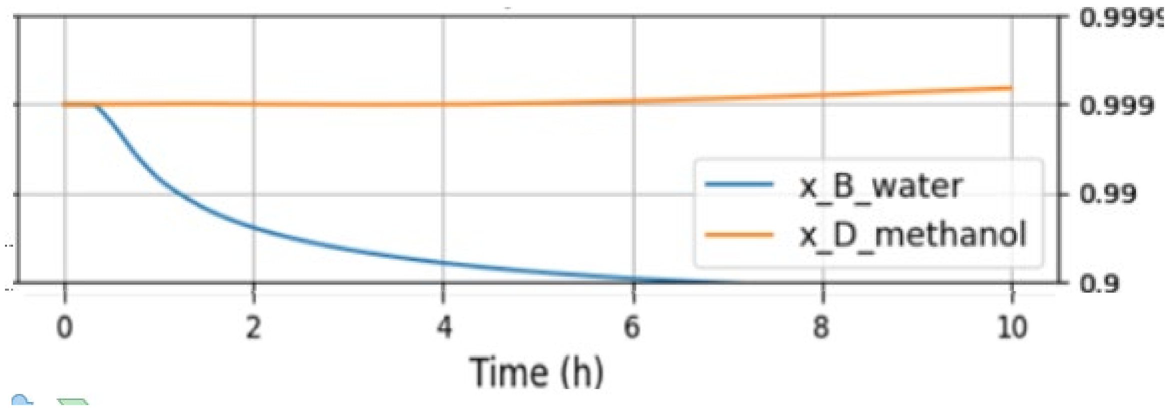
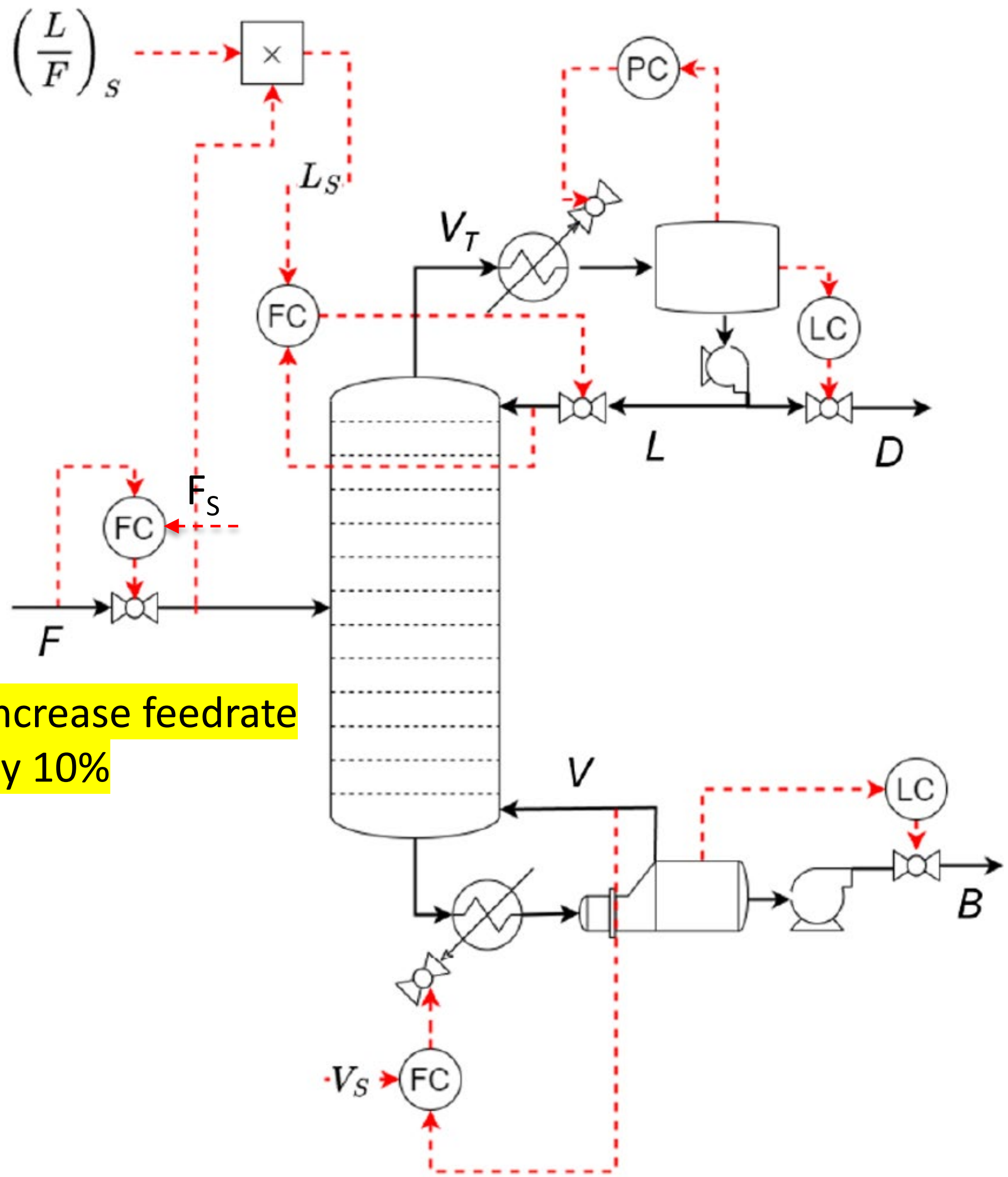
B1: L and V constant



