Advanced process control for the future

using the magic of feedback and simple elements*

Sigurd Skogestad

Department of Chemical Engineering
Norwegian University of Science and Technology (NTNU)
Trondheim

11 Nov. 2024
Center of Technological Qualification in Industrial Automation (CTAI)
UFBA, Salvador

*a.k.a. Advanced regulatory control (ARC)


Midnight or midday?


Geiranger fjord


Trondheim

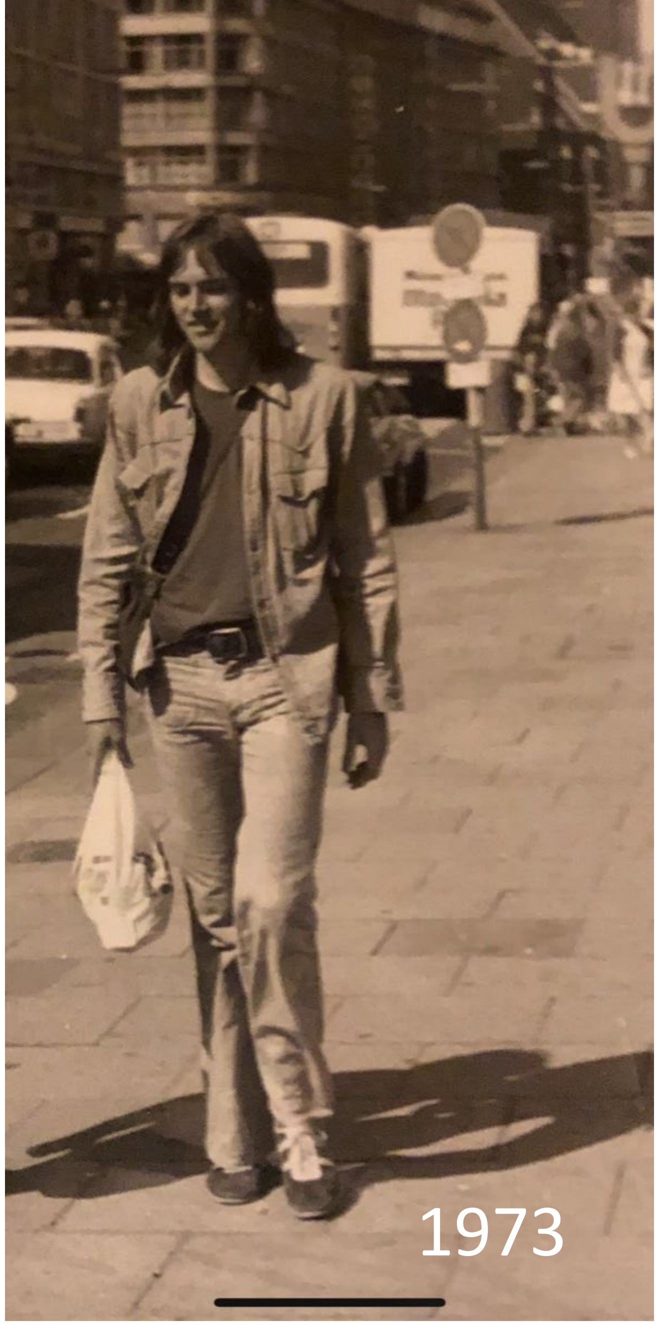

Norwegian Sea
Trondheim
Arctic circle
Sweden
Faroe Islands
Norway
Oslo
Stockholm
Baltic
North Sea
Edinburgh
Copenhagen
Denmark
United Kingdom
Isle of Man
Manchester
Dublin
Ireland
Liverpool
Hamburg
Amsterdam
Berlin
Poland
Netherlands
London
Brussels
Cologne
Germany
Belgium
Frankfurt
Prague
Luxembourg
Czechia

About Sigurd Skogestad

- 1955: Born in Flekkefjord, Norway
- 1974-1978: MS (Siv.ing.) studies in chemical engineering at NTNU
- 1979: Military service (FFI)
- 1980-1983: Worked at Norsk Hydro F-senter (process simulation)
- 1983-1987: PhD student at Caltech (supervisor: Manfred Morari)
- 1987-present: Professor of chemical engineering at NTNU
- 1994-95: Visiting Professor UC Berkeley
- 2001-02: Visiting Professor UC Santa Barbara
- 1999-2009: Head of ChE Department, NTNU
- 2015-..: Director SUBPRO (Subsea research center at NTNU)

Non-professional interests:

- **mountain skiing (cross country)**
- **orienteering (running around with a map)**
- **grouse hunting**



NTNU

“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

Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), N7491 Trondheim, Norway

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).

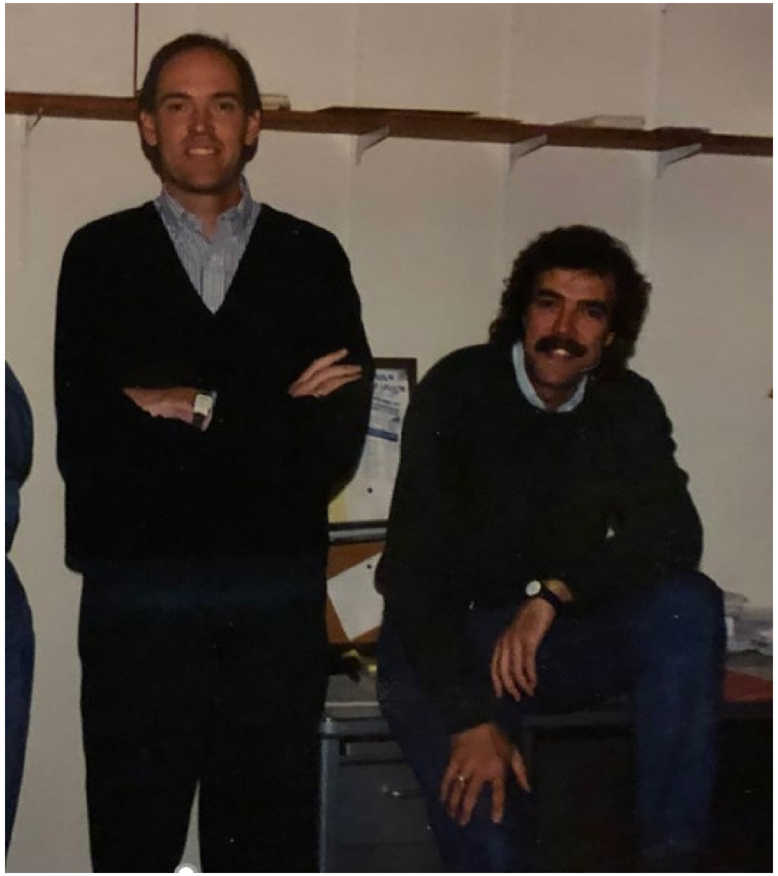
More information ...

- **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.

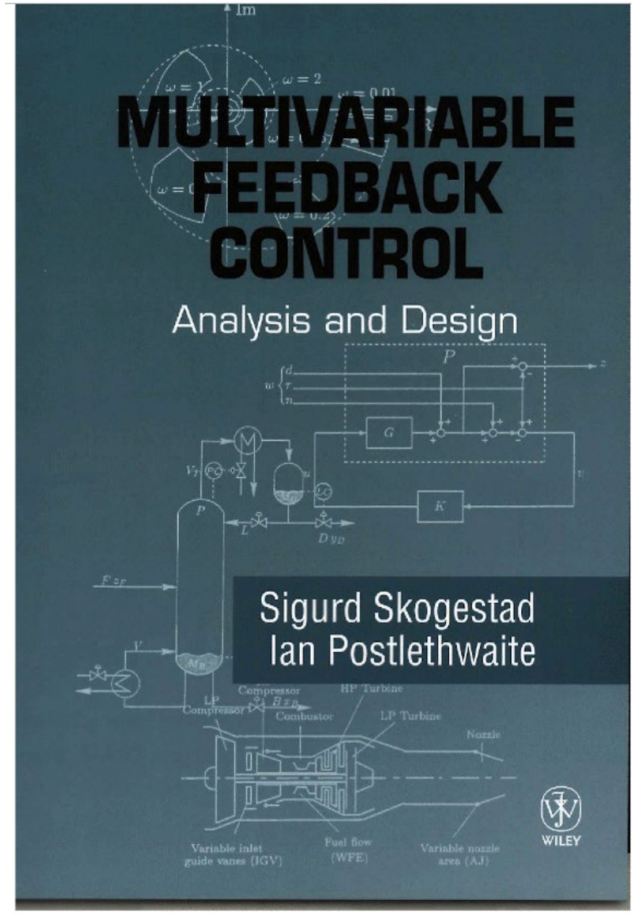
My research focus

- Control for economic optimization
  - Control of changing active constraints
- Control for linearization, stabilization and robustness
- Keep it simple!
  - Make use of the magic of feedback

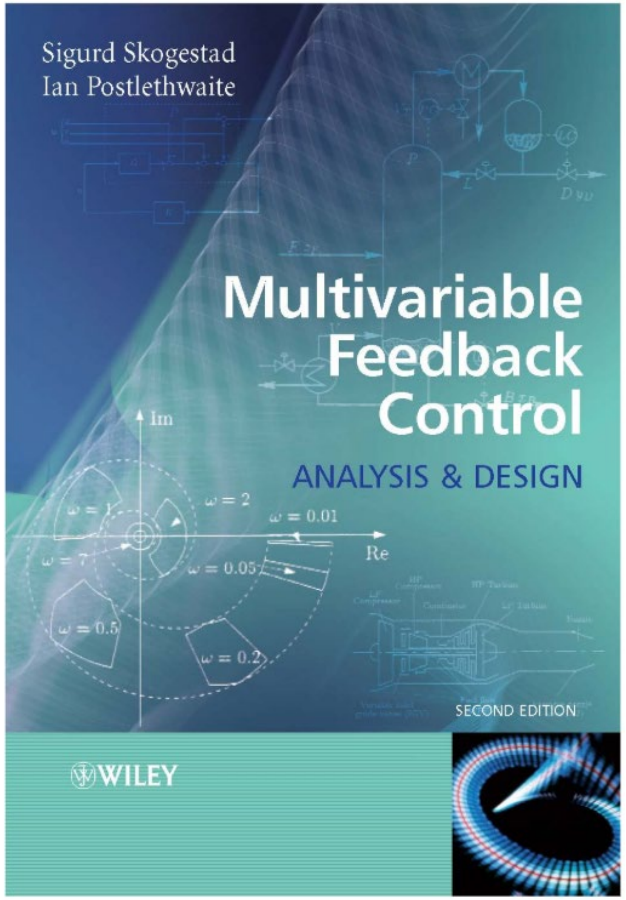
Robust control



Berkeley, Dec. 1994



1996



2005

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.

IMC PID tuning rule (1984, 1986)

AMERICAN CONTROL CONFERENCE
San Diego, California
June 6-8, 1984

IMPLICATIONS OF INTERNAL MODEL CONTROL FOR PID CONTROLLERS

Manfred Morari
Sigurd Skogestad

California Institute of Technology
Department of Chemical Engineering
Pasadena, California 91125

Daniel F. Rivera

University of Wisconsin
Department of Chemical Engineering
Madison, Wisconsin 53706

252 Ind. Eng. Chem. Process Des. Dev. **1986**, *25*, 252–265

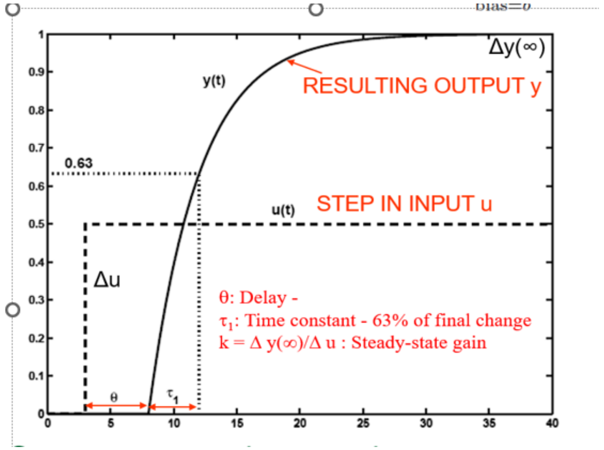
Internal Model Control. 4. PID Controller Design

Daniel E. Rivera, Manfred Morari,* and Sigurd Skogestad

Chemical Engineering, 206-41, California Institute of Technology, Pasadena, California 91125

For a large number of single input–single output (SISO) models typically used in the process industries, the Internal Model Control (IMC) design procedure is shown to lead to PID controllers, occasionally augmented with a first-order lag. These PID controllers have as their only tuning parameter the closed-loop time constant or, equivalently, the closed-loop bandwidth. On-line adjustments are therefore much simpler than for general PID controllers. As a special case, PI- and PID-tuning rules for systems modeled by a first-order lag with dead time are derived analytically. The superiority of these rules in terms of both closed-loop performance and robustness is demonstrated.

SIMC* PID tuning rule (2001,2003)



$$g(s) = \frac{k}{(\tau_1 s + 1)(\tau_2 s + 1)} e^{-\theta s}$$

$$K_c = \frac{1}{k}\frac{\tau_1}{\tau_c + \theta}$$

$$\tau_I = \min\{\tau_1, 4(\tau_c + \theta)\}$$

$$\tau_D = \tau_2$$

Tuning parameter:

$\tau_c \geq \theta$

$=\lambda$

[19] S. Skogestad, Probably the best simple PID tuning rules in the world. AIChE Annual Meeting, Reno, Nevada, November 2001

ELSEVIER

Journal of Process Control 13 (2003) 291–309

JOURNAL OF PROCESS CONTROL

www.elsevier.com/locate/jprocont

Simple analytic rules for model reduction and PID controller tuning☆

Sigurd Skogestad*

Department of Chemical Engineering, Norwegian University of Science and Technology, N-7491 Trondheim, Norway

Received 18 December 2001; received in revised form 25 June 2002; accepted 11 July 2002

Abstract

The aim of this paper is to present analytic rules for PID controller tuning that are simple and still result in good closed-loop behavior. The starting point has been the IMC-PID tuning rules that have achieved widespread industrial acceptance. The rule for the integral term has been modified to improve disturbance rejection for integrating processes. Furthermore, rather than deriving separate rules for each transfer function model, there is a just a single tuning rule for a first-order or second-order time delay model. Simple analytic rules for model reduction are presented to obtain a model in this form, including the "half rule" for obtaining the effective time delay.

*SIMC = Simple/Skogestad IMC

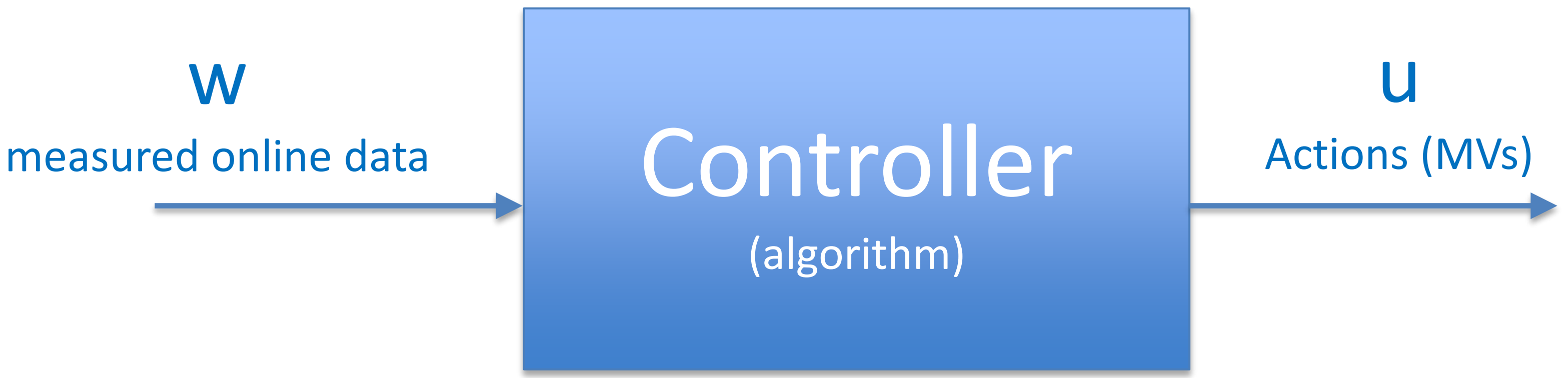
NTNU

Congratulations on the 20 years anniversary!

- Not so long….