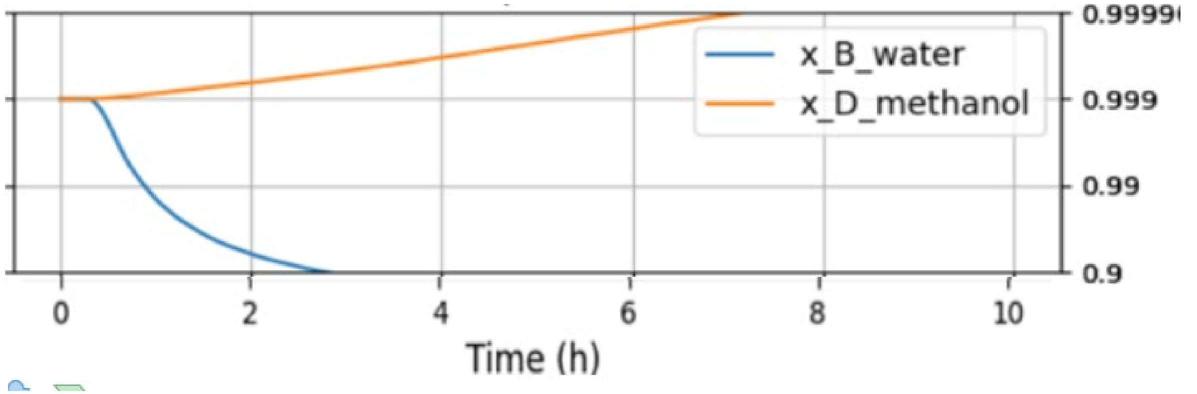
Increase feedrate by 10%

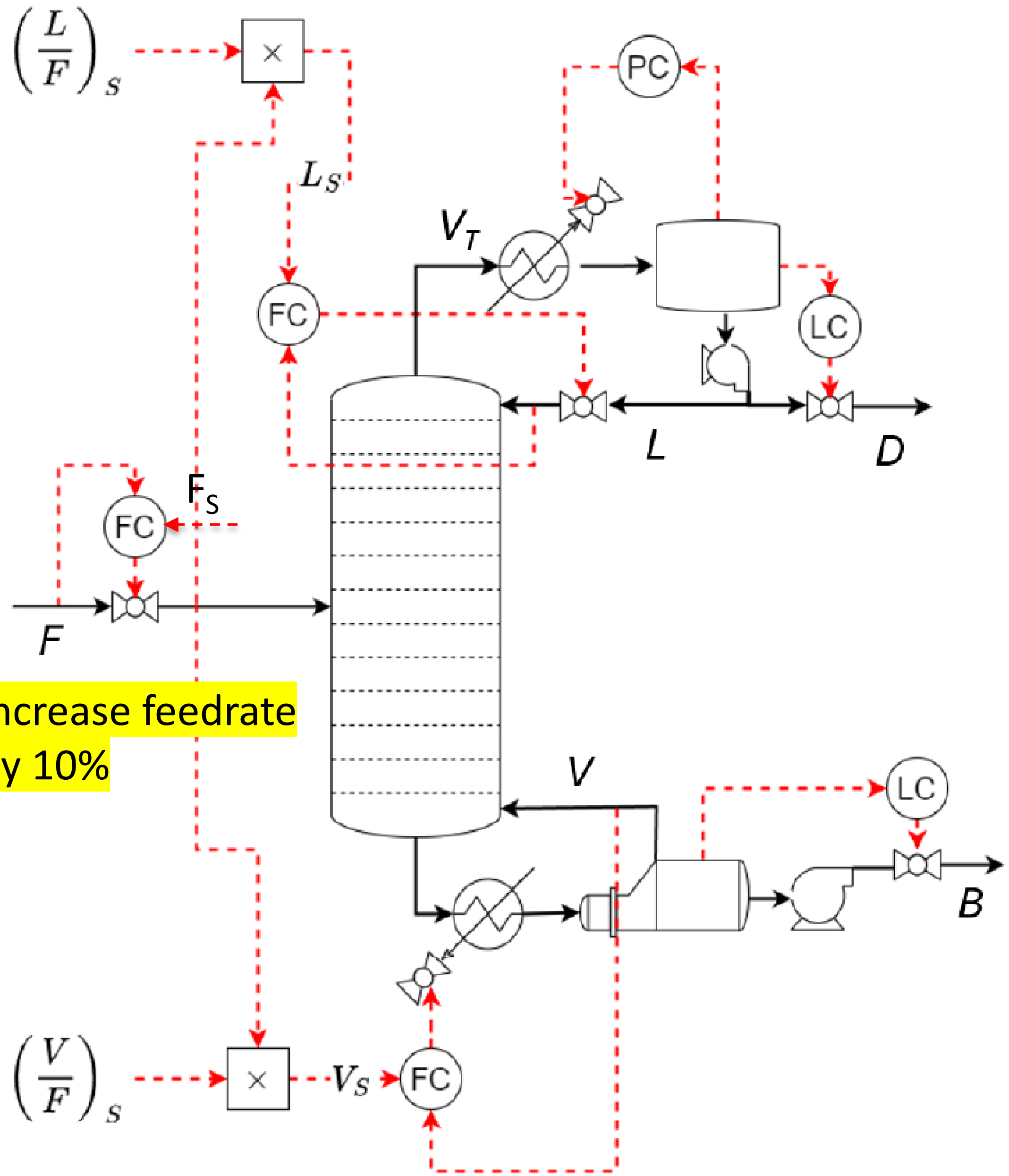


B2: L/F and V constant



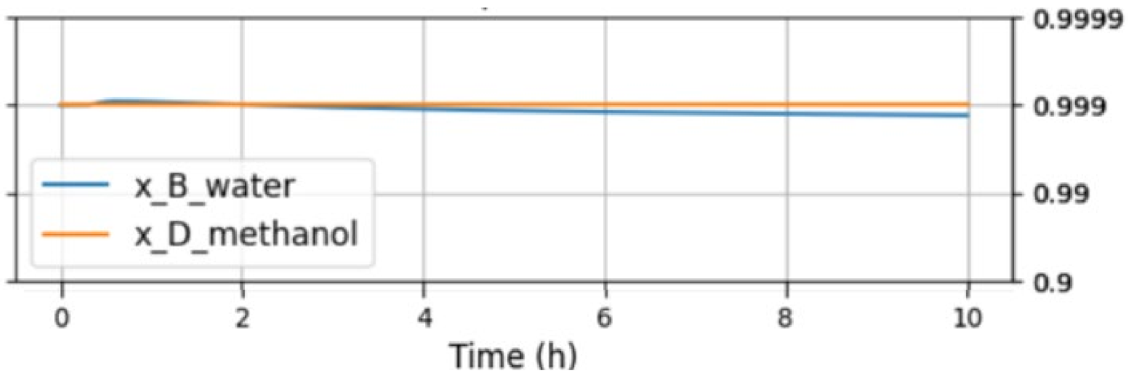
Increase feedrate by 10%





Increase feedrate by 10%

B4: L/F and V/F constant

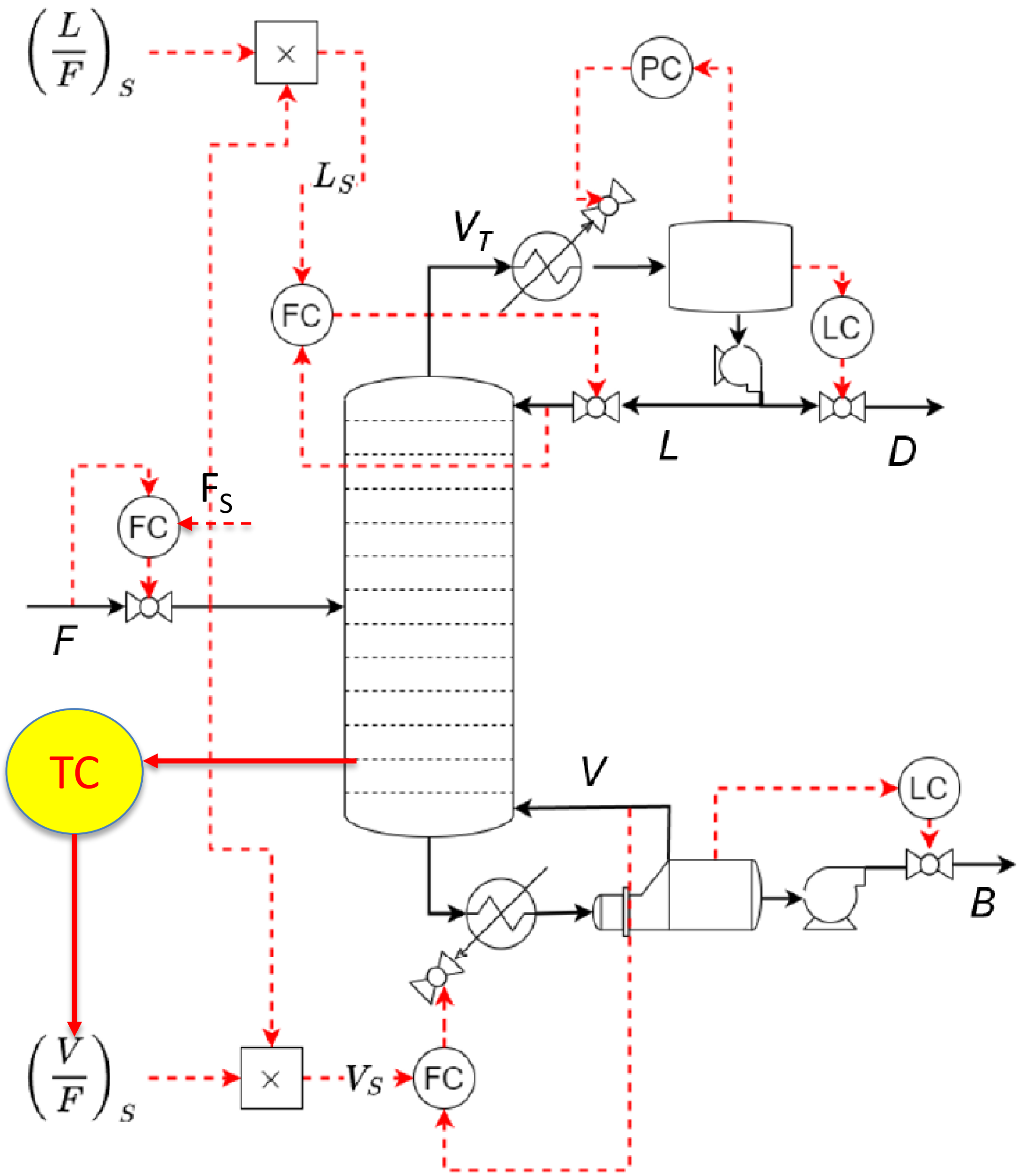


Oooops…. Drifting a little (even in simulation)….

…. Becuse feedforward

B3: L/F and T constant

Ratio with Cascade



B3: L/F and T constant

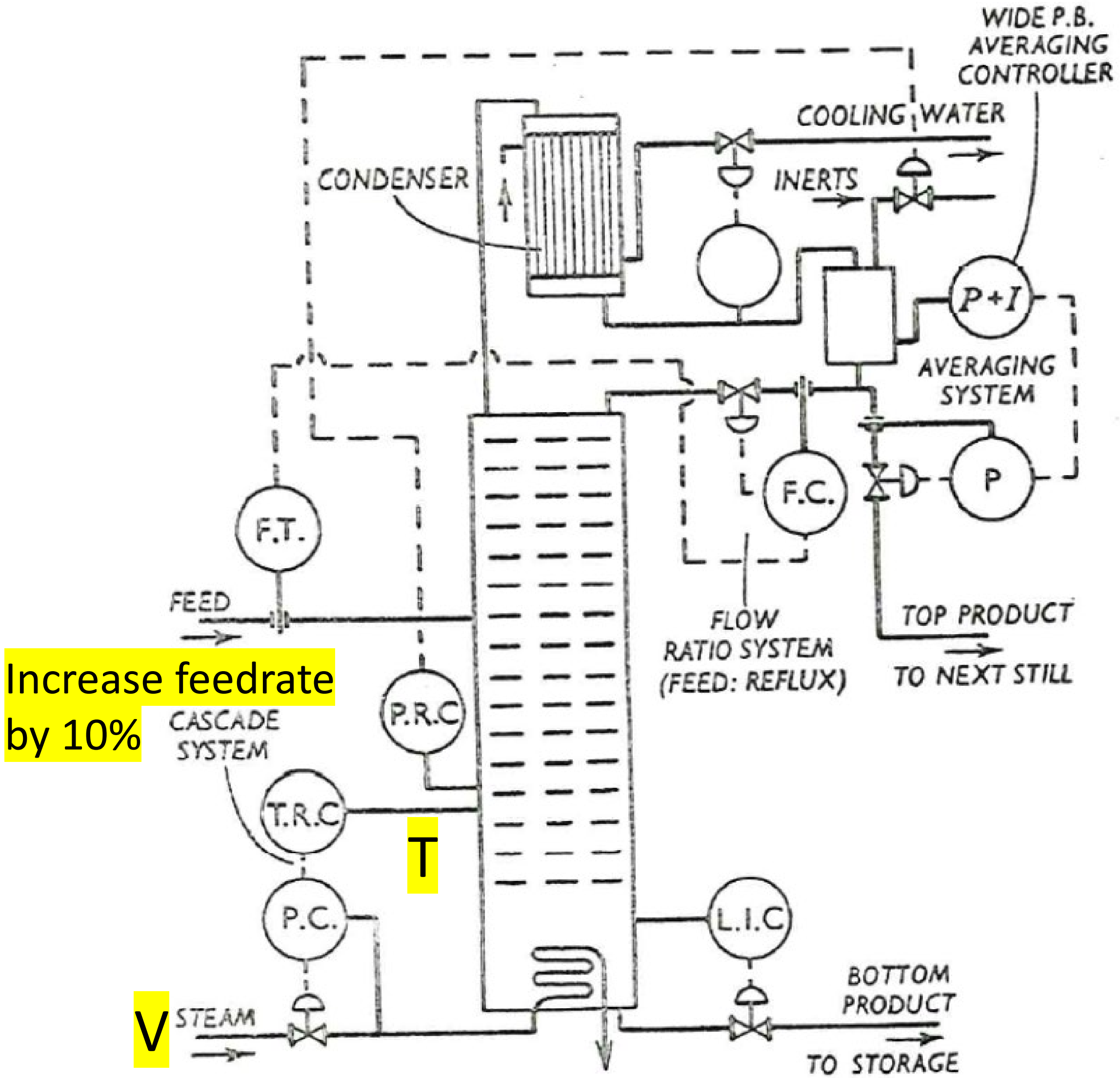
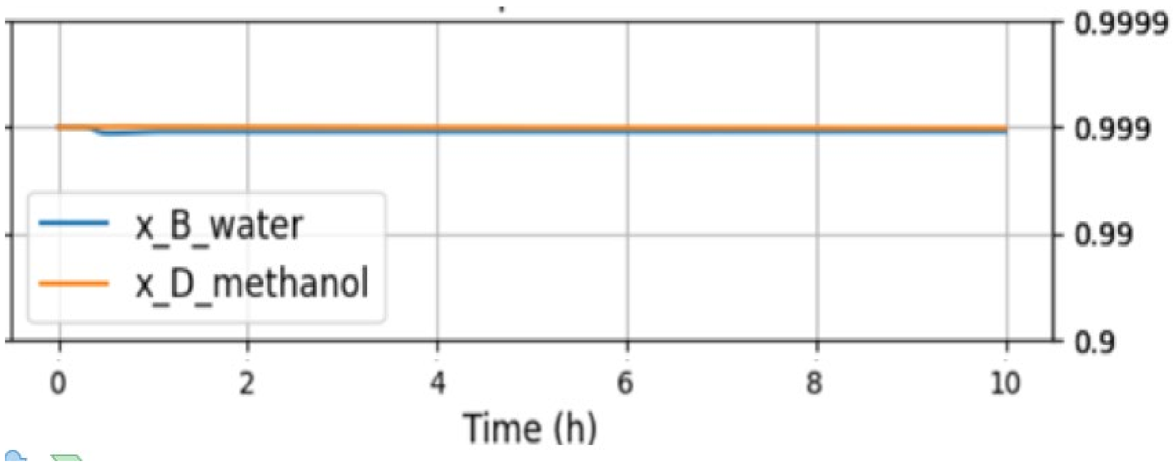


FIG. 18.11. Examples of the use of complex systems.



The best!

… but still one problem… V must not saturate!

Young, A.J. (1955). *An introduction to process control system design*. Longman.

Summary

- Scheme B1. Constant L and V (bad)
- Scheme B2. Constant L/F and V (even worse)
- Scheme B3. Constant L/F and temperature in bottom section (best, but V must not saturate)
- Scheme B4. Constant L/F and V/F (good, but V must not saturate)

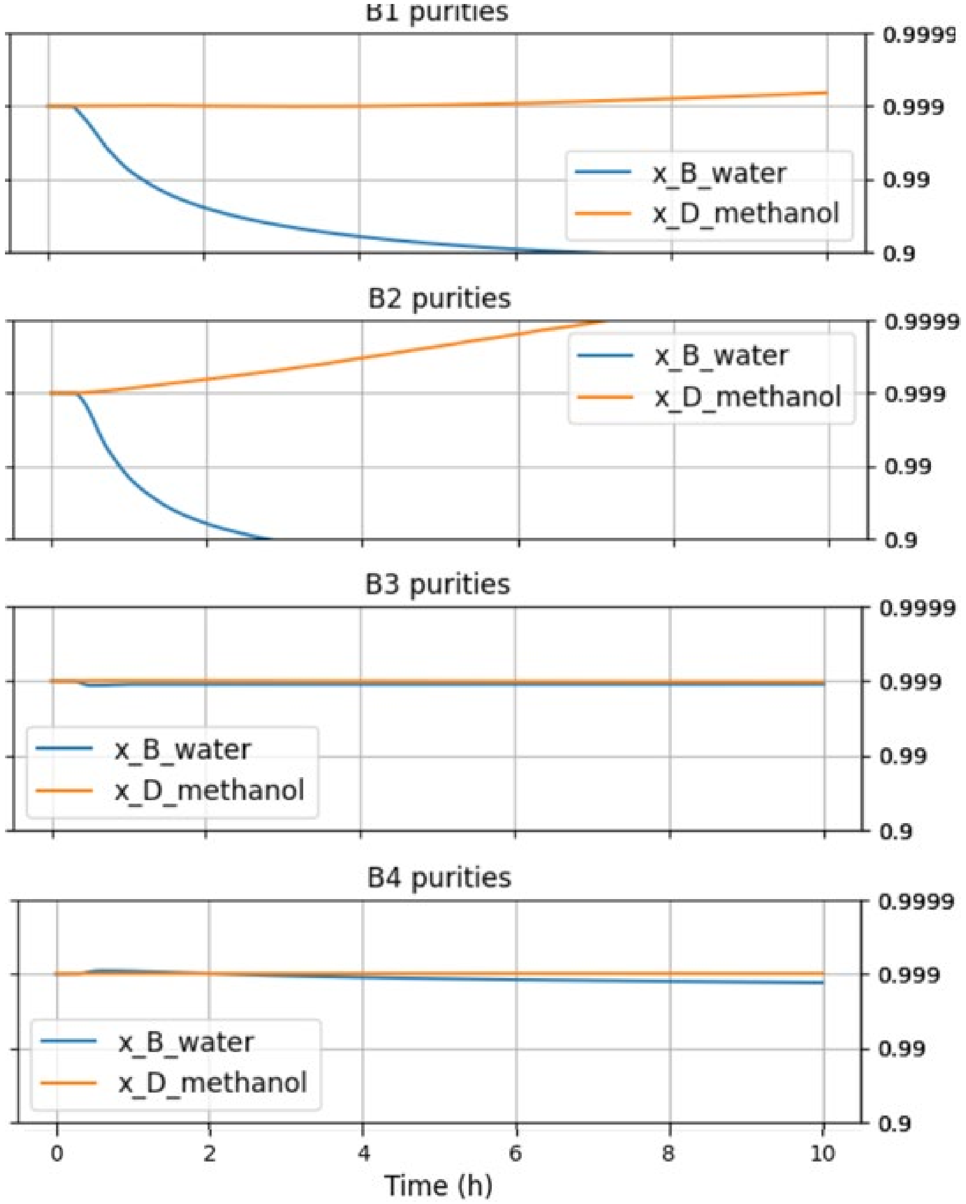
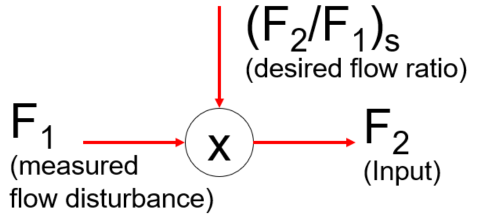


Fig. 4. Ratio control: Product composition responses for 10% feed flow disturbance with control structures B1, B2, B3 and B4.

When should we use ratio control?



Answer: When the **scaling property** holds such thay fixing the ratio of all **extensive** variables (e.g., F_2/F_1) gives constant **intensive** variables (y) at steady state.

- Similar to in thermodynamics, the scaling property holds exactly for equilibrium systems:
  - **Mixing**
  - **Equilibrium reactors**
  - **Distillation**
- General: If we have n steady-state independent extensive variables (flows), then we must fix n-1 ratios.
- Distillation: n=3 (with constant levels and pressure) $\Rightarrow$ Fix 2 ratios

Mathematically: Theory of ratio control

SCALING PROPERTY (steady state):

- Scaling <u>all</u> independent extensive variables by X the same factor,
- with all independent intensive variables x constant,
- scales all the dependent extensive extensive variables Y by the same factor
- And keeps all intensive variables y constant

$$y \text{ intensive}: \quad \underbrace{f_y(x_1, x_2, kX_1, kX_2, kX_3)}_{y(k)} = \underbrace{f_y(x_1, x_2, X_1, X_2, X_3)}_{y}$$

$$Y \text{ extensive}: \quad \underbrace{f_Y(x_1, x_2, kX_1, kX_2, kX_3)}_{Y(k)} = k\, \underbrace{f_Y(x_1, x_2, X_1, X_2, X_3)}_{Y}$$

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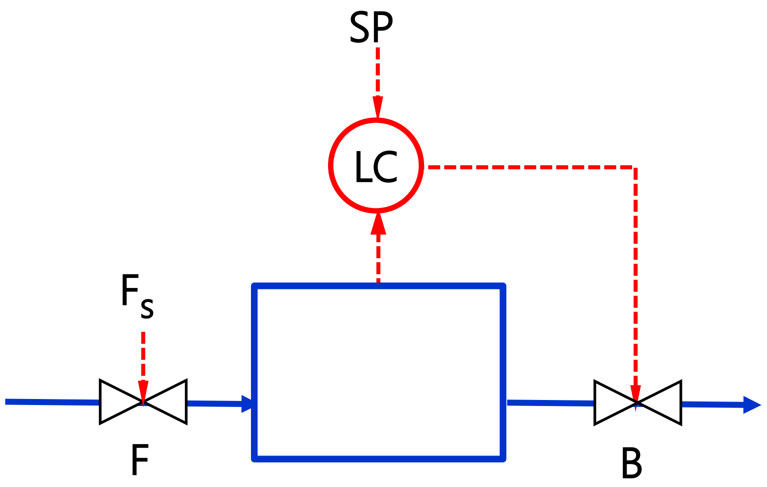
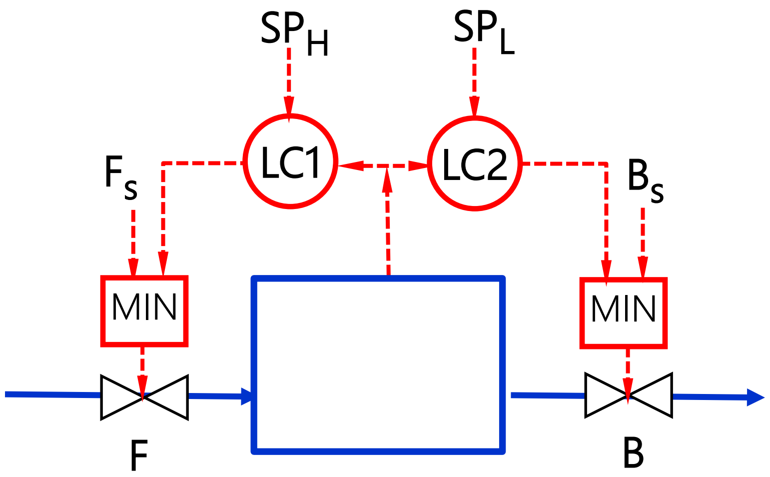
Conclusion ratio control