- But a good oppurtunity to look a bit at
  - What are we really trying to do?
  - Can we learn from history?
  - What is the future of process control

What is control?
Generating actions u from measurements w (data)



Want: $y = y_s$

1. Feedback: w=y (with setpoint y_s) depends on u
   - P-control (negative feedback): $u = K_c(y_s - y)$
   - Good control ($y \approx y_s$): Need gain $K_c \gg 1$
   - Don't need accurate value for K_c (but too large gives instability)

2. Feedforward: w=d or y_s , independent of u
   - Perfect feedforward: $u = G^{-1}\, y_s - G^{-1} G_d\, d$
   - Good control ($y \approx y_s$): Need accurate model (G)

y = CV (controlled variable)
u = MV (manipulated varoable)

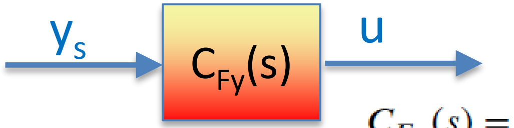
Feedforward versus feedback

Example: Setpoint response

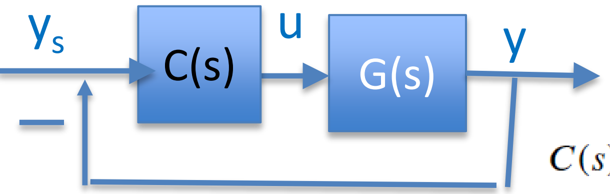
$$G(s) = \frac{k}{\tau s + 1}, \quad k = 3,\ \tau = 6$$

$$\text{Desired response :} \quad y = \frac{1}{\tau_c s + 1} y_s = \frac{1}{4s + 1} y_s$$

This can be achieved both with feedback (PI) and feedforward (nominally):



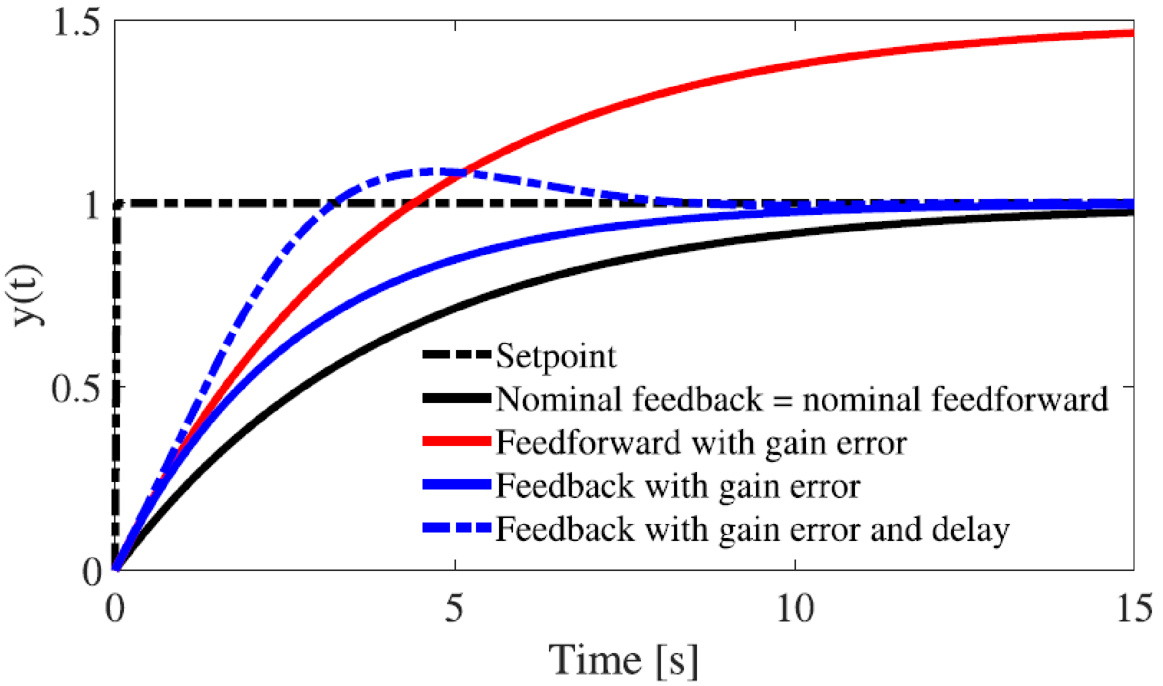
$$C_{Fy}(s) = \frac{1}{\tau_c s + 1} G(s)^{-1} = \frac{1}{k} \frac{\tau s + 1}{\tau_c s + 1} = \frac{1}{3} \frac{6s + 1}{4s + 1}$$



$$C(s) = K_c \left(1 + \frac{1}{\tau_I s}\right) = 0.5 \frac{6s + 1}{6s}$$

Identical nominally

Change process gain from k=3 to k'=4.5



Feedforward is sensitive to model error

With MPC: Not clear which solution you get

NTNU

Feedback is needed / used for

- Uncertainty about disturbances
- Uncertainty in model
- Linearization
- Stabilization

NOTE

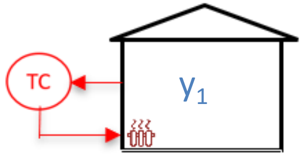
- «That something works doesn't mean that it couldn't be much better or simpler (PID), or even both better and simpler at the same time».
- Example: Sensor in wrong room

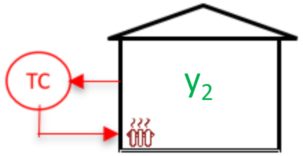
MULTIVARIABLE FEEDBACK CONTROL
Analysis and Design

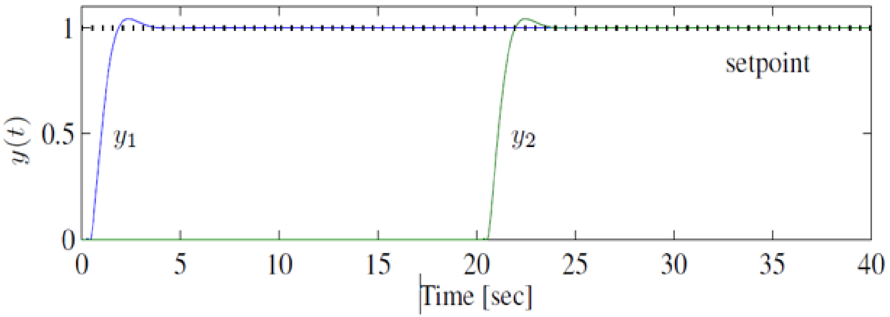
Second Edition

Sigurd Skogestad
Norwegian University of Science and Technology

Ian Postlethwaite
University of Leicester



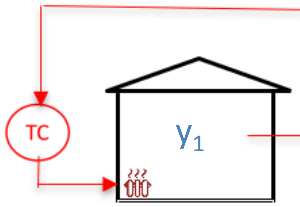


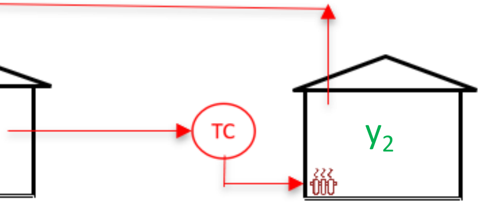


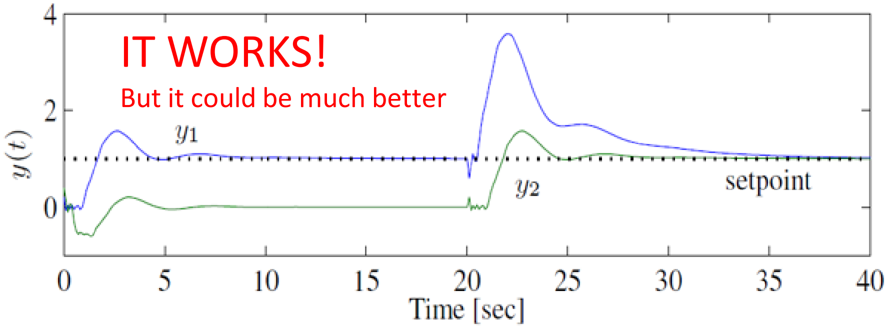
(a) Diagonal pairing; controller (10.51) with $\tau_1 = \tau_2 = 1$

$$G = \begin{bmatrix} g_{11} & g_{12} \\ g_{21} & g_{22} \end{bmatrix} = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

$$K = \begin{bmatrix} \frac{1}{\tau_1 s} & 0 \\ 0 & \frac{1}{\tau_2 s} \end{bmatrix}$$







(b) Off-diagonal pairing; plant (10.53) and controller (10.54)

$$G = \begin{bmatrix} g_{12} & g_{11} \\ g_{22} & g_{21} \end{bmatrix} = \begin{bmatrix} 0 & 1 \\ 1 & 0 \end{bmatrix}$$

$$K^*(s) = \begin{bmatrix} \frac{-(0.5s+0.1)}{s} & 0 \\ 0 & \frac{(0.5s+2)}{s} \end{bmatrix}$$

$$G_{\text{sim}} = Ge^{-0.5s}.$$

y = temperature = output = controlled variable

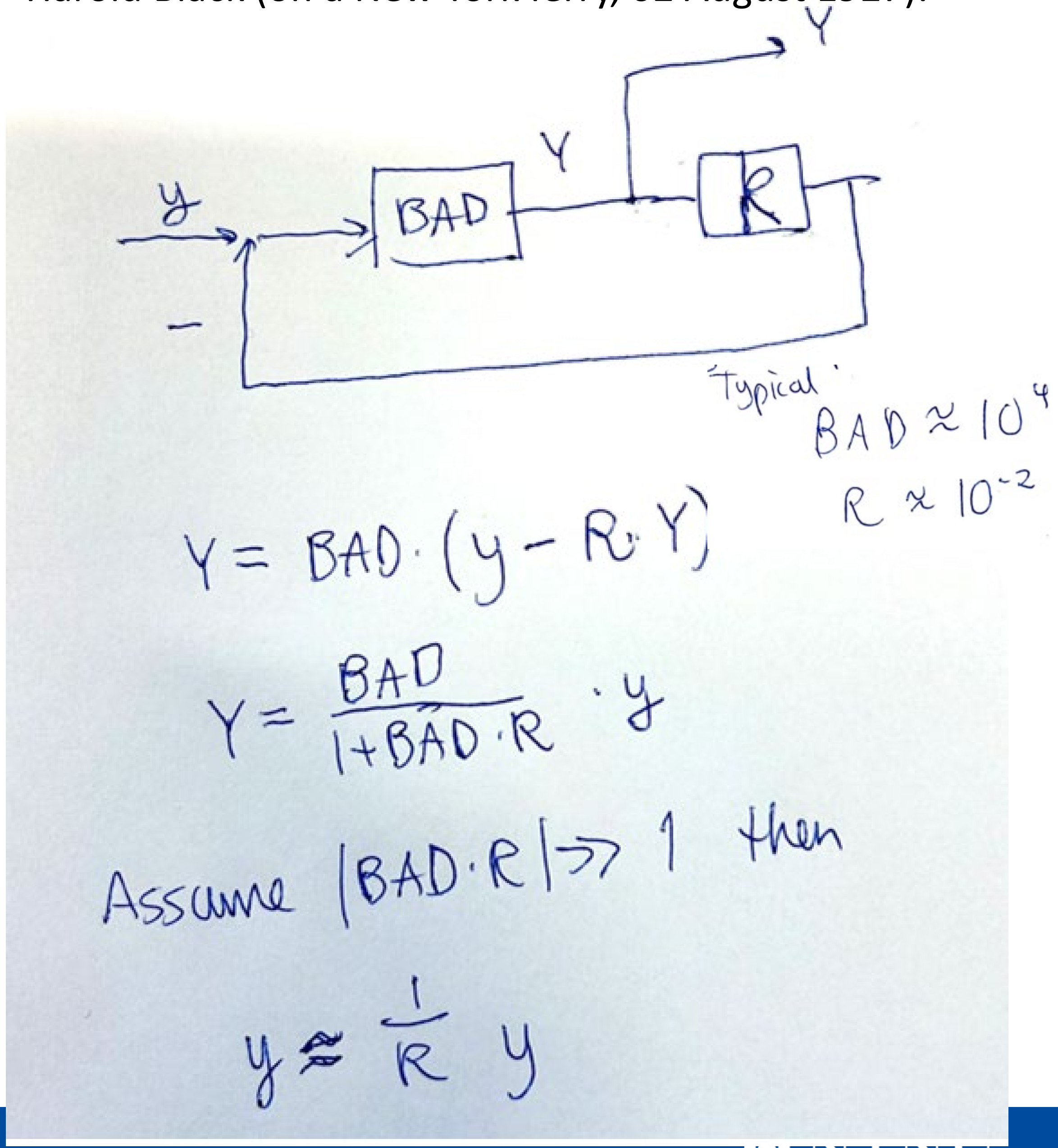
Linearizing effect of feedback

The negative feedback amplifier

- Amplification was required to send telephone signals across the US in the 1920s and it required 12 amplifications on the way, so they better be fairly accurate.
- The original idea of all engineers is to think feedforward (Bell Labs)
- Y = BAD y

He submitted an extremely long application (52 pages, 126 claims) in 1928, but the patent office objected to many of the claims, apparently because his concept of negative feedback flew in the face of accepted theory. The examiners finally awarded the patent nine years later, in December 1937 [10], after Black and others at AT&T developed both a practical amplifier and a theory of negative feedback.

Harold Black (on a New York ferry, 02 August 1927):



The Bell System Technical Journal

January, 1934

Stabilized Feedback Amplifiers*

By H. S. BLACK

This paper describes and explains the theory of the feedback principle and then demonstrates how stability of amplification and reduction of modulation products, as well as certain other advantages, follow when stabilized feedback is applied to an amplifier. The underlying principle of design by means of which singing is avoided is next set forth. The paper concludes with some examples of results obtained on amplifiers which have been built employing this new principle.

The carrier-in-cable system dealt with in a companion paper [1] involves many amplifiers in tandem with many telephone channels passing through each amplifier and constitutes, therefore, an ideal field for application of this feedback principle. A field trial of this system was made at Morristown, New Jersey, in which seventy of these amplifiers were operated in tandem. The results of this trial were highly satisfactory and demonstrated conclusively the correctness of the theory and the practicability of its commercial application.

Conclusion

The feedback amplifier dealt with in this paper was developed primarily with requirements in mind for a cable carrier telephone system, involving many amplifiers in tandem with many telephone channels passing through each amplifier. Most of the examples of feedback amplifier performance have naturally been drawn from amplifiers designed for this field of operation. In this field, vacuum tube amplifiers normally possessing good characteristics with respect to stability and freedom from distortion are made to possess superlatively good characteristics by application of the feedback principle.

However, certain types of amplifiers in which economy has been secured by sacrificing performance characteristics, particularly as regards distortion, can be made to possess improved characteristics by the application of feedback. Discussion of these amplifiers is beyond the scope of this paper.

* Presented at Winter Convention of A. I. E. E., New York City, Jan. 23–26, 1934. Published in *Electrical Engineering*, January, 1934.

[1] "Carrier in Cable" by A. B. Clark and B. W. Kendall, presented at the A. I. E. E. Summer Convention, Chicago, Ill., June, 1933; published in *Electrical Engineering*, July, 1933, and in *Bell Sys. Tech. Jour.*, July, 1933.

1

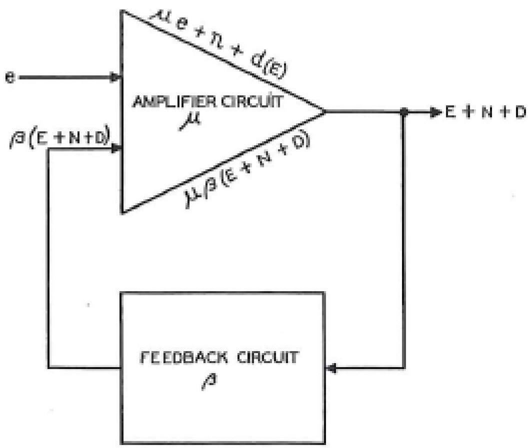


Fig. 1—Amplifier system with feedback.

e—Signal input voltage.
μ—Propagation of amplifier circuit.
μe—Signal output voltage without feedback.
n—Noise output voltage without feedback.
$d(E)$—Distortion output voltage without feedback.
β—Propagation of feedback circuit.
E—Signal output voltage with feedback.
N—Noise output voltage with feedback.
D—Distortion output voltage with feedback.

The output voltage with feedback is $E + N + D$ and is the sum of $\mu e + n + d(E)$, the value without feedback plus $\mu\beta[E + N + D]$ due to feedback.

$$E + N + D = \mu e + n + d(E) + \mu\beta[E + N + D]$$

$$[E + N + D](1 - \mu\beta) = \mu e + n + d(E)$$

$$E + N + D = \frac{\mu e}{1 - \mu\beta} + \frac{n}{1 - \mu\beta} + \frac{d(E)}{1 - \mu\beta}$$

If $|\mu\beta| \gg 1$, $E \doteq -\frac{e}{\beta}$. Under this condition the amplification is independent of μ but does depend upon β. Consequently the over-all characteristic will be controlled by the feedback circuit which may include equalizers or other corrective networks.

Experiment

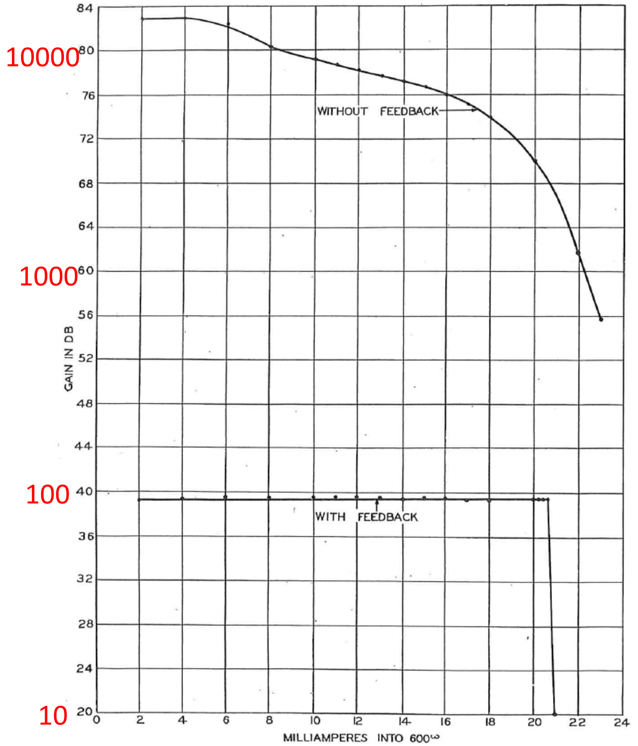


Fig. 10—Gain-load characteristic with and without feedback for a low level amplifier designed to amplify frequencies from 3.5 to 50 kc.

Bad amplifier:

$$\mu \approx 10000$$

Accurate resistance:

$$\beta \approx 0.01$$

Get

$$\mu\beta \approx 100 \gg 1$$

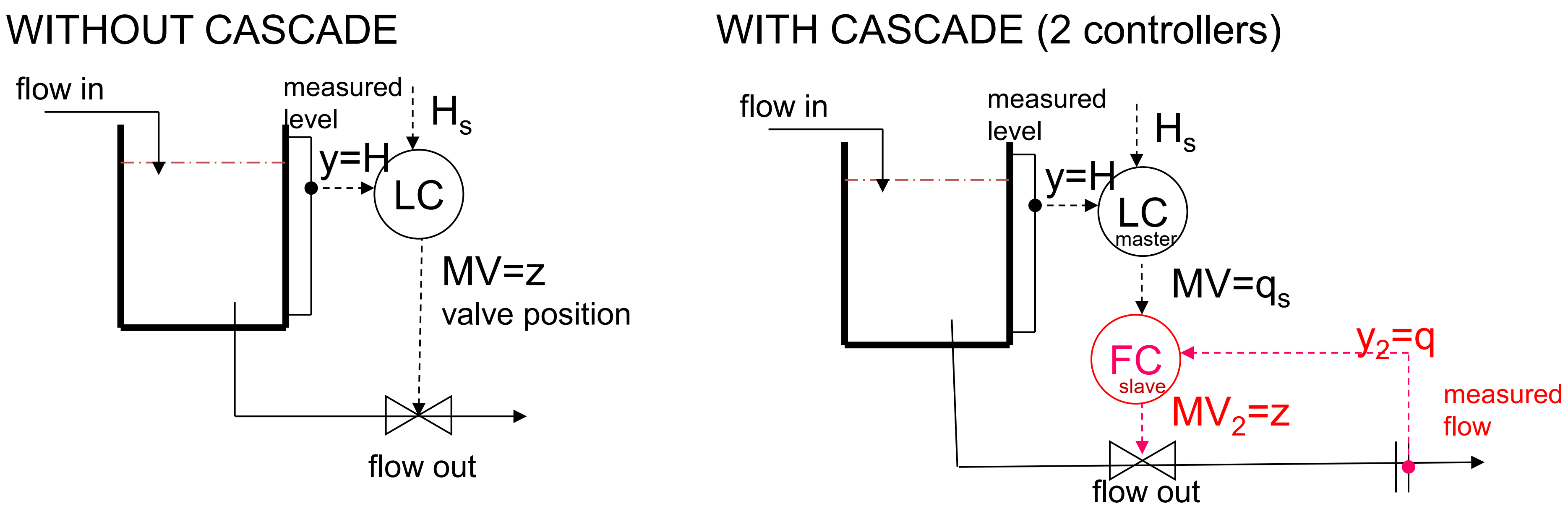
Resulting amplification (negative feedback)

$$\frac{\mu}{1+\mu\beta} \approx \frac{1}{\beta} = 100$$

Proving that Black's invention worked also initiated the idea of freqyency analysis(Nyquist)

Linearization of valve using cascade control

- Benefits: 1. Local distrurbance rejection, 2. Linearization
- Does nonlinearity disappear?



NTNU

No, it moves to the time constant for slave loop – OK if we we have time scale separation between master and slave

Nonlinear valve with varying gain k_2: $G_2(s) = k_2(z) / (\tau_2 s+1)$

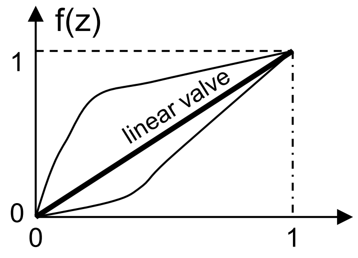
- Slave (flow) controller K_2: PI-controller with gain K_{c2} and integral time $\tau_I = \tau_2$ (SIMC-rule). Get

$$L_2 = K_2(s)G_2(s) = \frac{K_{c2}k_2}{\tau_2 s}$$

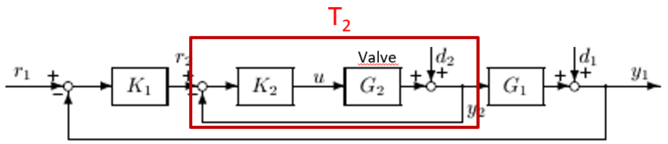
- With slave controller: Transfer function T_2 from y_{2s} to y_2 (as seen from master loop):

$$T_2 = L_2/(1+L_2) = 1/(\tau_{C2}\, s + 1), \text{ where } \tau_{C2} = \tau_2/(k_2 K_{c2})$$

- Linearization: Gain for T_2 is always 1 (independent of k_2) because of intergal action in the inner (slave) loop
- But: Gain variation in k_2 (inner loop) translates into variation in closed-loop time constant τ_{C2}. This may effect the master loop



$k_2(z)$ = slope = df/dz



G_1T_2 = «Process» for tuning master controller K_1

Stabilization

- The only way we can change dynamics (poles) and stabilize is by feedback
- So: Stabilization with feedforward does NOT work
  - Example: level control
    - $G(s) = k'/s$
    - Integrating process with pole at s=0. At the limit to instability
    - It is practically impossible to control level by trying to set $q_{out}=q_{in}$ using feedforward.
    - We get «internal instability»: Level will eventually go out of bound
- But we need to be careful: Feedback often causes oscillations and even instability

Stabilization of the grid using P-control. Early 1930s

- Initial idea (not workabler): Centralized coordination of power producers
- 1930s: Use grid frequency («level») as (local) measure of imbalance between supply and demand. And: Stabilize frequency using local P-control for all power producers
- It's the same as level control with many flows in and out
- Need to have some back-off from maximum power for this to work (90%)
- Many local P-controllers, but only one centralized I-controller

C. Zotică, L.O. Nord and J. Kovács et al./Computers and Chemical Engineering 141 (2020) 106995

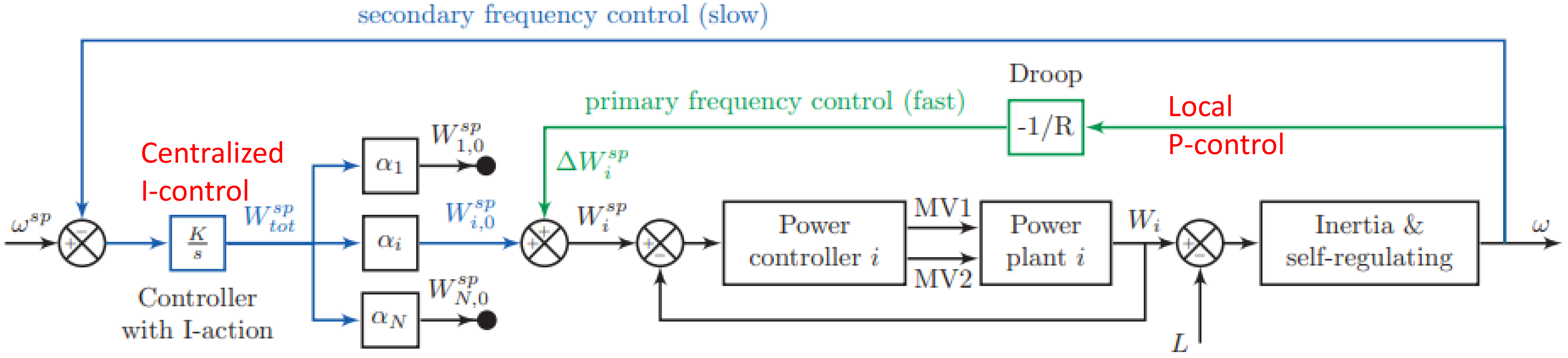
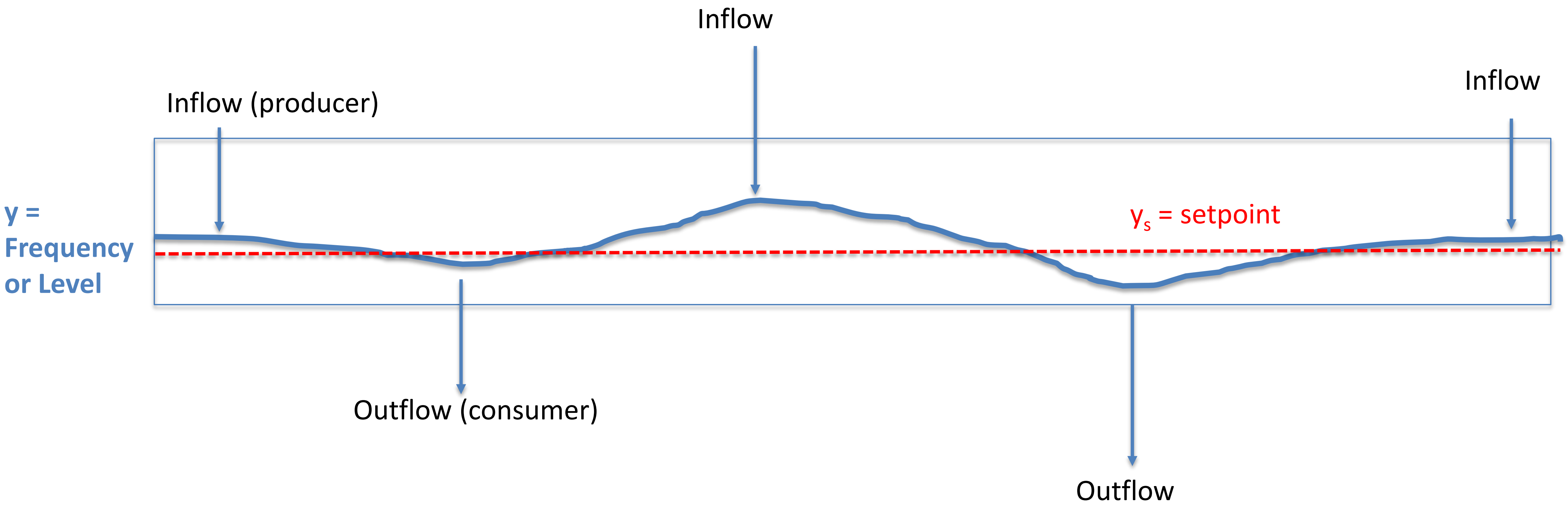


Fig. 15. Primary (green) and secondary (blue) frequency control for power plant i in an area with N power plants participating in grid frequency control (adapted from Wood et al., 2014.). (For interpretation of the references to colour in this figure legend, the reader is referred to the web version of this article.)

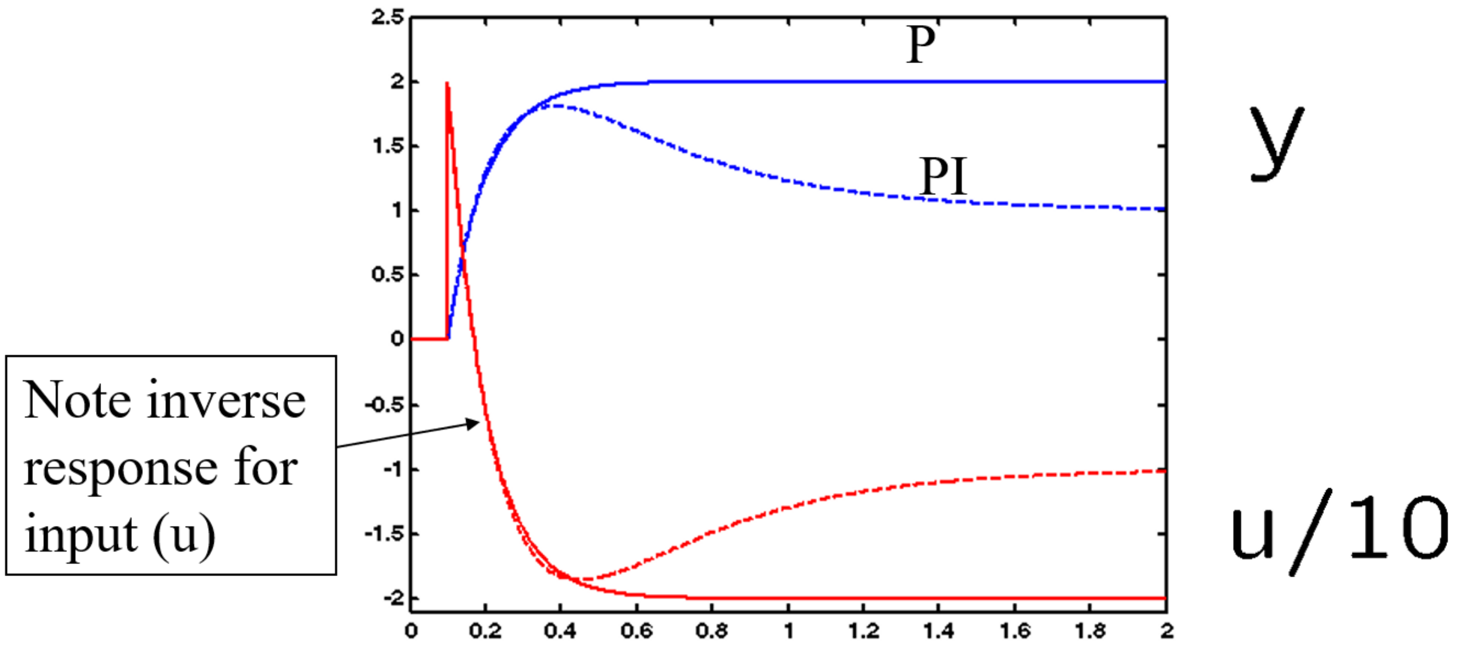


$$\frac{dy}{dt} = In1 + In2 + In3 - Out1 - Out2 = \underbrace{\Sigma_i (u_{0i} + K_{ci}(y_s - y))}_{\text{Everyone contributes to stabilization of y}}$$

NTNU

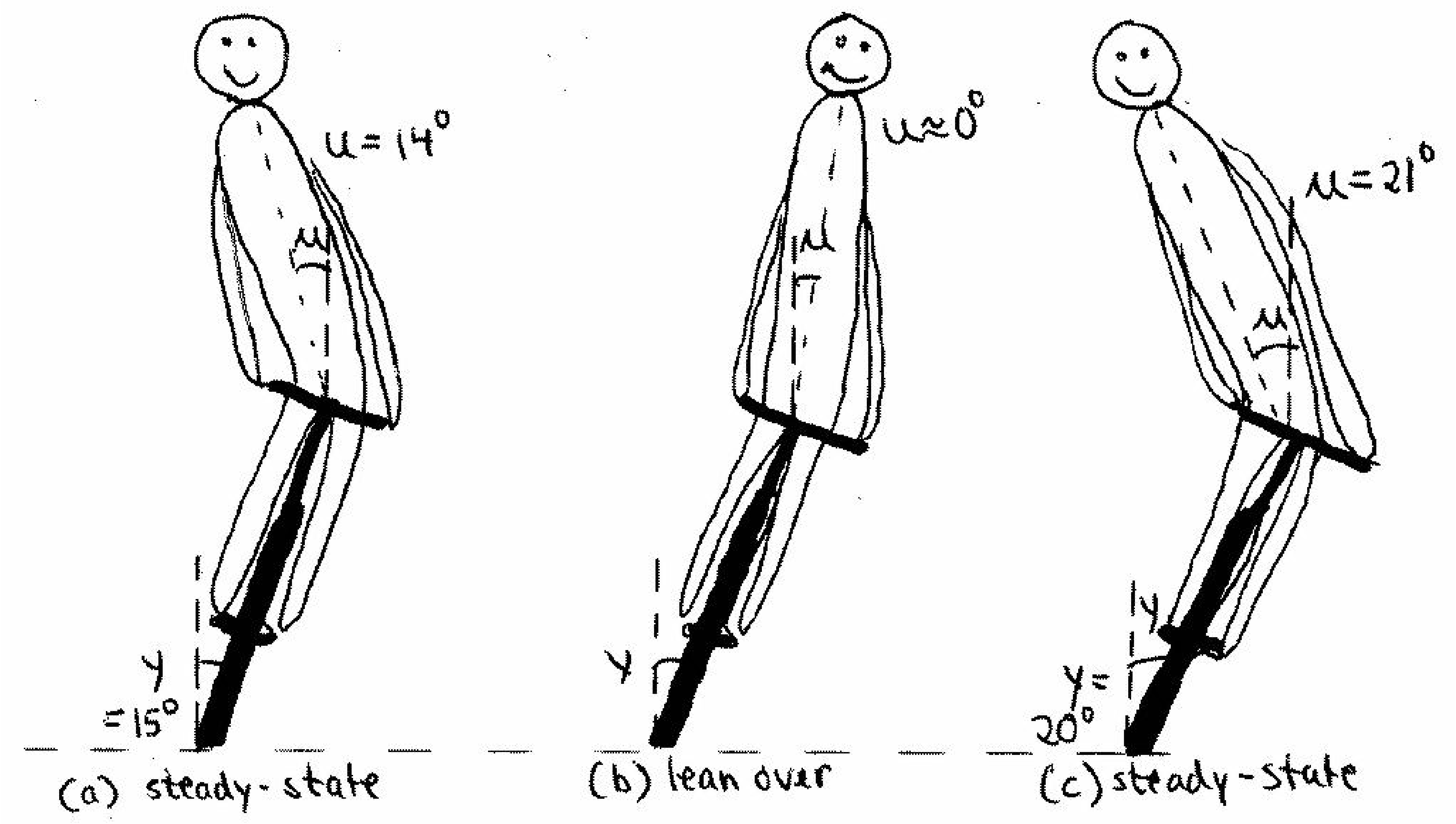
Stabilization using feedback: Instability moves to input (RHP-zero)

$$G = \frac{1}{s - 10}$$



Setpoint change $r = 1$. $G(s) = \frac{1}{s-10}$. $K_c = 20$. For PI: $\tau_I = 1$