- It’s a great idea!
  - Very simple to implement
  - Gives nonlinear feedforward action
- Requires no model, just insight
  - Theoretical basis: Scaling principle
- For distillation: Must keep 2 ratios constant
  - Or 1 ratio and 1 temperature/composition
  - Don’t use ratio control for distillation if one flowrate saturates
  - Could use L/D rather than L/F (even when F is the disturbance)

Bidirectional control

- What should we do if B, D, L, V_T or V saturates (at max) so that we lose control?
- One solution: «Override» to reduce throughput (feedrate F)





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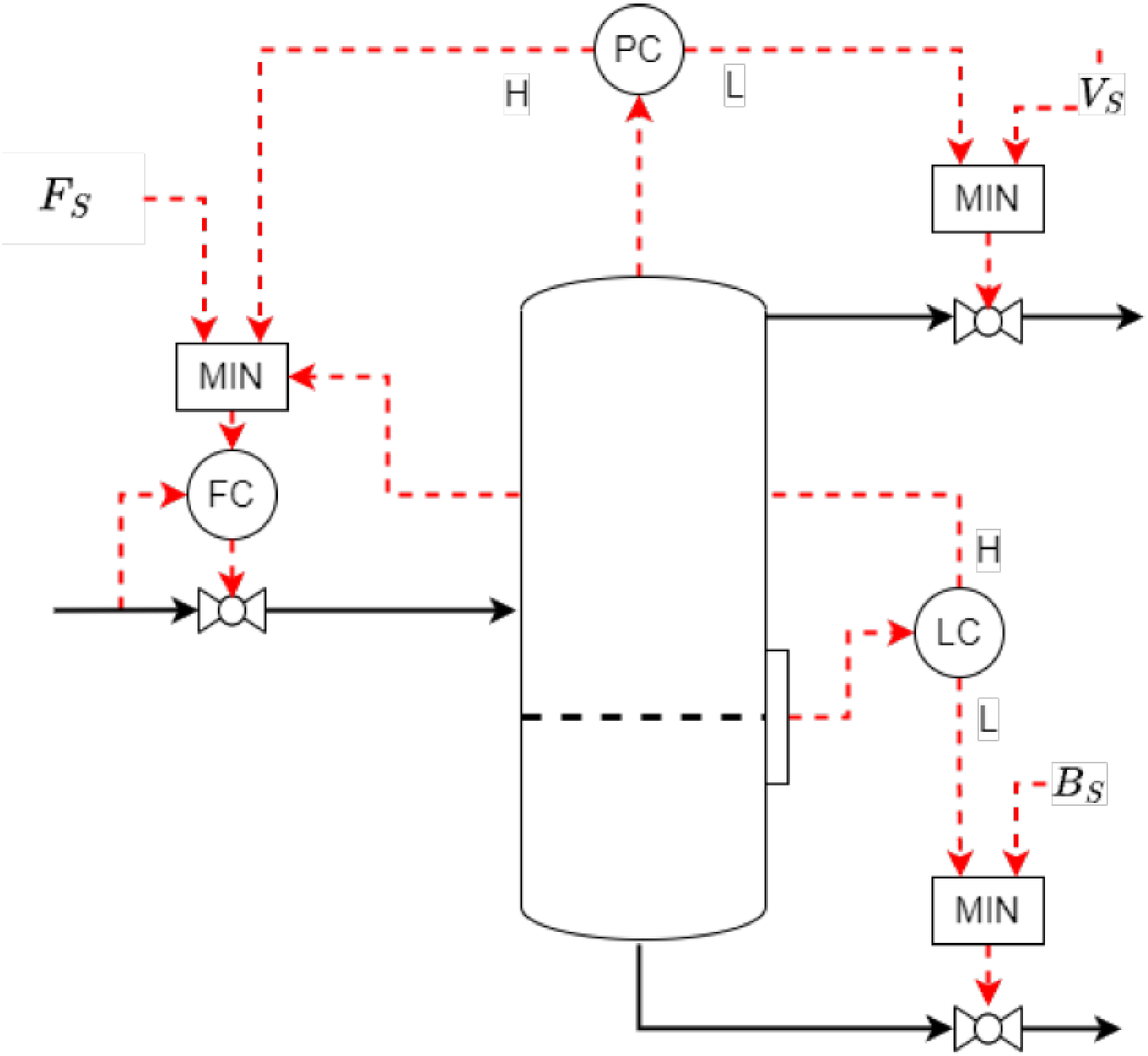
Bidirectional control of gas-liquid separator



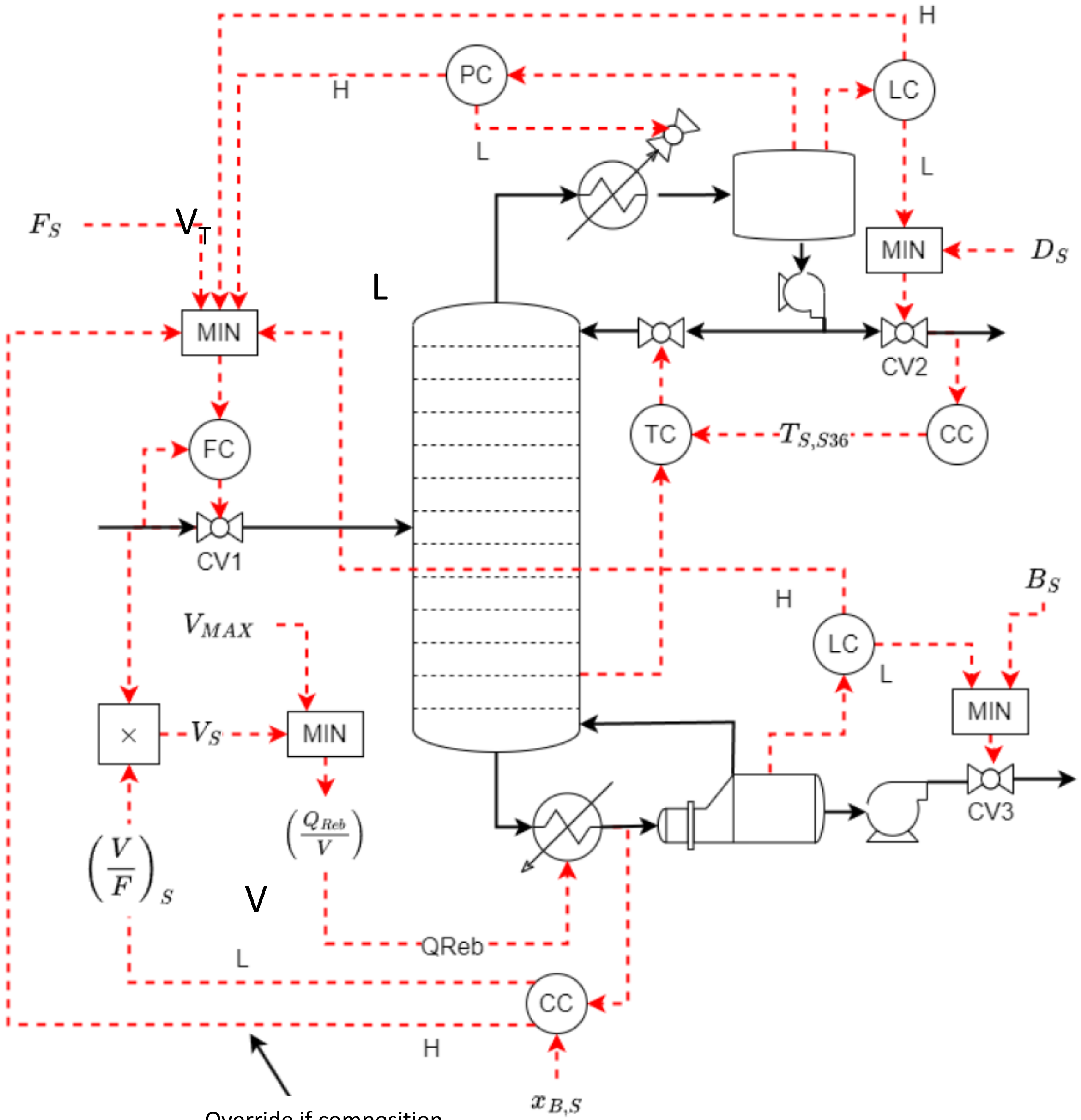
Bidirectional control of distillation

V/F with cascade in bottom
Top composition most important

4 flows may saturate at max:

- B
- D
- V_T (cooling)
- V (heating)