Inverse response for bicycle / motorcycle caused by underlying instability

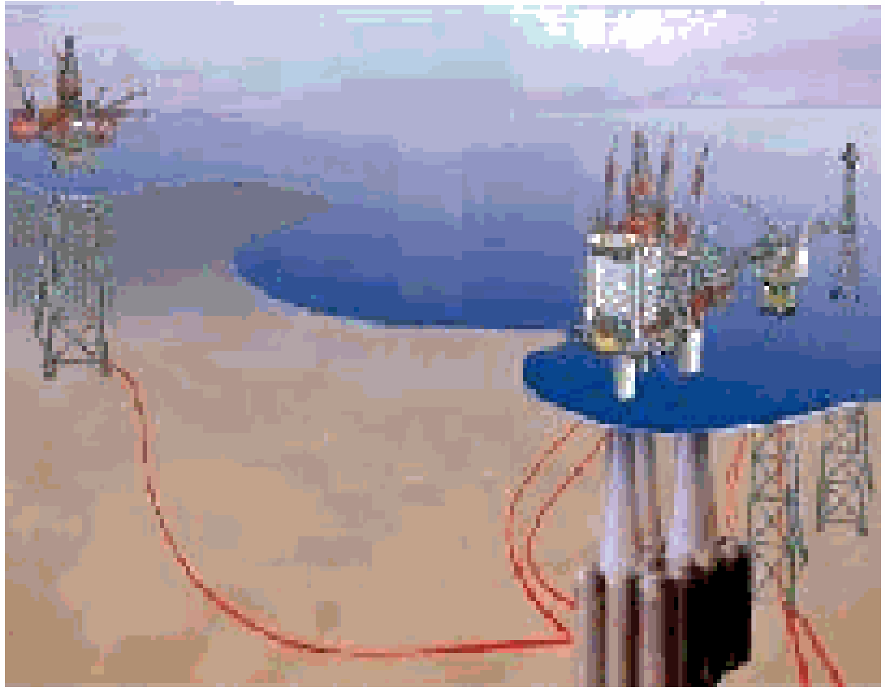


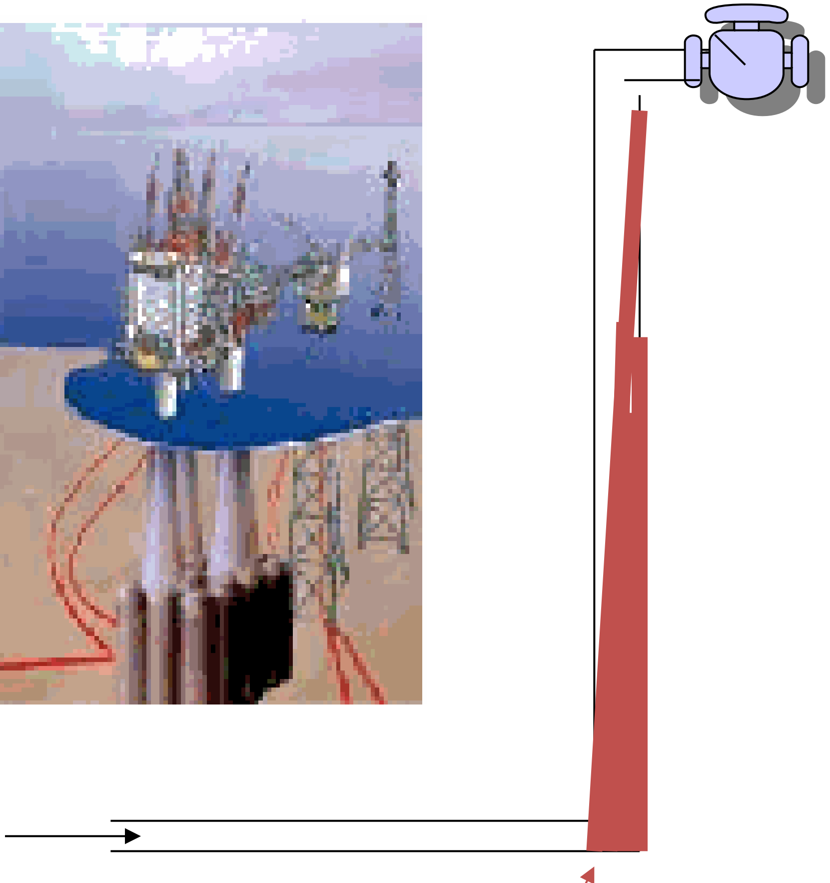
Move first in right direction

But end up moving in opposite direction

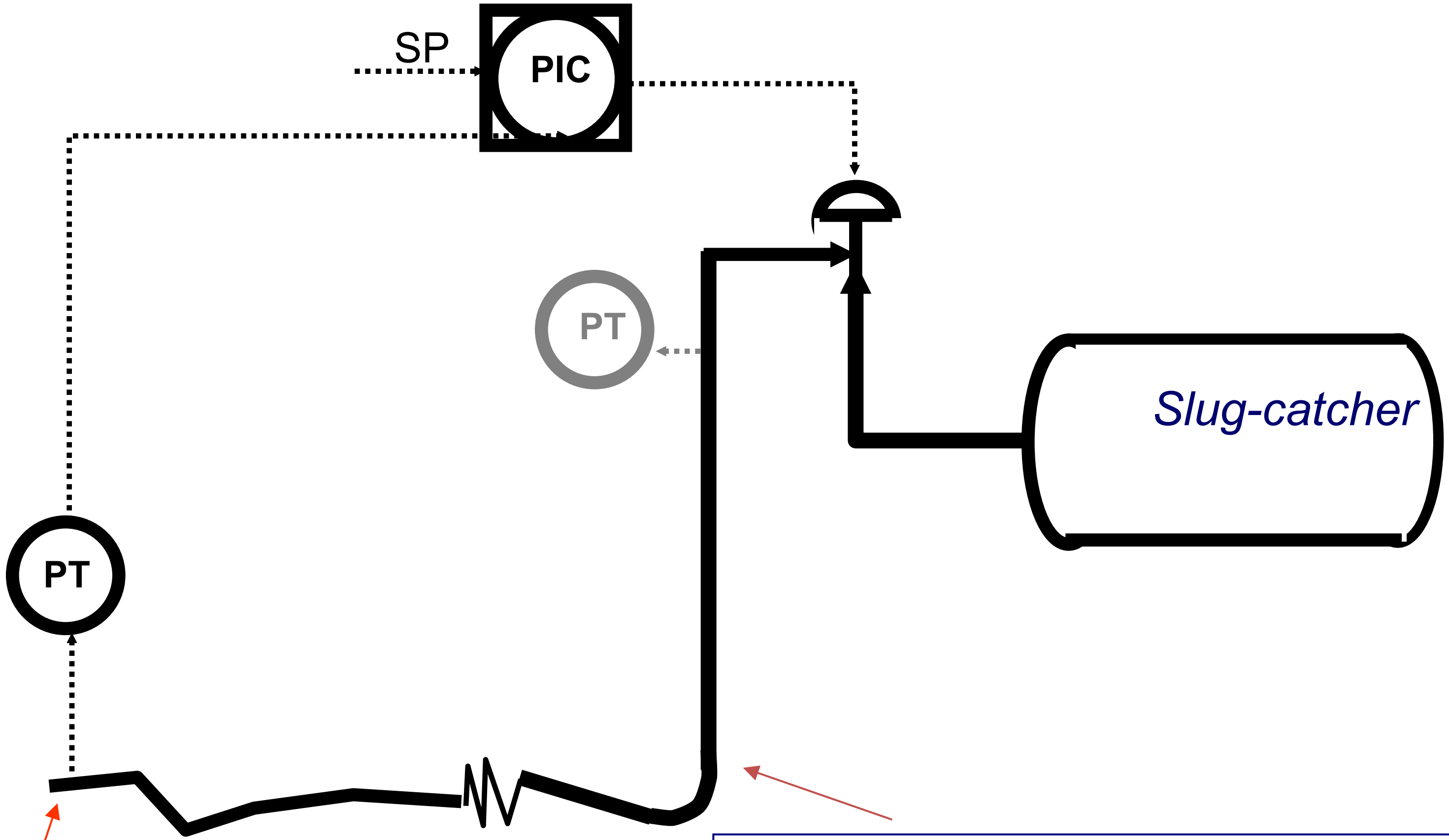
NTNU

Stabilization: Anti-slug control (IFAC, 2002)





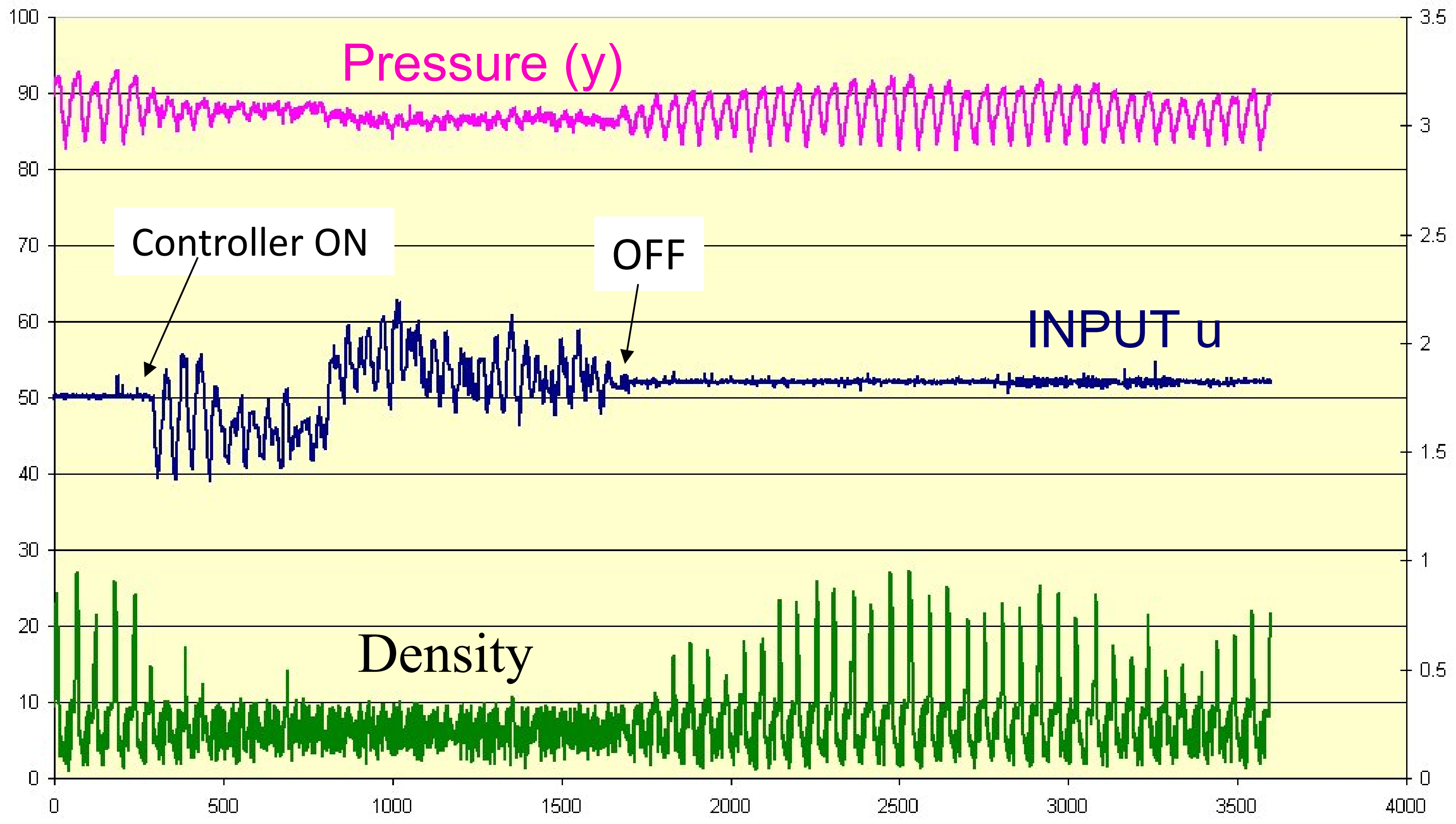
Anti slug-control - control structure



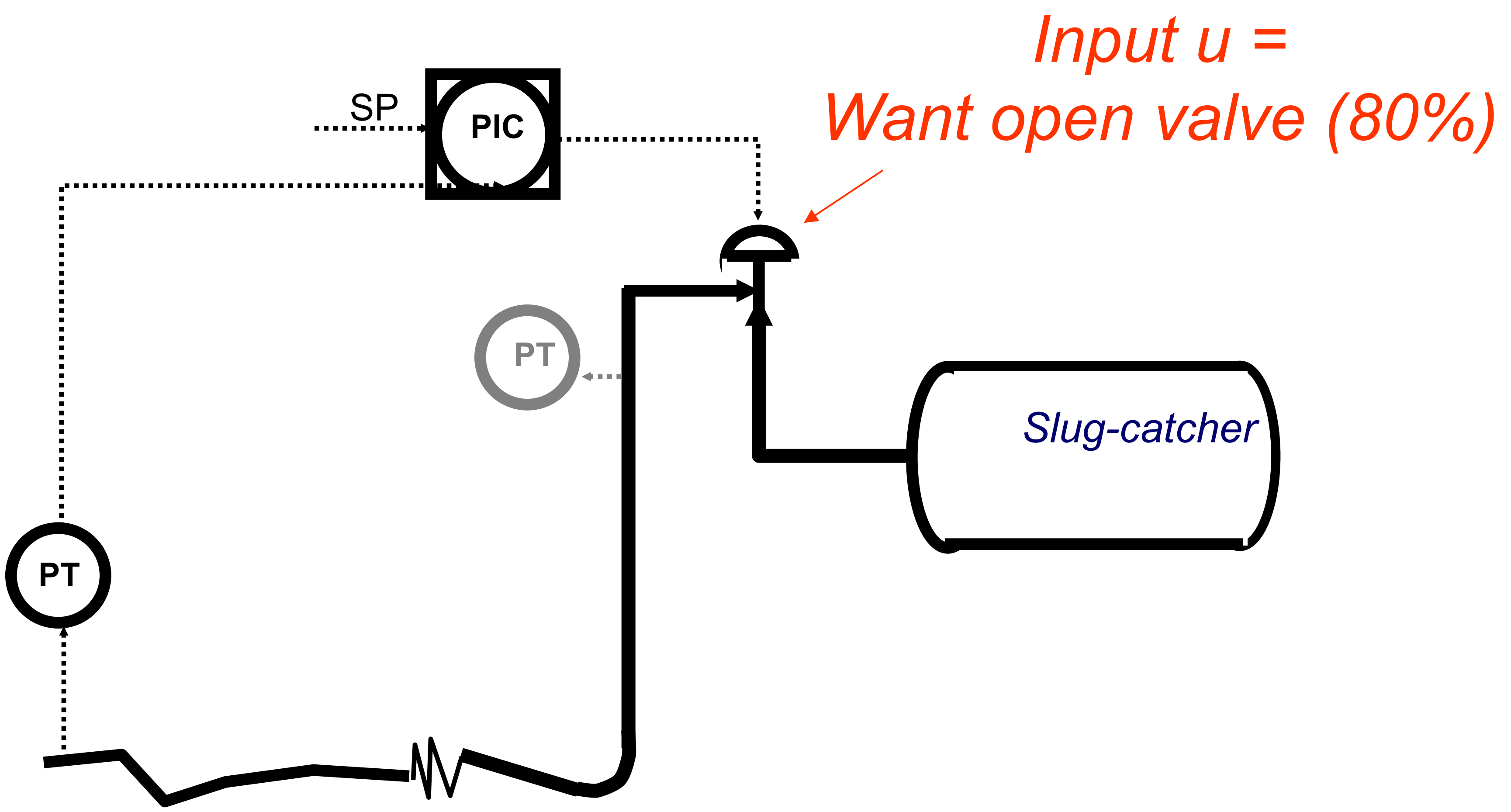
Measurement y

Undesired slug flow (limit cycle) unless feedback control is used to stabilize a steady flow regime (desired, but open-loop unstable)

Anti slug control – experimental data
(Statoil/SINTEF)

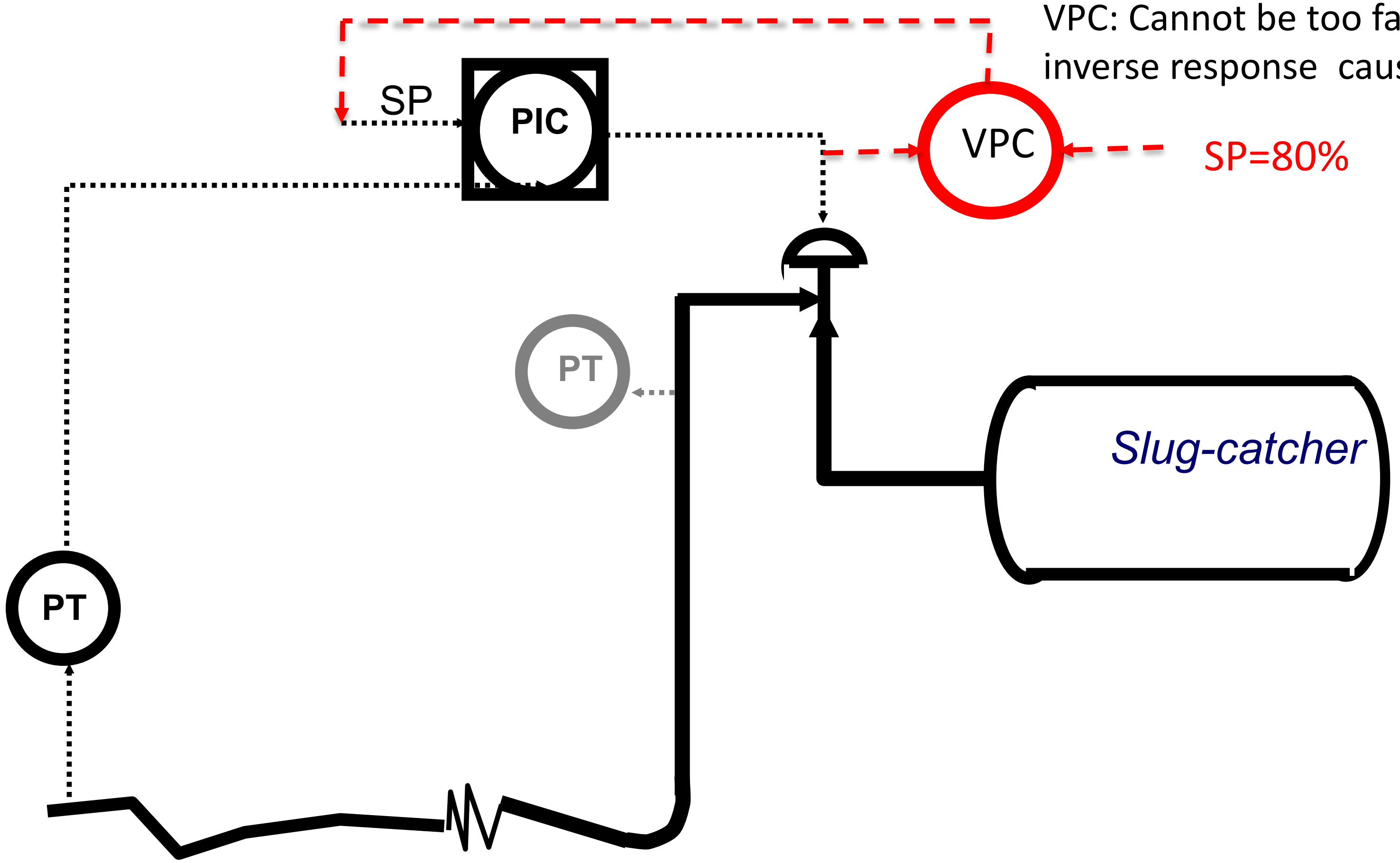


Anti slug-control - control structure



Anti slug-control - control structure

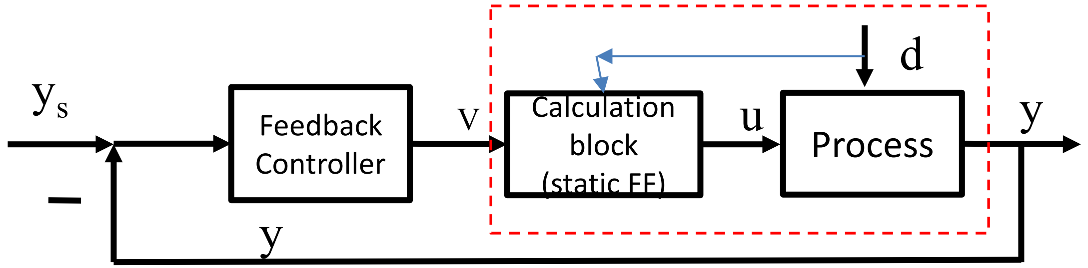
VPC: Cannot be too fast because of inverse response caused by underlying instability



«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

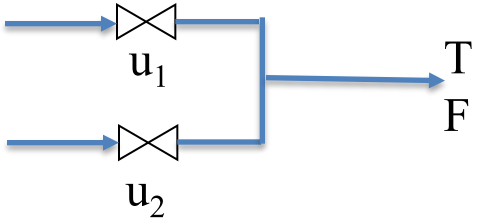
1. Static model: $y = f(u,d)$
2. Select transformed inputs (= controller outputs): $v = f(u,d)$
3. Invert to get physical inputs: $u = f^{-1}(v,d)$
4. Then response from v to y is: $y = I\,v$ (linear, decoupled, perfect disturbance rejection)

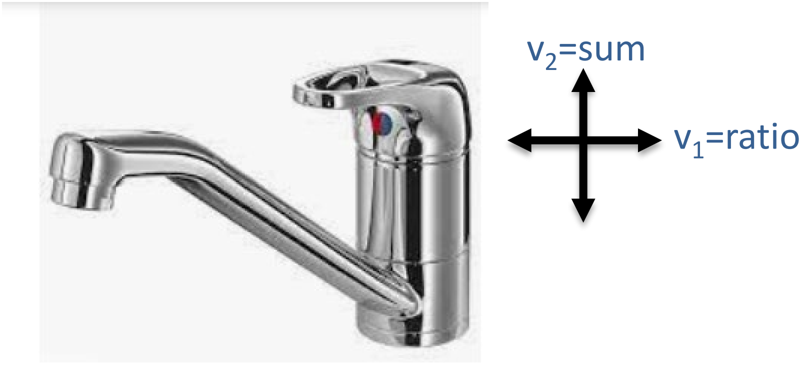


Looks like magic but ut works

 S. Skogestad, C. Zotica, N. Alsop., «Transformed inputs for linearization, decoupling and feedforward control», J. Proc. Control, 2023

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

$$u_1 = v_1 v_2 / (1+ v_1)$$
$$u_2 = v_2 / (1 + v_1)$$

NTNU

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

Example: Mixing of hot and cold water

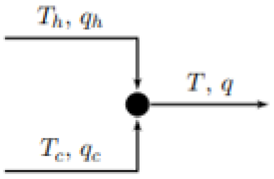
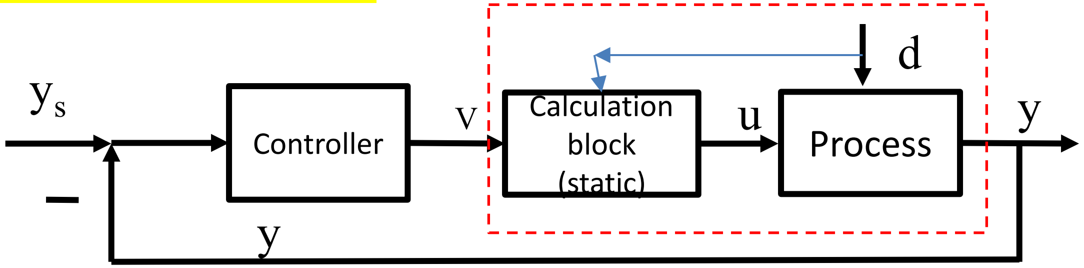


Figure 1: Mixer system



$$u = \begin{pmatrix} q_h \\ q_c \end{pmatrix}$$

$$d = \begin{pmatrix} T_h \\ T_c \end{pmatrix}$$

$$y = \begin{pmatrix} T \\ q \end{pmatrix}$$

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

$$T = \frac{q_h T_h + q_c T_c}{q_h + q_c}$$

$$q = q_c + q_h$$

Select transformed inputs as right hand side, v = f

$$v_1 = \frac{q_h T_h + q_c T_c}{q_h + q_c} \quad (1)$$

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

$$q_c = v_2 - q_h$$

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

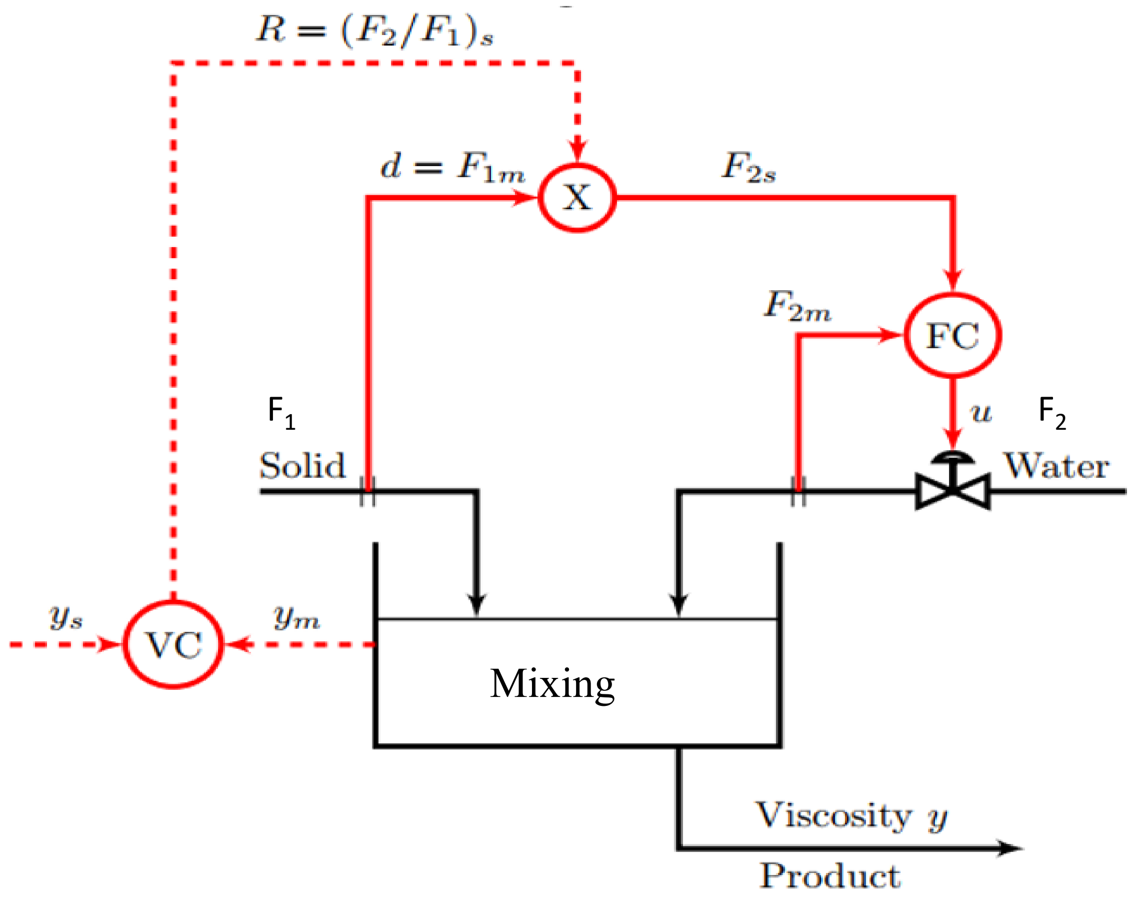
Ratio control

- 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

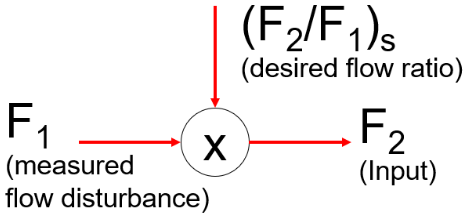


Typical ratio control scheme (solid red lines) with outer feedback loop (dashed red lines) to set the correct ratio.
In the figure, y is viscosity, but generally it can be any intensive variable

- **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)

NTNU

When should we use ratio control?



Answer: When fixing the ratio of all **extensive** variables (e.g., F_2/F_1) gives constant **intensive** variables (y).

The use of ratio control assumes the

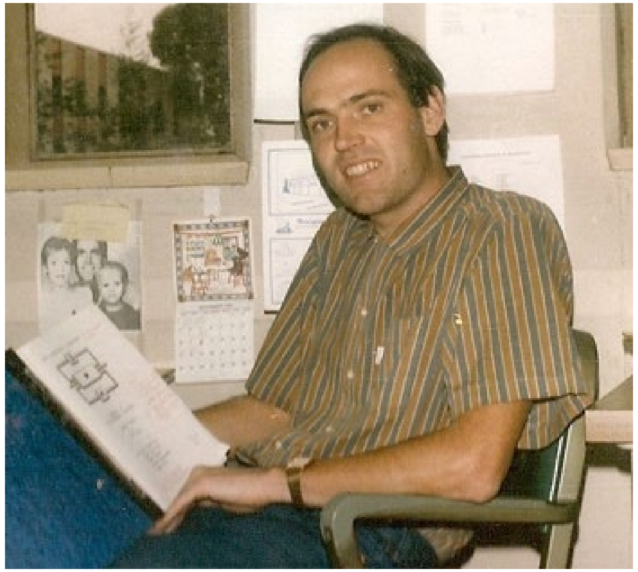
- **Scaling or scale-up property:** We get the same steady-state solution if we increase all extensive variables (flows and heat rates) by the same factor compared to a basis. (Reklaitis, 1981)

For this to hold we must assume constant process efficiencies (Skogestad, 2009):

- **The scaling property** holds if the process efficiencies are independent of the throughput.

- Similar to the use in thermodynamics, the scaling property holds for equilibrium systems. The scaling property (and thus the use of ratio control) applies to many process units, including
  - **Mixing**
  - Equilibrium reactors
  - Equilibrium flash and equilibrium distillation

G.V. Reklaitis, *Introduction to material & energy balances*. Wiley (1983)
S. Skogestad. *Chemical end energy process engineering*, CRC press (2009)



Sigurd at Caltech (1984)

How we design a control system for a complete chemical plant?

- Where do we start?
- What should we control? and why?
- etc.
- etc.

AIChE

Economic Plantwide Control of the Ethyl Benzene Process

Rahul Jagtap, Ashok S Pathak, and Nitin Kaistha
Dept. of Chemical Engineering, Indian Institute of Technology Kanpur, Kanpur 208016,Uttar Pradesh, India

DOI 10.1002/aic.13964
Published online December 10, 2012 in Wiley Online Library (wileyonlinelibrary.com).