L (reflux) never saturates

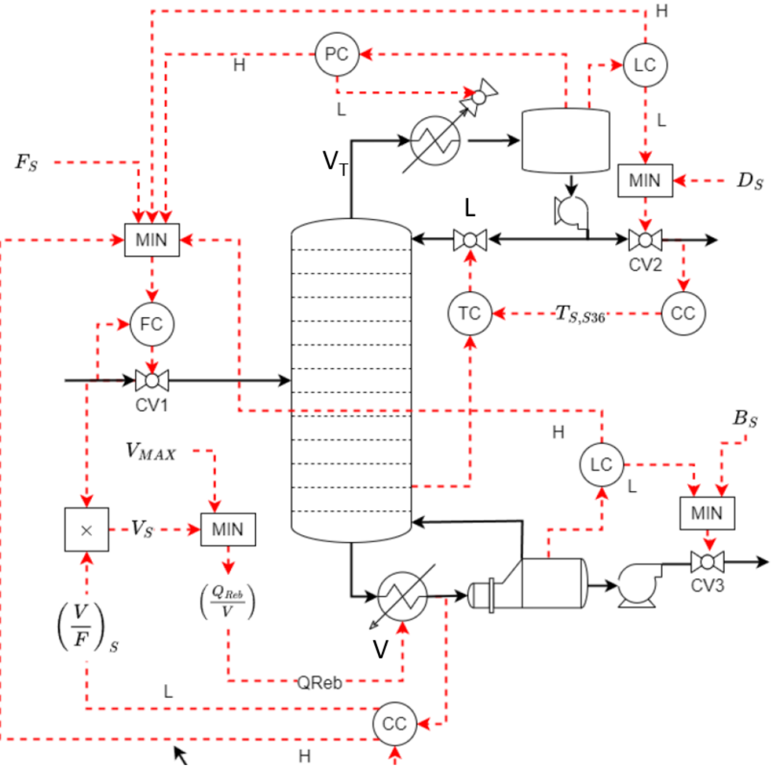


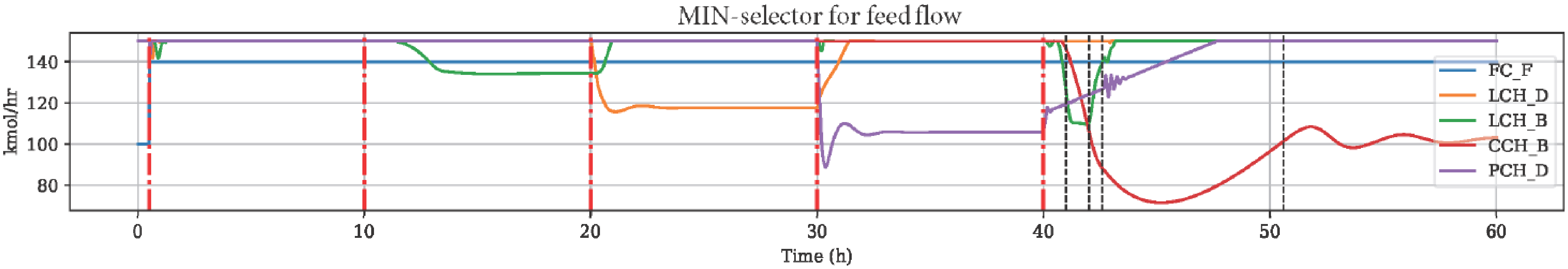
Override if composition
control is lost because
V (heat input) saturates

Dynamic simulation with Aspen

Time (h)	Constraint (TPM)	Initial value	New Limit
0.5	F_s [kmol/h]	100	140 (+40%)
10	CV3 (bottoms valve)	31.98%	22.39% (-30%)
20	CV2 (distillate valve)	44.90%	31.45% (-30%)
30	QCond [Mcal/h]		-0.963
40	QReb [Mcal/h]		1.107

Table 3. Summary of constraint limit changes that result in activating H-overrides and movement of the TPM in Figure 8.





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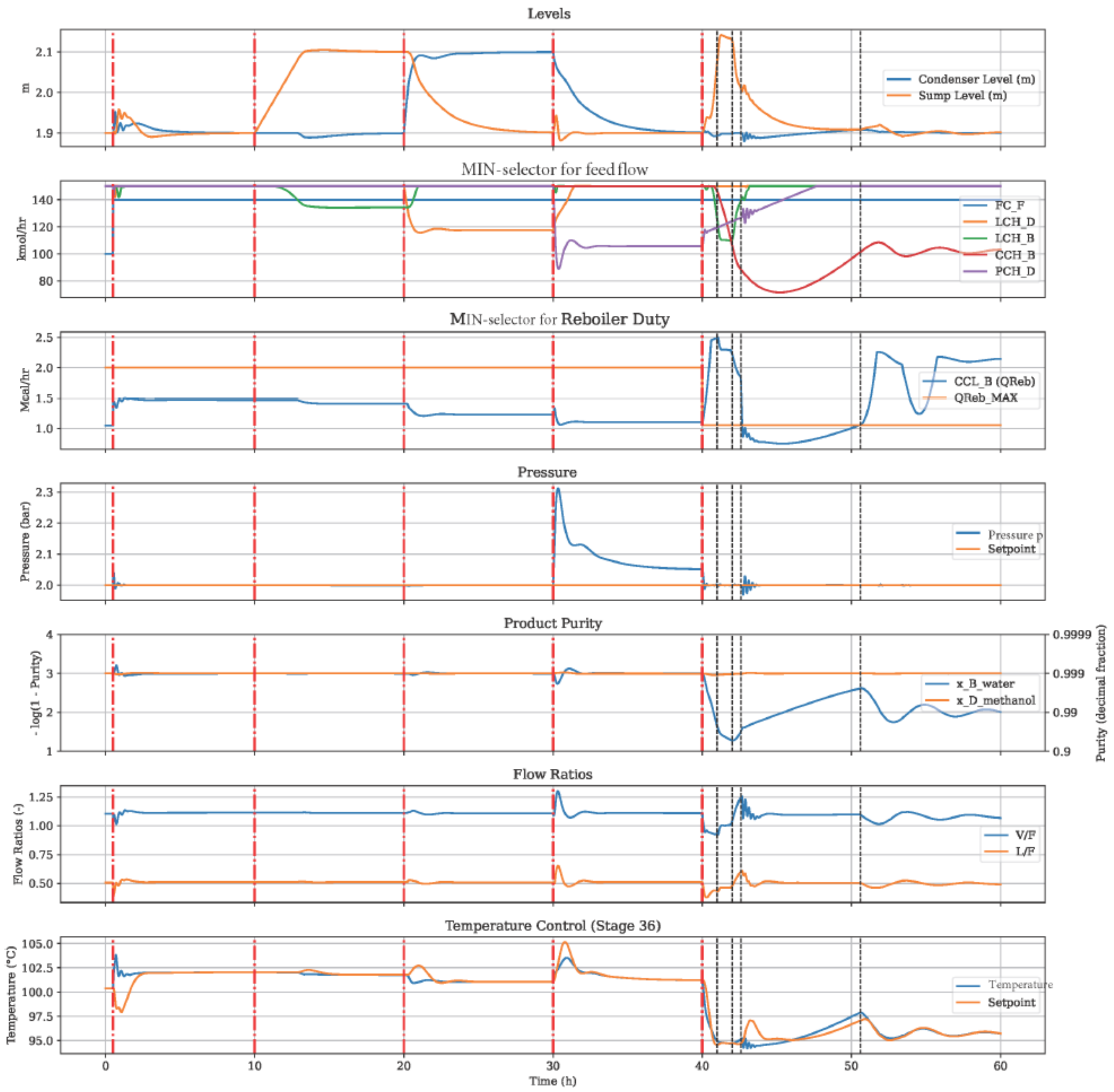

Fig. 9. Simulation results for the advanced bidirectional distillation column control scheme in Figure 8. Each vertical red line signifies one of the five events described in Table 3. Each vertical black line after t=40 h signifies the activation or deactivation of an override controller.

CONCLUSION
«Data-based» control using Advanced regulatory control

- Simple control elements put together in an archirecture
- No overall process model
- Data (measurements) is the most important

Academic process control community fish pond

Optimal centralized Solution (**EMPC**)

Simple solutions that work (**ARC** = PID++)

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Review article

Advanced control using decomposition and simple elements

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ARTICLE INFO

Keywords:
Control structure design
Feedforward control
Cascade control
PID control
Selective control
Override control
Time scale separation
Decentralized control
Distributed control
Horizontal decomposition
Hierarchical decomposition
Layered decomposition
Vertical decomposition
Network architectures

ABSTRACT

The paper explores the standard advanced control elements commonly used in industry for designing advanced control systems. These elements include cascade, ratio, feedforward, decoupling, selectors, split range, and more, collectively referred to as "advanced regulatory control" (ARC). Numerous examples are provided, with a particular focus on process control. The paper emphasizes the shortcomings of model-based optimization methods, such as model predictive control (MPC), and challenges the view that MPC can solve all control problems, while ARC solutions are outdated, ad-hoc and difficult to understand. On the contrary, decomposing the control systems into simple ARC elements is very powerful and allows for designing control systems for complex processes with only limited information. With the knowledge of the control elements presented in the paper, readers should be able to understand most industrial ARC solutions and propose alternatives and improvements. Furthermore, the paper calls for the academic community to enhance the teaching of ARC methods and prioritize research efforts in developing theory and improving design method.

Contents

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Backup slides

QUIZ
What are the three most important inventions of process control?