A1: Benzene
A2: Ethylene
B: Ethylbenzene (product)
C: Diethylbenzene (undersired, recycled to extinction)
$A1+A2 \rightarrow B$
$B + A2 \rightarrow C$
$C + A1 \rightarrow 2B$

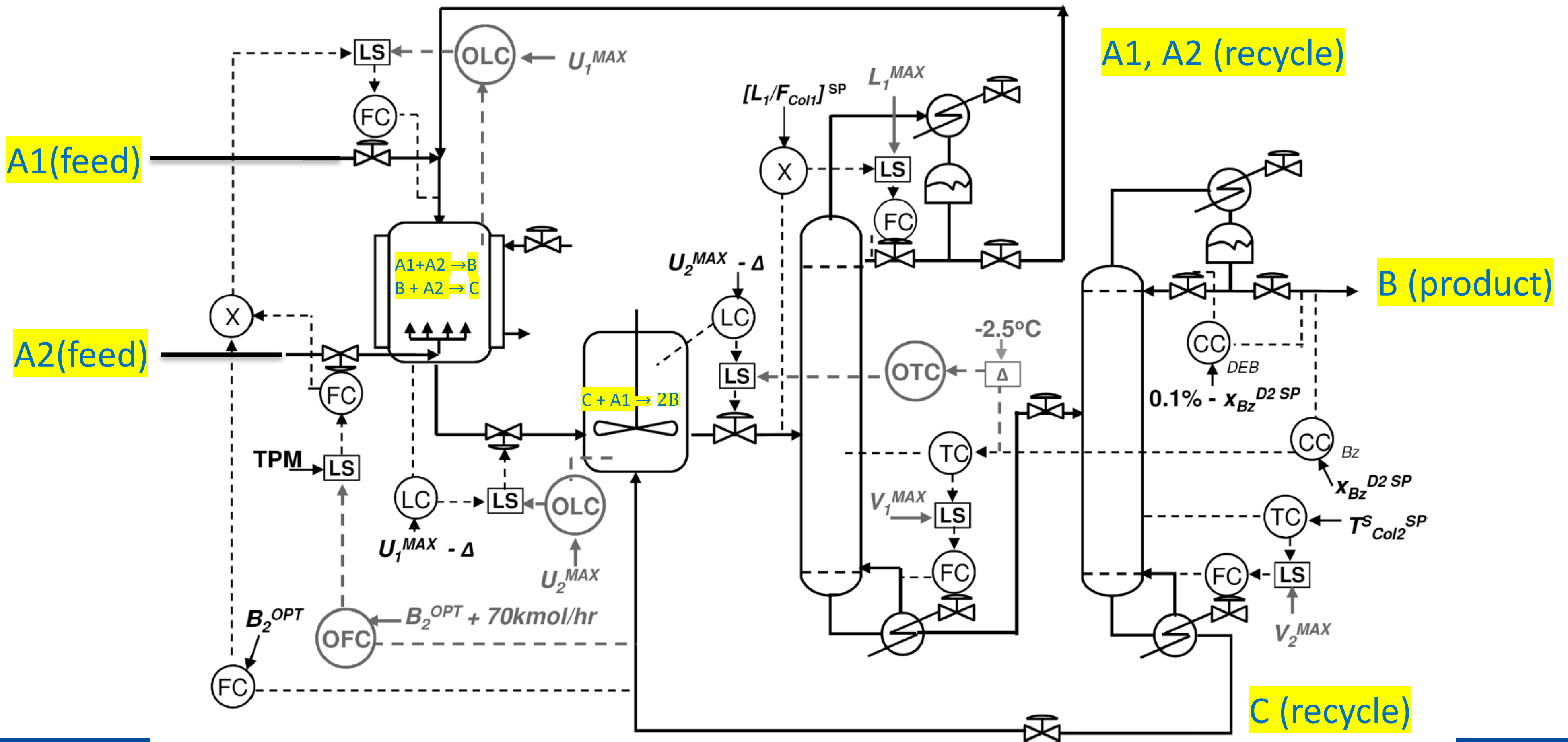


Figure 7. CS2 with overrides for handling equipment capacity constraints.

NTNU

Control system structure*

Alan Foss ("Critique of chemical process control theory", AIChE Journal,1973):

The central issue to be resolved ... is the determination of control system structure.*

Which variables should be measured, which inputs should be manipulated and which links should be made between the two sets?

*Current terminology: Control system architecture

Main objectives of a control system

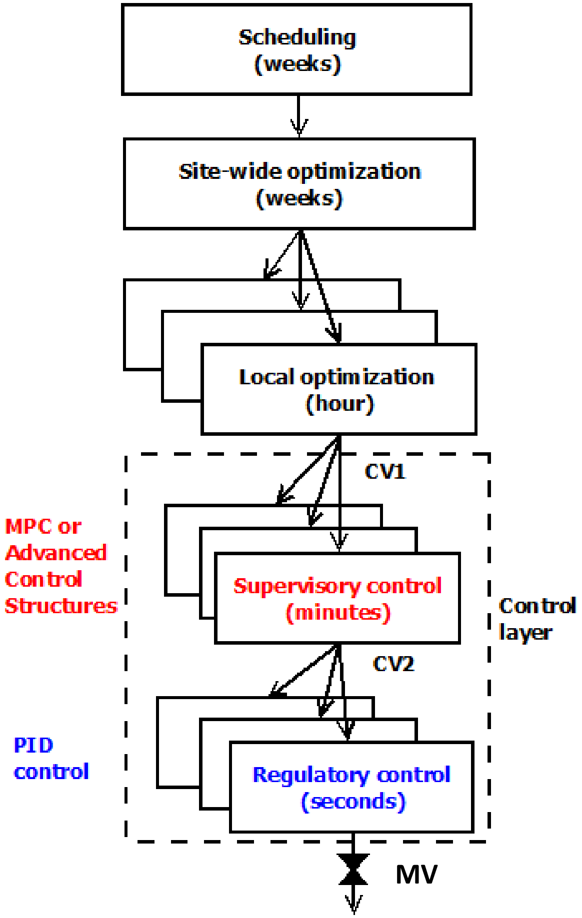
1. Economics: Implementation of acceptable (near-optimal) operation
2. Regulation: Stable operation

ARE THESE OBJECTIVES CONFLICTING?

- Usually NOT
  - Different time scales
    - Stabilization → fast time scale
  - Stabilization doesn’t “use up” any degrees of freedom
    - Reference value (setpoint) available for layer above
    - But it “uses up” part of the time window (frequency range)

Two fundamental ways of decomposing the controller

- Vertical (hierarchical; cascade)
- Based on time scale separation
- Decision: Selection of CVs that connect layers



PROCESS

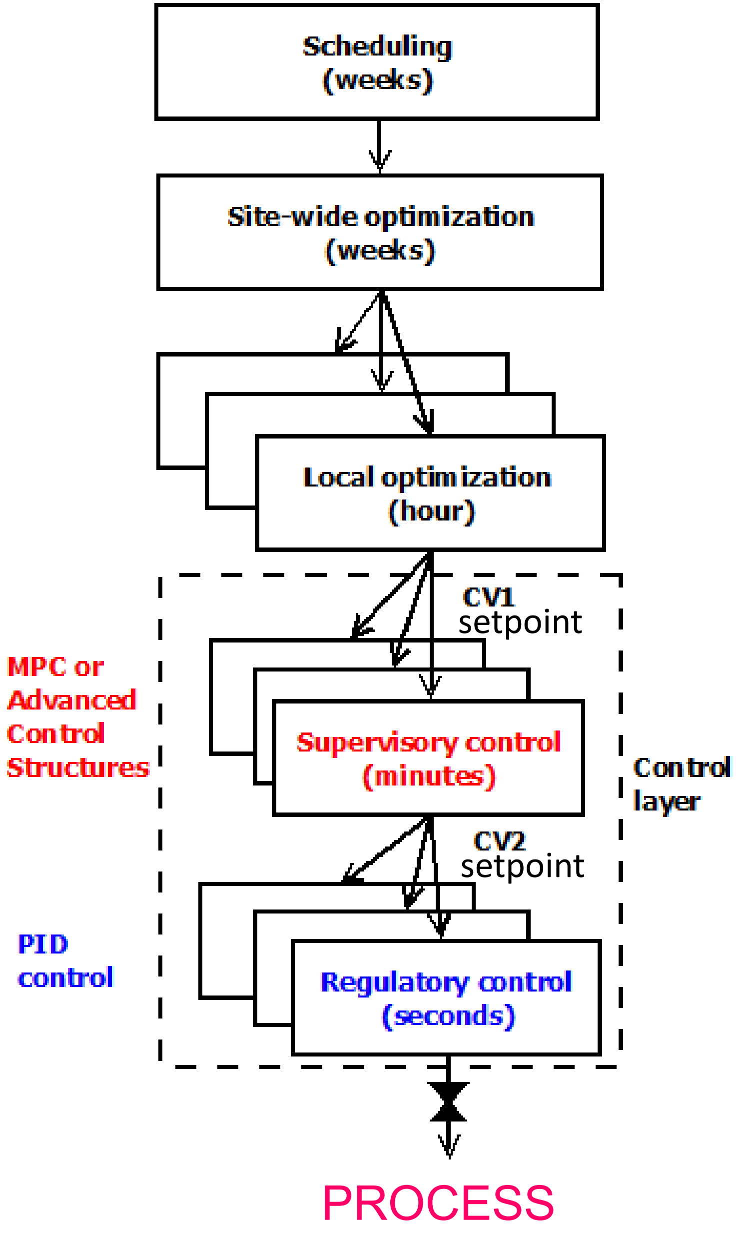
- Horizontal (decentralized)
- Usually based on distance
- Decision: Pairing of MVs and CVs within layers

In addition: Decomposition of controller into smaller elements (blocks):
Feedforward element, nonlinear element, estimators (soft sensors), switching elements

CV = controlled variable MV = manipulated variable

«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



NTNU

Combine control and optimization into one layer?

EMPC: Economic model predictive "control"'

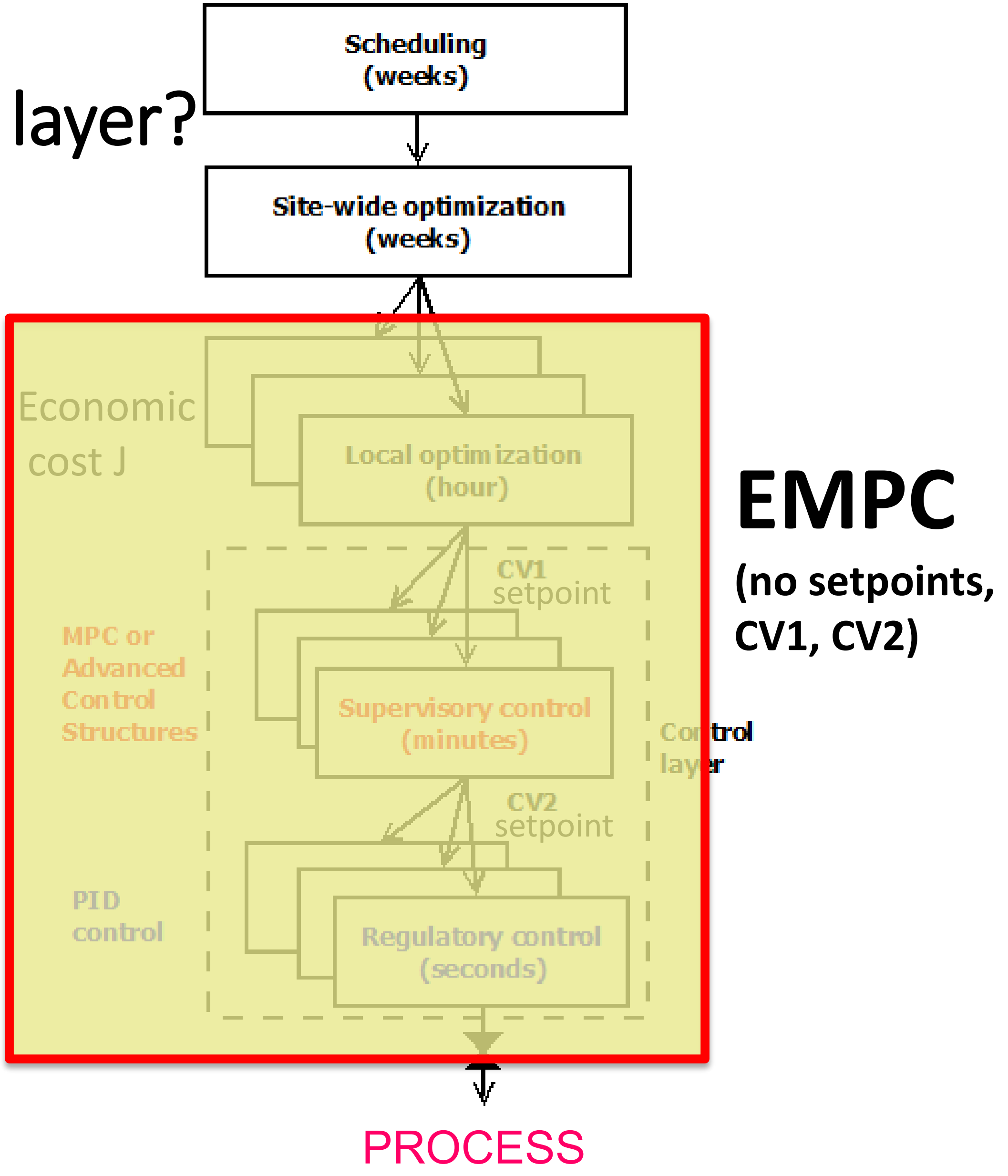
$J_{EMPC} = J + J_{control}$

Penalize input usage, $J_{control} = \Sigma \Delta u_i^2$

NO, combining layers is generally not a good idea!
(the good idea is to separate them!)

One layer (EMPC) is optimal theoreretically, but

- Need detailed dynamic model of everything
- Tuning difficult and indirect
- Slow! (or at least difficult to speed up parts of the control)
- Robustness poor
- Implementation and maintainance costly and time consuming



Typical economic cost function:

J [$/s] = cost feed + cost energy – value products

CV = controlled variable

RTO = real-time optimization

A more fundamental problem with MPC / EMPC

- All claimed stability results are «wrong»
- *They assume that we can measure all states perfectly*
- … or estimate them with a separate estimator block (LQG)

John Doyle (1985):

There are two ways a theorem can be wrong

(from an engineering point of view):

- ***Either it's simply wrong***
- ***Or the assumptions make no sense***

Fact:

Essentially all stability and convergence results for optimal control, MPC and nonlinear control assume full state information, that is, perfect measurement of all states

IEEE TRANSACTIONS ON AUTOMATIC CONTROL, VOL. AC-23, NO. 4, AUGUST 1978

Guaranteed Margins for LQG Regulators

JOHN C. DOYLE

Abstract—There are none.

INTRODUCTION

Considerable attention has been given lately to the issue of robustness of linear–quadratic (LQ) regulators. The recent work by Safonov and Athans [1] has extended to the multivariable case the now well-known guarantee of 60° phase and 6 dB gain margin for such controllers.

Practical «Solution»: Loop transfer recovery (LTR).
Artificial small weight on measurement noise to make estimator fast

“Advanced control”

- Would like: Feedback solutions that can be implemented with minimum need for models

- **“Classical advanced regulatory control“ (ARC) based on single-loop PIDs?**
  – YES!
  – Extensively used by industry
  – Problem for engineers: Lack of design methods
    - Has been around since 1930’s
    - But almost completely neglected by academic researchers
  – Main fundamental limitation: Based on single-loop (need to choose pairing)

ARC = Advanced regulatory control

ARC: Standard Advanced control elements

Each element links a subset of inputs with a subset of outputs. Results in simple local design and tuning

First, there are some elements that are used to improve control for cases where simple feedback control is not sufficient:

E1*. Cascade control[2]
E2*. Ratio control
E3*. Valve (input)[3] position control (VPC) on extra MV to improve dynamic response.

Next, there are some control elements used for cases when we reach constraints:

E4*. Selective (limit, override) control (for output switching)
E5*. Split range control (for input switching)
E6*. Separate controllers (with different setpoints) as an alternative to split range control (E5)
E7*. VPC as an alternative to split range control (E5)

All the above seven elements have feedback control as a main feature and are usually based on PID controllers. Ratio control seems to be an exception, but the desired ratio setpoint is usually set by an outer feedback controller. There are also several features that may be added to the standard PID controller, including

E8*. Anti-windup scheme for the integral mode
E9*. Two-degrees of freedom features (e.g., no derivative action on setpoint, setpoint filter)
E10. Gain scheduling (Controller tunings change as a given function of the scheduling variable, e.g., a disturbance, process input, process output, setpoint or control error)

In addition, the following more general model-based elements are in common use:

E11*. Feedforward control
E12*. Decoupling elements (usually designed using feedforward thinking)
E13. Linearization elements
E14*. Calculation blocks (including nonlinear feedforward and decoupling)
E15. Simple static estimators (also known as inferential elements or soft sensors)

Finally, there are a number of simpler standard elements that may be used independently or as part of other elements, such as

E16. Simple nonlinear static elements (like multiplication, division, square root, dead zone, dead band, limiter (saturation element), on/off)
E17*. Simple linear dynamic elements (like lead–lag filter, time delay, etc.)
E18. Standard logic elements

[2] The control elements with an asterisk * are discussed in more detail in this paper.