- Hint 1: According to Sigurd Skogestad
- Hint 2: All were in use around 1940

SOLUTION

1. PID controller, in particular, I-action
2. Cascade control
3. Ratio control

None of these are easily implemented using MPC

Cases where ratio control should <u>not</u> be used:

- Heat exchangers (because the area is fixed during operation)
  - This means, for example, that we should not fix the ratio between hot and cold flow.
  - For the scaling property to hold for a heat exchanger, we would need to increase the heat transfer area *A* proportionally to the flow rates. This is reasonable during design but not during operation.
- Non-equilibrium reactors (because the volume is fixed)
  - Extent of reaction depends on kinetics and reactor volume
- Compressors with varying thermodynamic efficiency
  - This includes most real compressors
- Mixing processes where one feed stream is fixed
- Distillation with a fixed stream (e.g. heat input)

Controller	τ_C	K_C	τ_I [s]	Setpoint
$LC_{L,D}$	*	-50	7200	1.9 m
$LC_{L,B}$	*	-50	7200	1.9 m
PC_L	*	3	28	2.0 bar
TC_{S36}	60s	5.3	2336	set by CC
CC_D	600s	-528	3600	1.e-3
$LC_{H,D}$	*	10	7200	2.1 m
$LC_{H,B}$	*	10	7200	2.1 m
PC_H	*	1	3600	2.05 bar
$CC_{L,B}$	*	7.95	7500	1.e-3
$CC_{H,B}$	*	1	600	1.e-2

Table 2. Tuning parameters for the distillation column in Figure 8. The controllers were tuned sequentially in the order given in the table. From the desired τ_C, we obtain K_C and τ_I from the SIMC rules and open-loop experiments performed in Aspen Plus. The controllers marked (*) were tuned manually based on qualitative process dynamics. Simple anti-windup schemes with bounds on the controller output are used for most PI controllers.

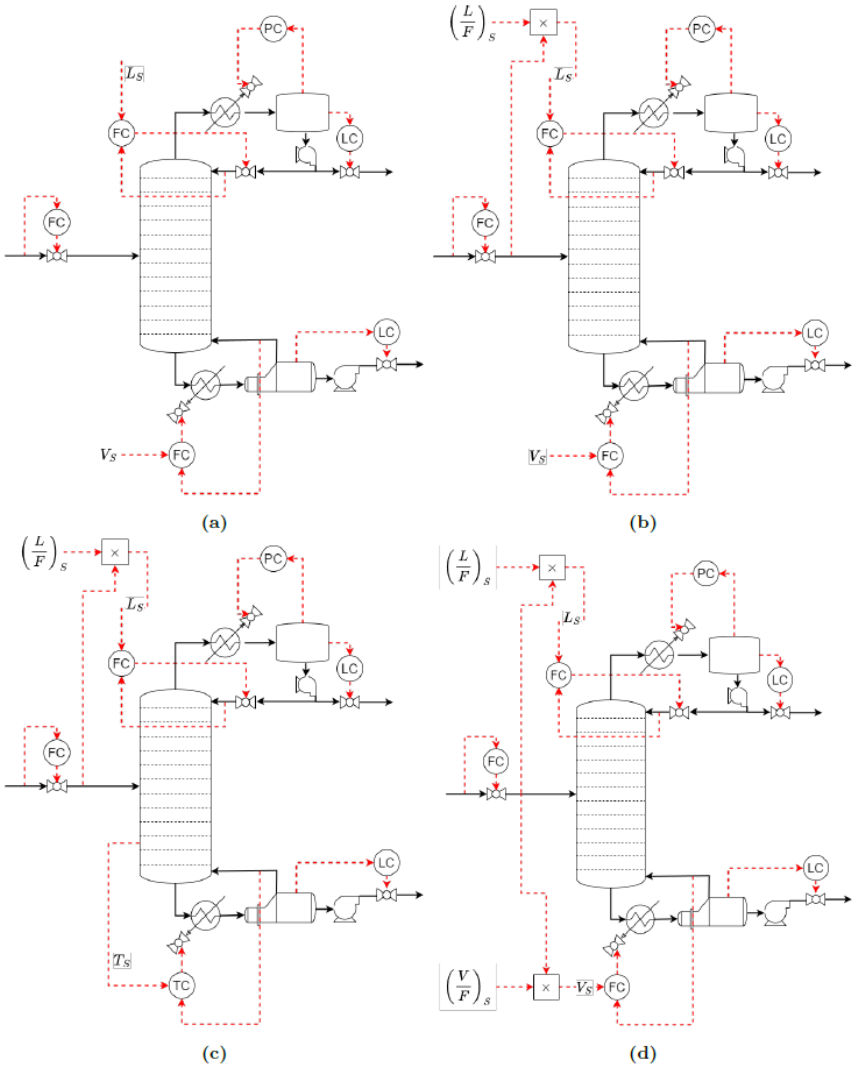


Figure 2.3 – Four "simple" distillation column control schemes.
(a) B1: L and V constant.
(b) B2: L/F and V constant.
(c) B3: L/F constant and temperature control on stage 38.
(d) B4: L/F and V/F constant.

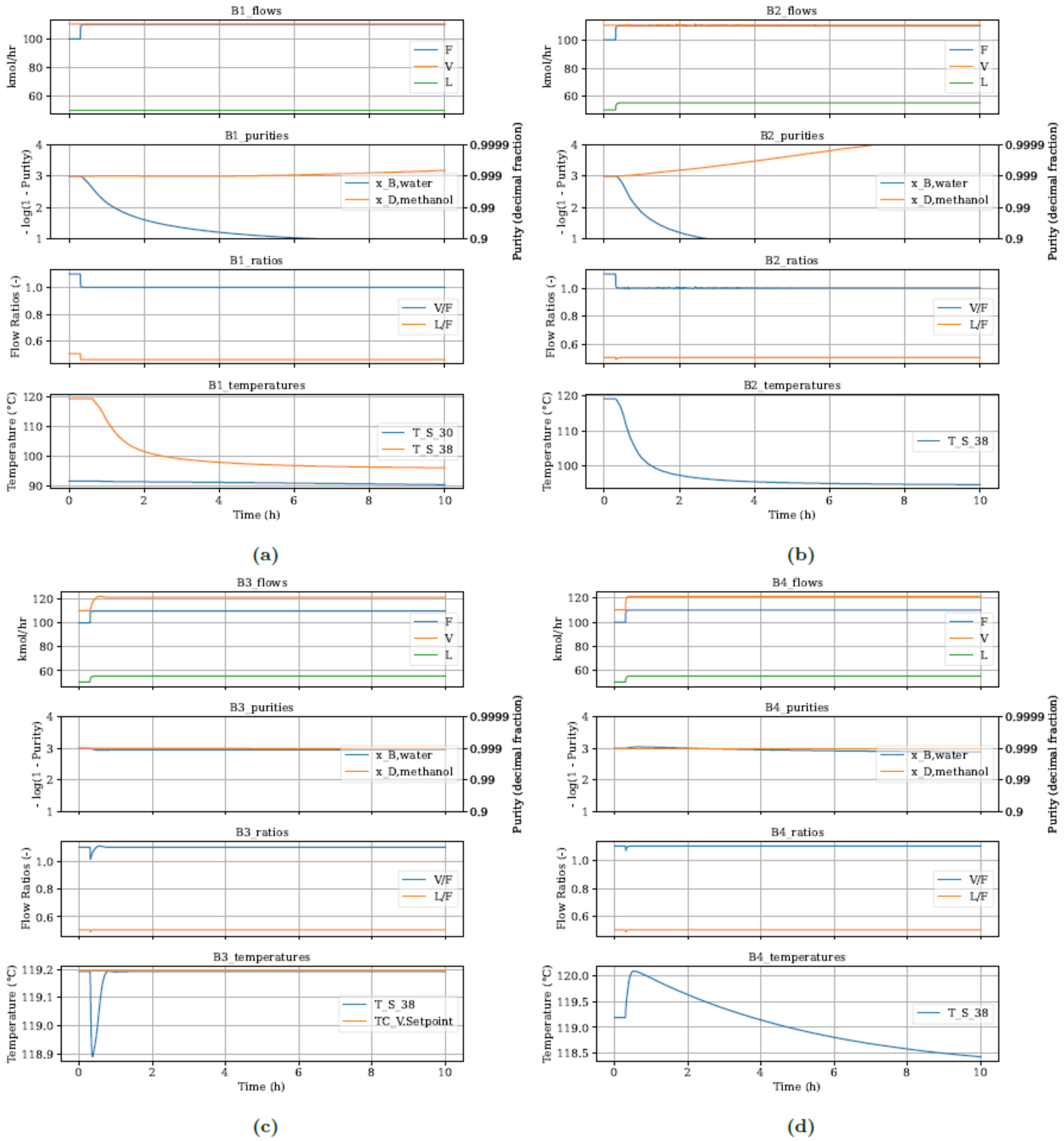
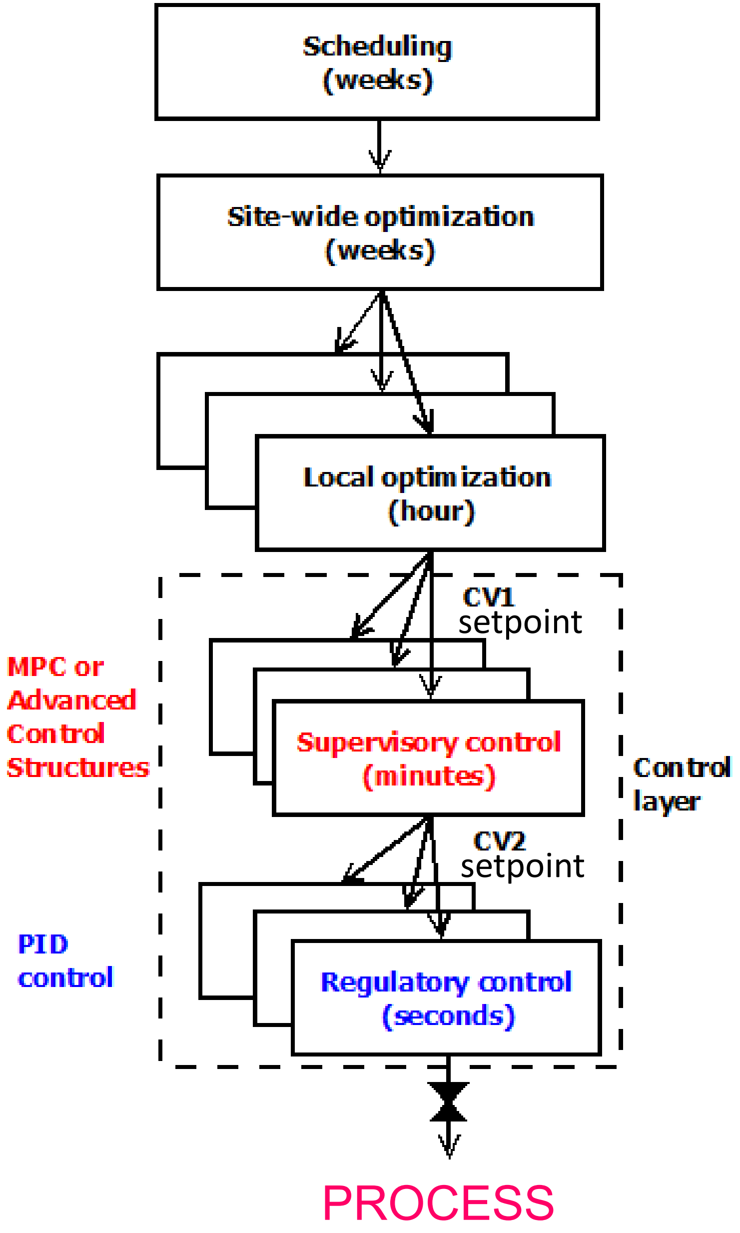