ARC = advanced reguklatory control

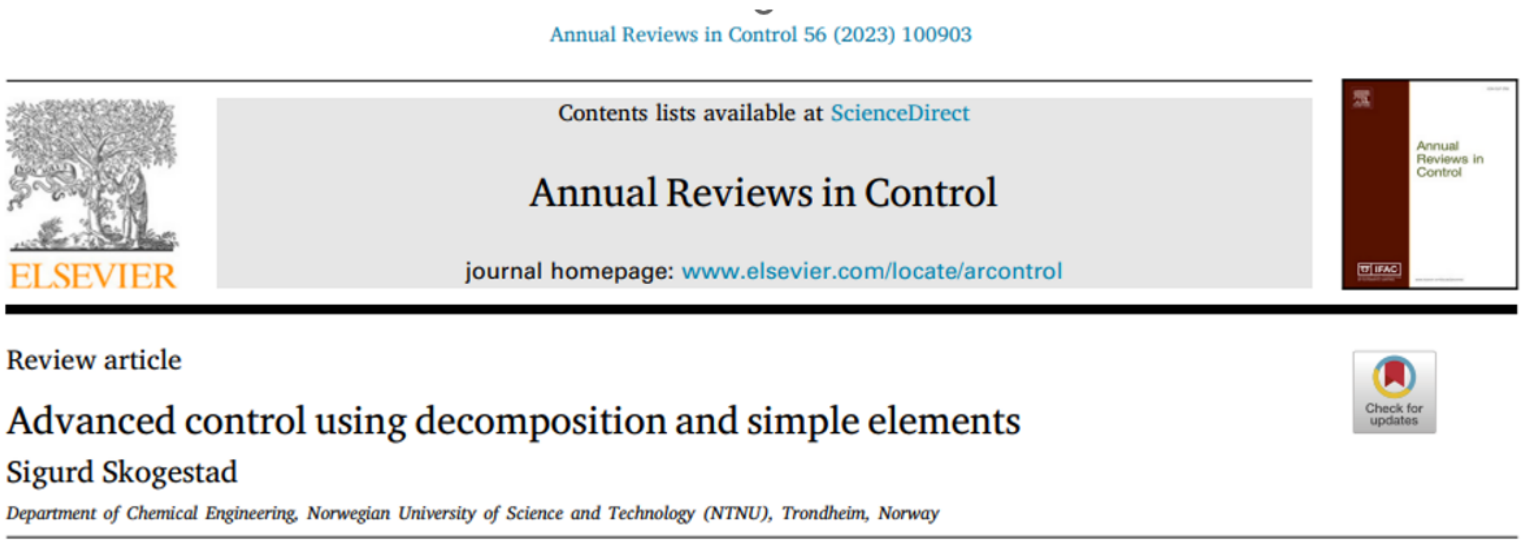
Sigurd Skogestad, "Advanced control using decomposition and simple elements". Annual Reviews in Control, vol. 56 (2023), Article 100903 (44 pages).

How design standard ARC elements?

- Industrial literature (e.g., Shinskey).

  Many nice ideas. But not systematic. Difficult to understand reasoning

- Academia: Very little work
  - I feel alone



Annual Reviews in Control 56 (2023) 100903

Contents lists available at ScienceDirect

Annual Reviews in Control

journal homepage: www.elsevier.com/locate/arcontrol

ELSEVIER

Review article

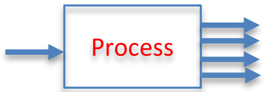
Advanced control using decomposition and simple elements

Sigurd Skogestad

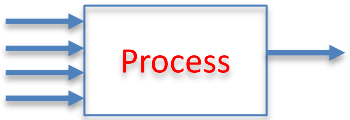
Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), Trondheim, Norway

Constraint switching
(because it is optimal at steady state)

- CV-CV switching
  - Control one CV at a time
  - Use selectors



- MV-MV switching
  - Use one MV at a time
  - Use split range control, multiple controllers or VPC



- MV-CV switching
  - MV saturates so must give up CV
  - Two alterntaives:
    - Simple («do nothing»). If we followed input saturation rule
    - Complex (repairing of loops). Need to combine MV-MV and CV-CV switching

Example adaptive cruise control:
CV-CV switch (min-selector) followed by MV-MV switch (split range control)

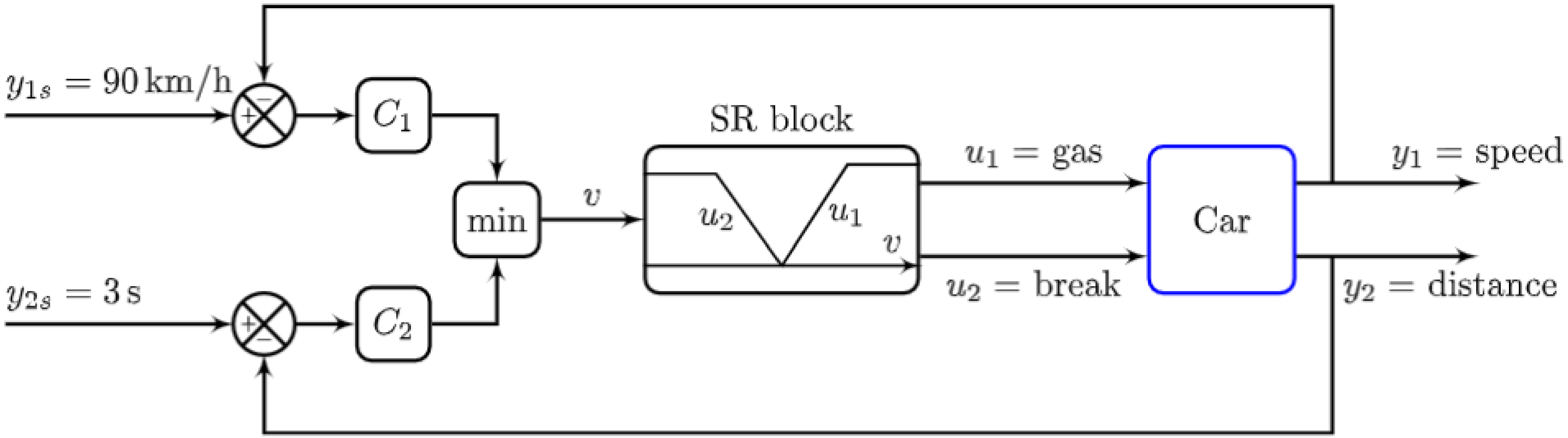
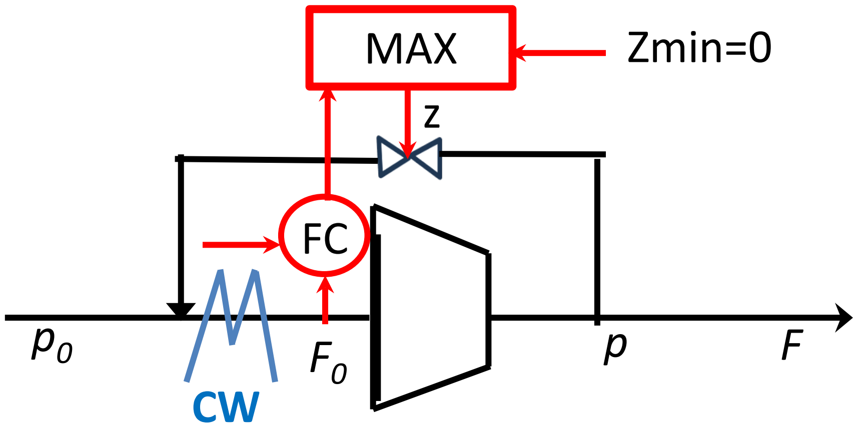


Fig. 31. Adaptive cruise control with selector and split range control.

QUIZ Compressor control

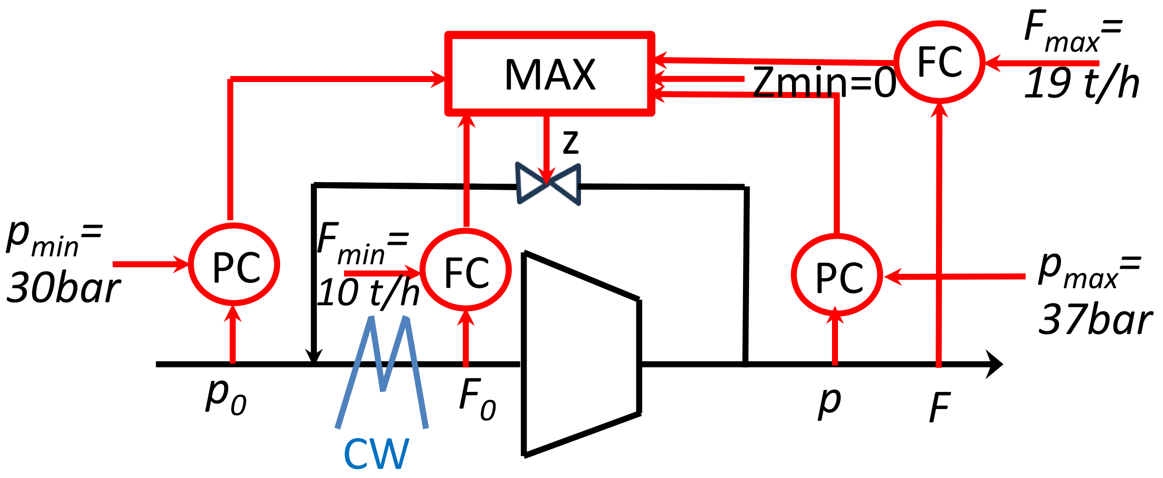


Suggest a solution which achieves

- *$p < p_{max} = 37$ bar (max delivery pressure)*
- *$P_0 > p_{min} = 30$ bar (min. suction pressure)*
- *$F < F_{max} = 19$ t/h (max. production rate)*
- *$F_0 > F_{min} = 10$ t/h (min. through compressor to avoid surge)*

All these 4 constraints are satisfied by a large z
-> MAX-selector

SOLUTION



Conclusion. The future of process control

1. **«Advanced regulatory control» (ARC) or «Advanced PID»:**
   - Works very well in many cases
   - Optimization by feedback (active constraint switching)
   - Need to pair input and output.
     - Advantage: The engineer can specify directly the solution
     - Problem: Unique pairing may not be possible for complex cases
   - Need model only for parts of the process (for tuning)
   - Challenge: Need better teaching and design methods

2. **MPC may be better (and simpler) for some complex multivariable cases**
   - But combine with lower-layer PID (cascade and ratio control)
   - Main challenge MPC: Need dynamic model for whole process (costly)
   - Other challenge: Tuning may be difficult

(3.= **Machine learning**: NO, not enough relevant data)

Academic process control community fish pond

Optimal centralized
Solution (**EMPC**)

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

Annual Reviews in Control 56 (2023) 100903

ELSEVIER

Contents lists available at ScienceDirect

Annual Reviews in Control

journal homepage: www.elsevier.com/locate/arcontrol

Annual Reviews in Control

Review article

Advanced control using decomposition and simple elements

Sigurd Skogestad

Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), Trondheim, Norway

ARTICLE INFO

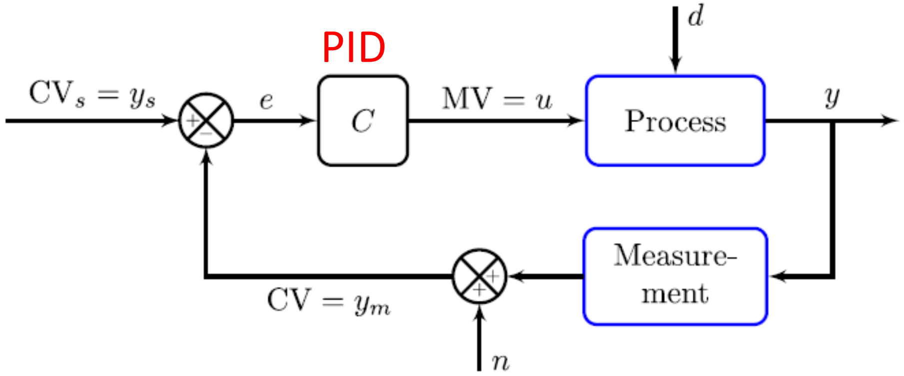
Keywords:
Control structure design
Feedforward control
Cascade control
PID control
Selective control
Override control
Time scale separation
Decentralized control
Distributed control
Horizontal decomposition
Hierarchical decomposition
Layered decomposition
Vertical decomposition
Network architectures

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

Most basic element: Single-loop PID control (E0)



MV-CV Pairing. Two main pairing rules:

1. ***"Pair-close rule" :*** *The MV should have a large, fast, and direct effect on the CV.*
2. ***"Input saturation rule":*** *Pair a MV that may saturate with a CV that can be .given up (when the MV saturates).*
   – *Exception: Have extra MV so we use MV-MV switching (e.g., split range control)*

Additional rule for interactive systems:

3. ***" RGA-rule".*** *Avoid pairing on negative steady-state RGA-element.*

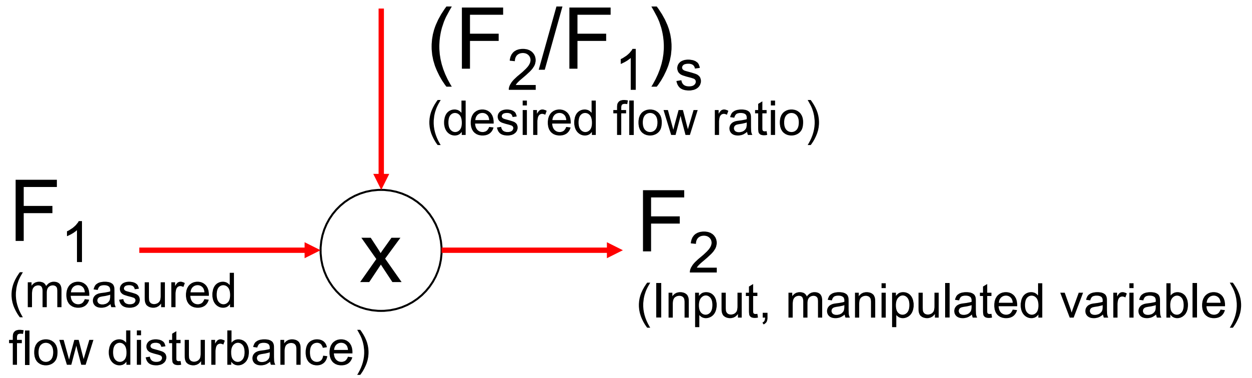
Ratio control

Special case of to feedforward, but don't need model, just process insight.

Always use for mixing streams

- Note: Disturbance needs to be a flow (or more generally an extensive variable)

Use multiplication block (x):



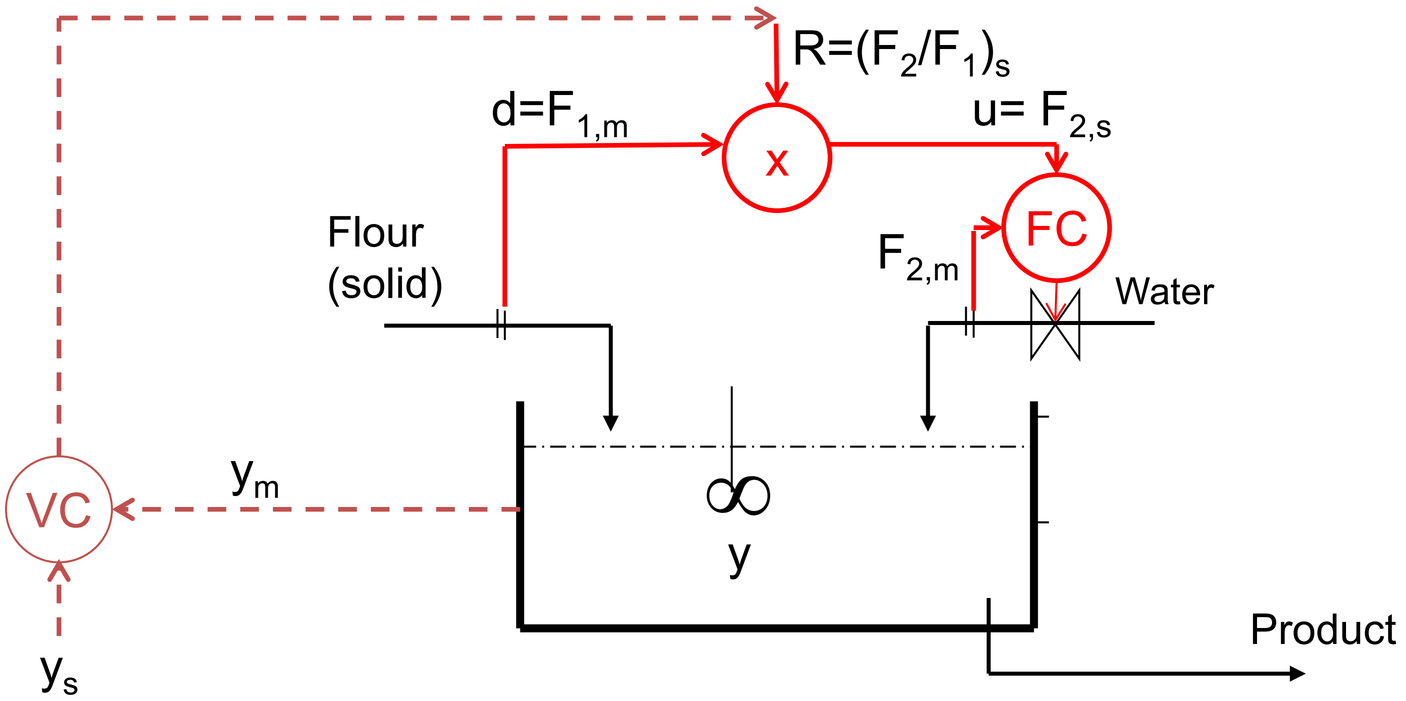
"Measure disturbance ($d=F_1$) and adjust input ($u=F_2$) such that ratio is at given value $(F_2/F_1)_s$"

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)*

EXAMPLE: CAKE BAKING MIXING PROCESS

RATIO CONTROL with outer feedback (to adjust ratio setpoint)

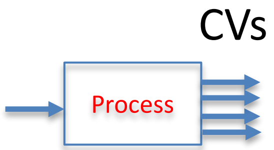


Feedback correction («trim»)

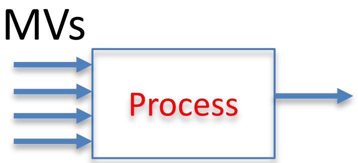
Want to control: Viscosity y [cP]
(or any intensive quality variable, like c, ρ or T)

Constraint switching
(because it is optimal at steady state)

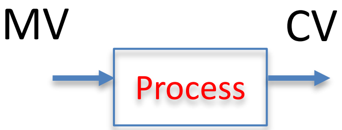
- CV-CV switching
  - Control one CV at a time



- MV-MV switching
  - Use one MV at a time



- MV-CV switching
  - MV saturates so must give up CV
  1. Simple («do nothing»)
  2. Complex (repairing of loops)



MV = manipulated variable
CV = controlled variable

MV-MV switching

- One CV, many MVs (to cover whole steady-state range because primary MV may saturate)*
- Use one MV at a time

Three alternatives:

Alt.1 Split-range control (SRC)

- Plus Generalized SRC (baton strategy)

Alt.2 Several controllers (one for each MV) with different setpoints for the single CV

Alt.3 Valve position control (VPC)

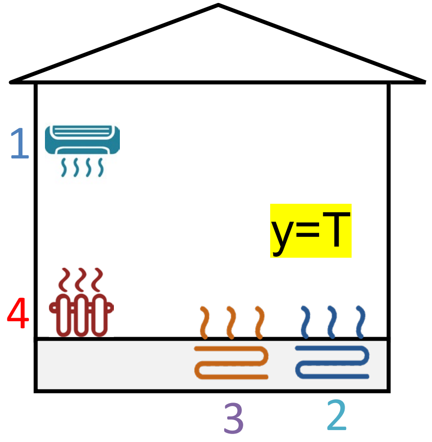
Which is best? It depends on the case!