Figure 2.4 – Feed disturbance (+10%) simulation results from the schemes in Figure 2.4.
(a) B1: L and V constant.
(b) B2: L/F and V constant.
(c) B3: L/F constant and temperature control on stage 38.
(d) B4: L/F and V/F constant.

«Advanced» control

- Mainly used in the «supervisory» control layer
- Two main options
  1. **Standard «Advanced regulatory control» (ARC) elements**
     - This option is preferred if it gives acceptable performance
  2. **Model predictive control (MPC)**
     - Requires a lot more effort to implement and maintain

  – **What about machine learning (AI)?**
     - No, it requires way too much data - would take years to learn



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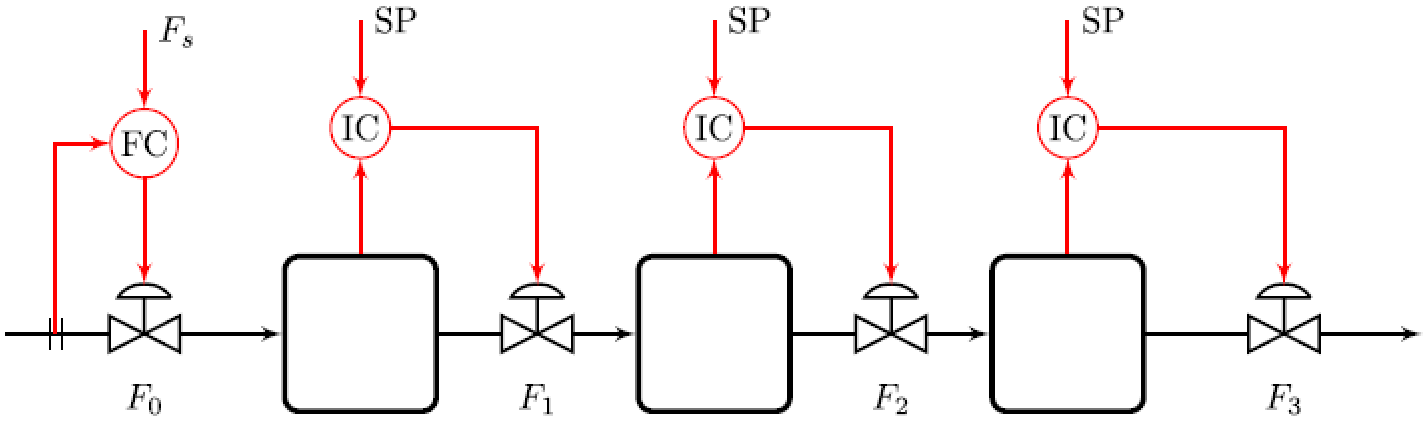
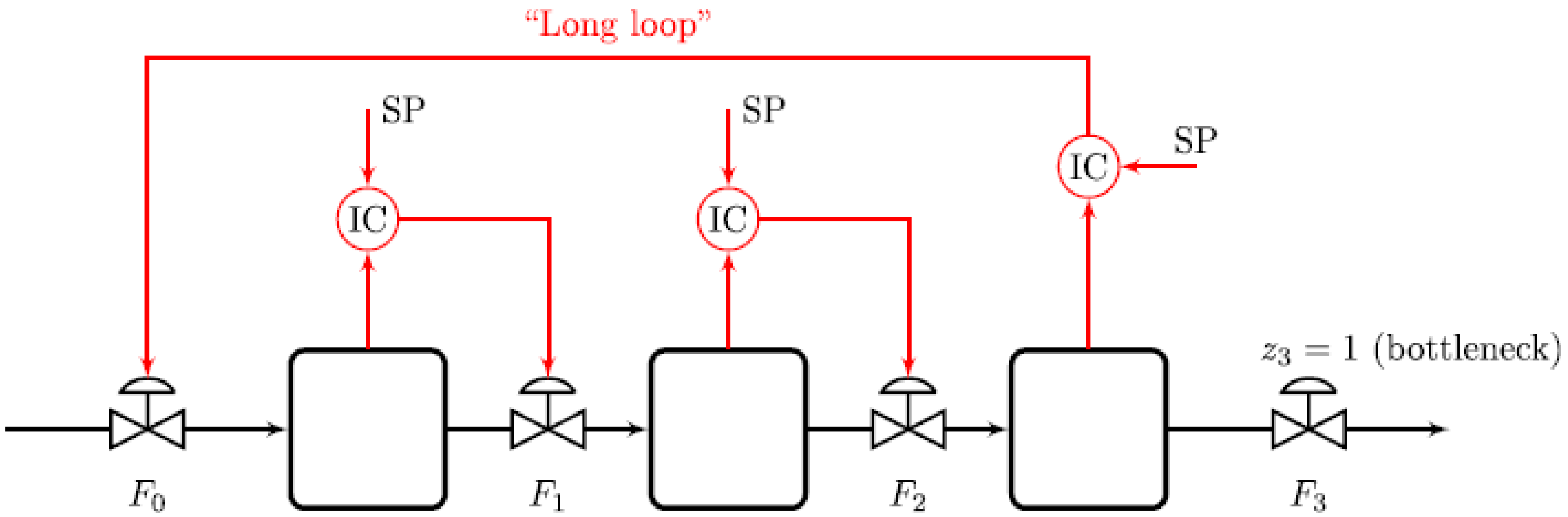
Example:
Inventory control



(a) Inventory control in direction of flow (for given feed flow, TPM = F_0)

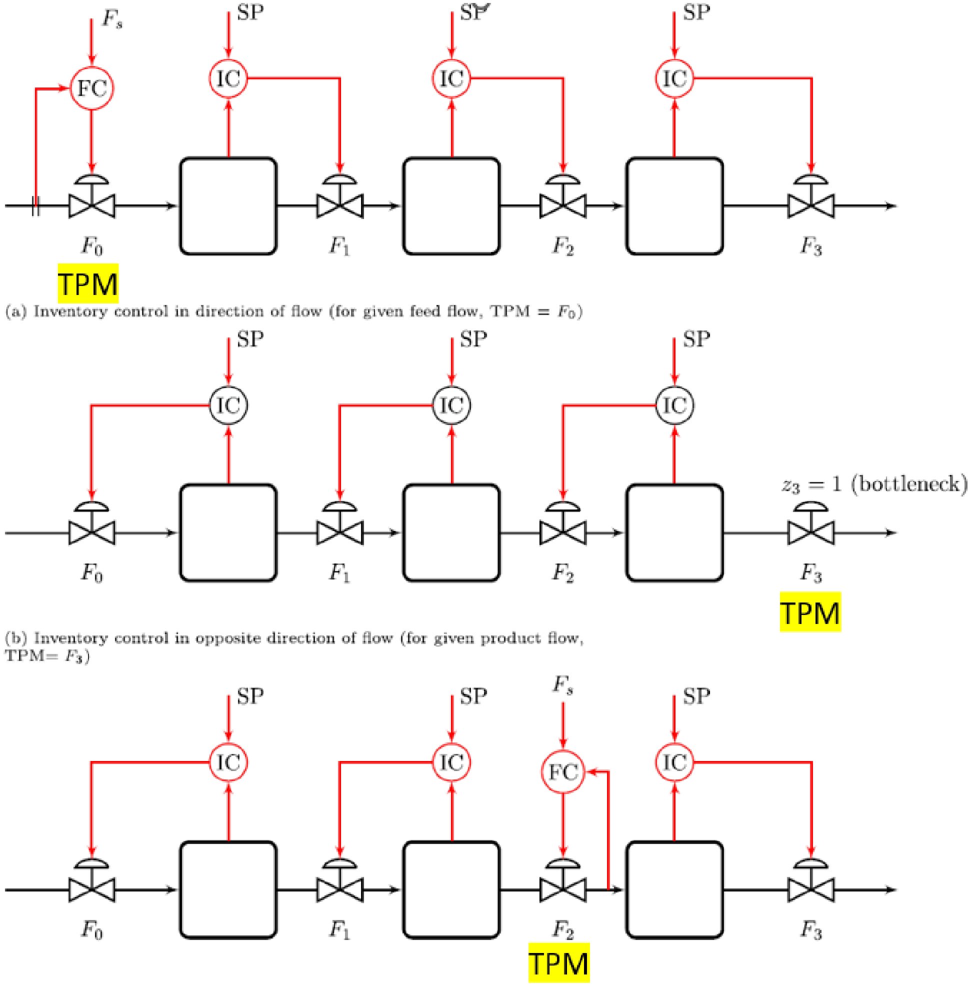


(d) Inventory control with undesired "long loop", not in accordance with the "radiation rule" (for given product flow, TPM= F_3)

TPM = throughput manipulator

Inventory control for units in series

Radiating rule:
Inventory control should be "radiating" around a given flow (TPM).



(a) Inventory control in direction of flow (for given feed flow, TPM = F_0)

(b) Inventory control in opposite direction of flow (for given product flow, TPM= F_3)

(c) Radiating inventory control for TPM in the middle of the process (shown for TPM = F_2)

Follows radiation rule

Need to reconfigure inventory loops if TPM moves

Generalization of bidirectional inventory control

Reconfigures TPM automatically with optimal buffer management!!

Maximize throughput:

$F_s=\infty$

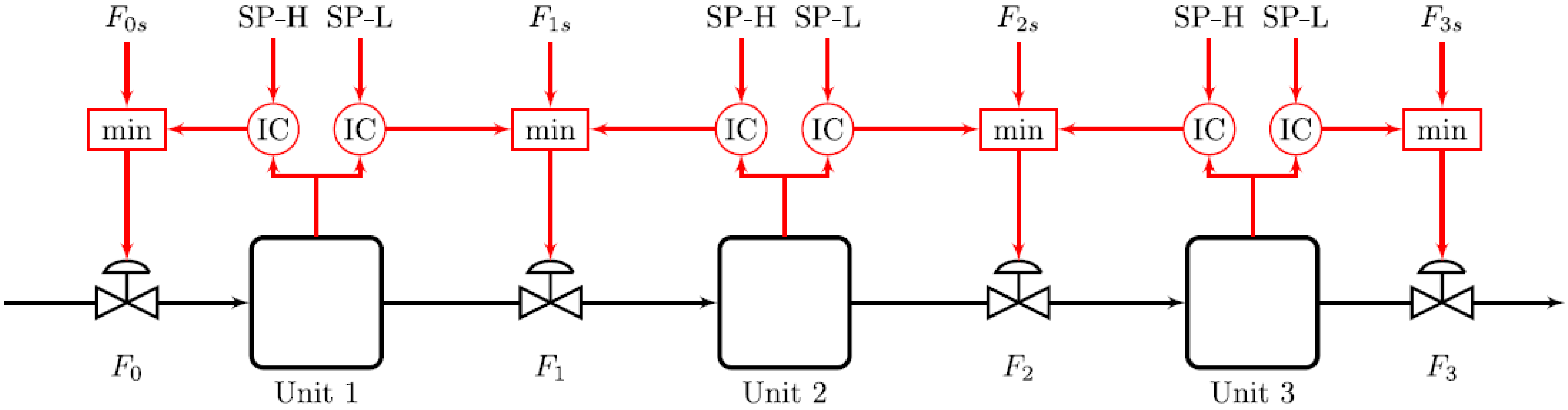


Fig. 36. Bidirectional inventory control scheme for automatic reconfiguration of loops (in accordance with the radiation rule) and maximizing throughput. Shinskey (1981) Zotica et al. (2022).

SP-H and SP-L are high and low inventory setpoints, with typical values 90% and 10%.

Strictly speaking, with setpoints on (maximum) flows ($F_{i,s}$), the four valves should have slave flow controllers (not shown). However, one may instead have setpoints on valve positions (replace $F_{i,s}$ by $z_{i,s}$), and then flow controllers are not needed.

F.G. Shinskey, «Controlling multivariable processes», ISA, 1981, Ch.3

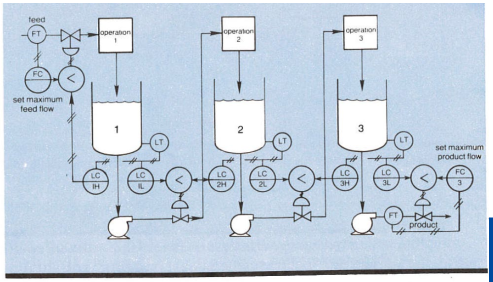


Fig. 3-7. Production rate can be set at either end of the process or constrained at any intermediate point without loss of inventory control.

When use MPC?

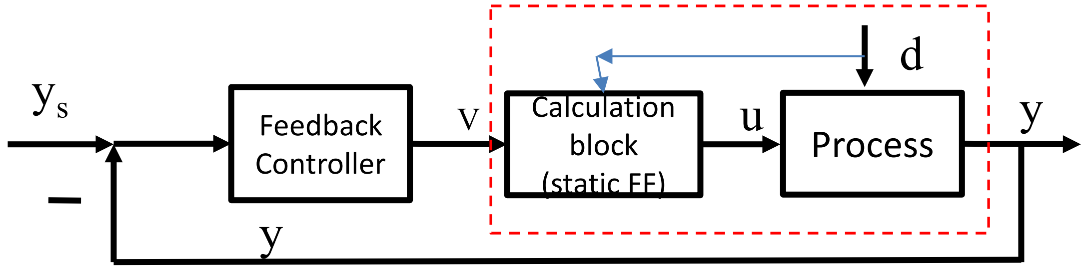
When conventional APC performs poorly or becomes complex

- Cases with many changing constraints (where we cannot assign one input to each constraint)
- Interactive process
- Know future disturbances and setpoint changes (predictive capability)

«Transformed inputs» v

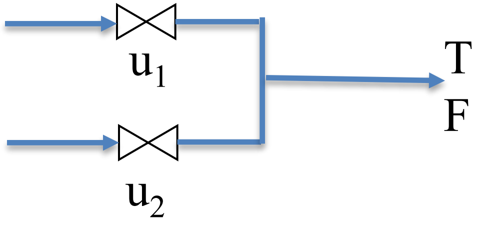
- Combining feedforward with feedback in an extremely simple way
- Most effective for static feedforward
  - Dynamic generalzation (= «feedback linearization») usually unrealistic because of many derivatives

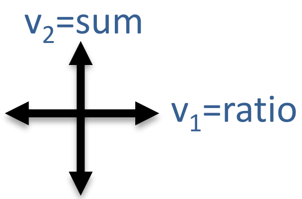
1. Static model: $y = f(u,d)$
2. Select transformed inputs (= controller outputs): $v = f(u,d)$
3. Invert to get physical inputs: $u = f^{-1}(v,d)$
4. Then response from v to y is: $y = I\,v$ (linear, decoupled, perfect disturbance rejection)



Looks like magic but ut works

Example decoupling: Mixing of hot (u_1) and cold (u_2) water





- Want to control
  - y_1 = Temperature T
  - y_2 = total flow F
- Inputs, u=flowrates
- May use two SISO PI-controllers
  - TC
  - FC
- Insight: Get decoupled response with transformed inputs
  - TC sets flow ratio, $v_1 = u_1/u_2$
  - FC sets flow sum, $v_2 = u_1 + u_2$
- Decoupler: Need «static calculation block» to solve for inputs

$$u_1 = v_1 v_2 / (1+ v_1)$$

$$u_2 = v_2 / (1 + v_1)$$

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TRANSFORMED INPUTS v=f(u,d): GENERAL APPROACH FOR COMBINED FEEDBACK, FEEDFORWARD, DECOUPLING AND LINEARIZATION

Example: Mixing of hot and cold water

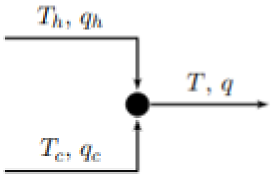
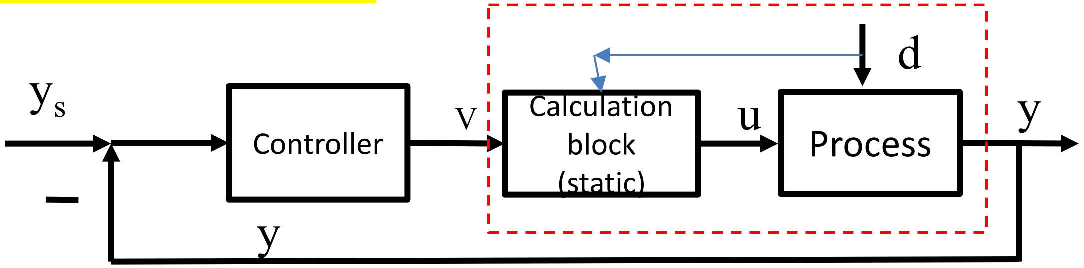


Figure 1: Mixer system



$$u = \begin{pmatrix} q_h \\ q_c \end{pmatrix} \quad d = \begin{pmatrix} T_h \\ T_c \end{pmatrix} \quad y = \begin{pmatrix} T \\ q \end{pmatrix}$$

Steady-state model written as y=f(u,d):

$$T = \frac{qhTh + qcTc}{qh + qc}$$

$$q = q_c + q_h$$

Select transformed inputs as right hand side, v = f

$$v_1 = \frac{qhTh + qcTc}{qh + qc} \quad (1)$$ Generalized ratio

$$v_2 = q_c + q_h \quad (2)$$

Model from v to y (red box) is then decoupled and with perfect disturbance rejection:

$$T = v_1$$

$$q = v_2$$

It's almost magic!

- Can then use two single-loop PI controllers for T and q!
  - These controllers are needed to correct for model errors and unmeasured disturbances
- Note that v_1 used to control T is a generalized ratio, but it includes also feedforward from Tc and Th.

Implementation (calculation block) : Solve (1) and (2) with respect to u=(qc qh):

Decoupler with feedforward: $$q_h = \frac{v_2(v_1 - T_c)}{T_h - T_c}$$

$$q_c = v_2 - q_h$$