***Optimal Operation with Changing Active Constraint Regions using Classical Advanced Control, Adriana Reyes-Lua Cristina Zotica, Sigurd Skogestad, Adchem Conference, Shenyang, China. July 2018 ,**

Example MV-MV switching

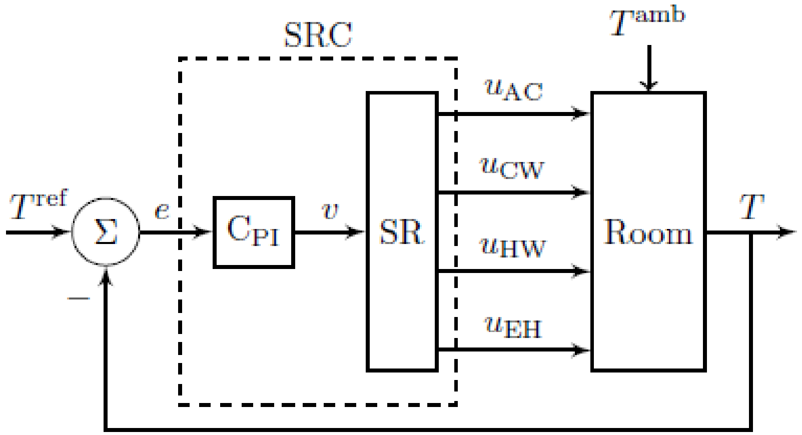
- Break and gas pedal in a car
- Use only one at a time,
- «manual split range control»

Example split range control: Room temperature with 4 MVs

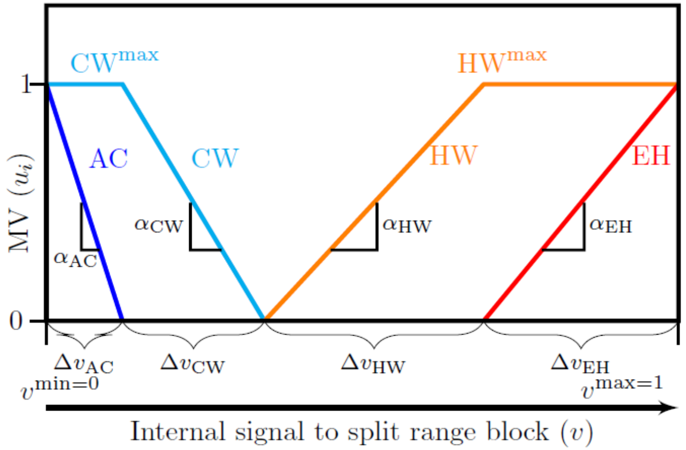


MVs (two for summer and two for winter):

1. AC (expensive cooling)
2. CW (cooling water, cheap)
3. HW (hot water, quite cheap)
4. Electric heat, EH (expensive)



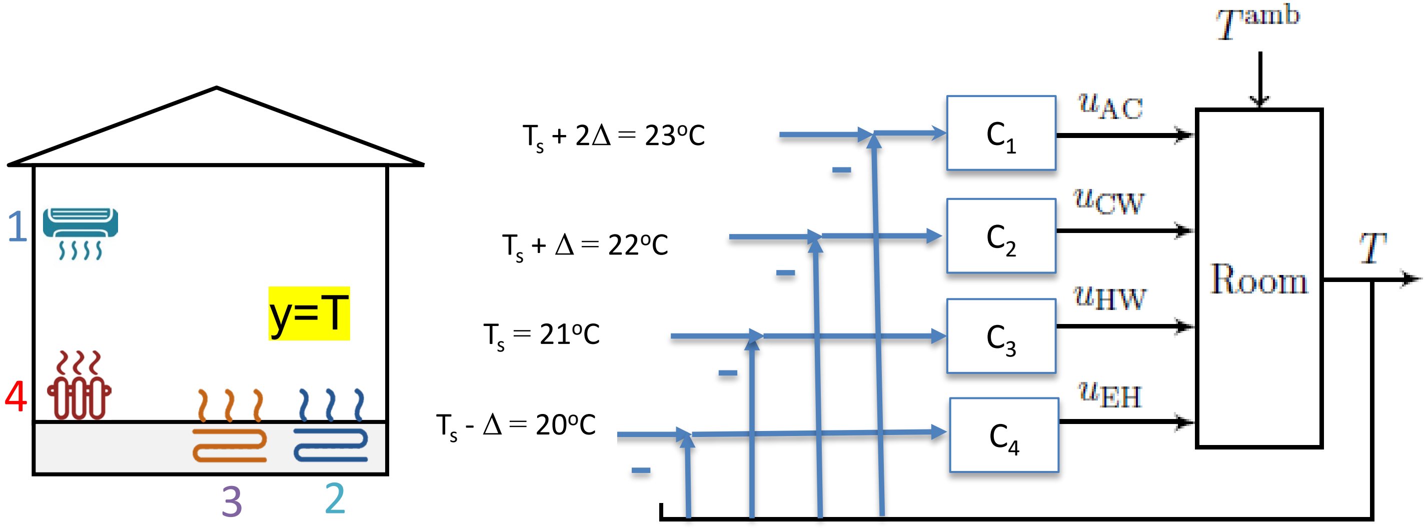
SR-block:



C_{PI} – same controller for all inputs (one integral time)

But get different gains by adjusting slopes α in SR-block

Alternative 2: Multipliple Controllers with different setpoints

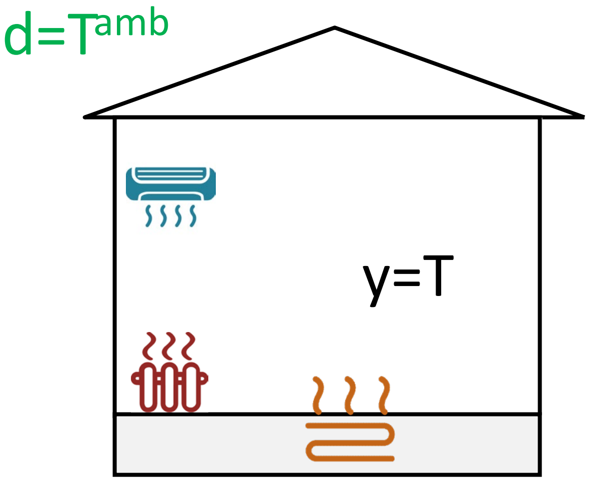


Disadvantage (comfort):
- Different setpoints

Advantage (economics) :
- Different setpoints (energy savings)

Simulation Room temperature

- Dashed lines: SRC (E5)
- Solid lines: Multiple controllers (E6)



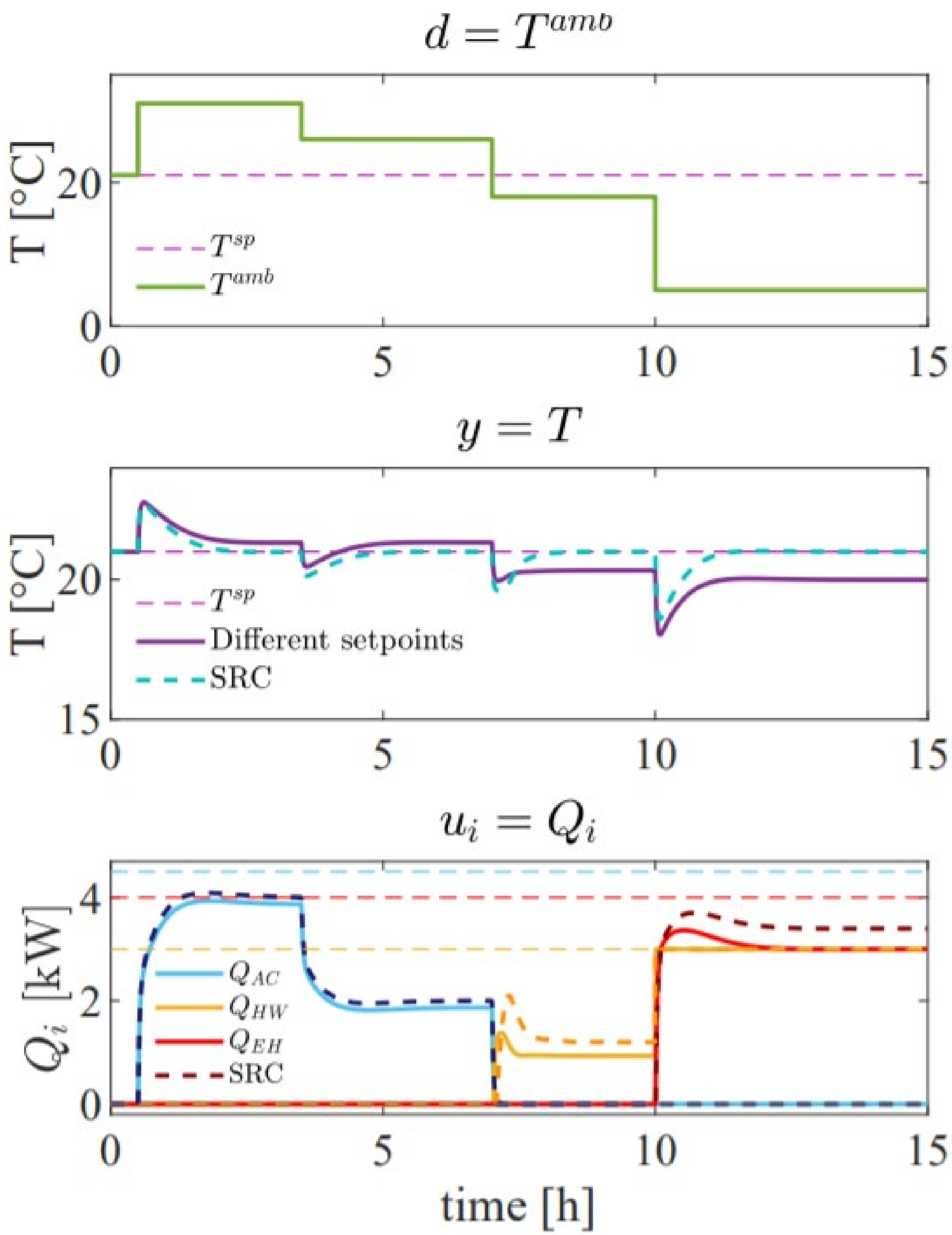


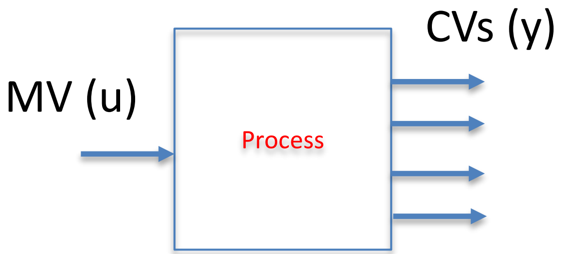
Table 1. Ranges for available inputs (u_k).

Input (u_k)	Description	Nominal	Min	Max	Units
$u_1 = Q_{AC}$	air conditioning	0	0	4.5	kW
$u_2 = Q_{HW}$	heating water	0	0	3.0	kW
$u_3 = Q_{EH}$	electrical heating	0	0	4.0	kW

SRC = split range control

A Reyes-Lua, S Skogestad. Multiple-Input Single-Output Control for Extending the Steady-State Operating Range - Use of Controllers with Different Setpoints. Processes 7 (12), 941 (2019)

CV-CV switching



- Only one input (MV) controls many outputs (CVs)
  - Typically caused by change in active constraint
- Always use MIN- or MAX-selector
  - Example: Control car speed (y_1) - but give up if too small distance (y_2) to car in front.

Example adaptive cruise control: CV-CV switch followed by MV-MV switch

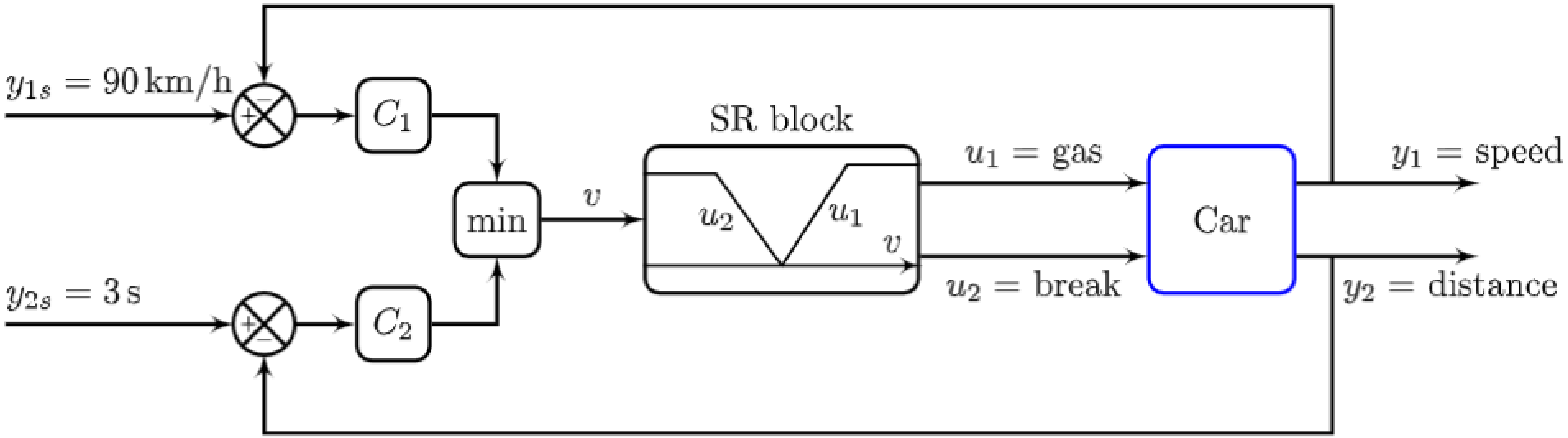
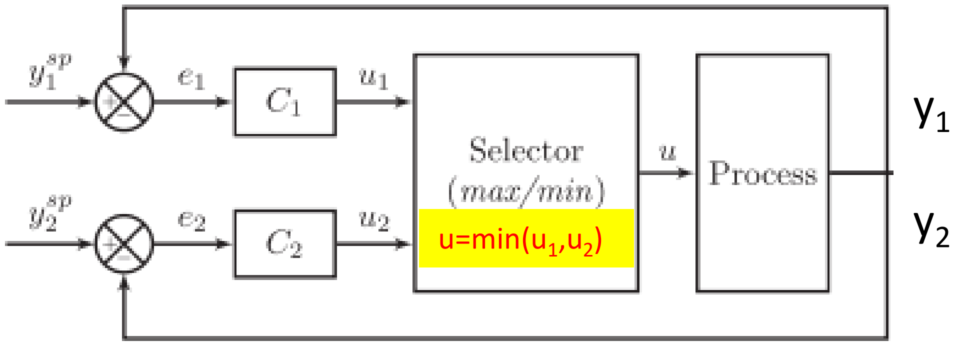


Fig. 31. Adaptive cruise control with selector and split range control.

Note: This is <u>not</u> Complex MV-CV switching, because then the order would be opposite.

Selector: One input (u), several outputs (y_1,y_2)



> = MAX = HS

< = MIN = LS

- Many CVs paired with one MV, but only CV controlled at a time
- This requires output-output (CV-CV) switching: Use selector*
- Note: The selector is usually on the input u, even though the setpoint/constraint is on the output y
- Sometimes called “override”
  - OK name for temporary dynamic fix, but otherwise a bit misleading**
- Selectors work well, but require pairing each constraint with a given input (not always possible)

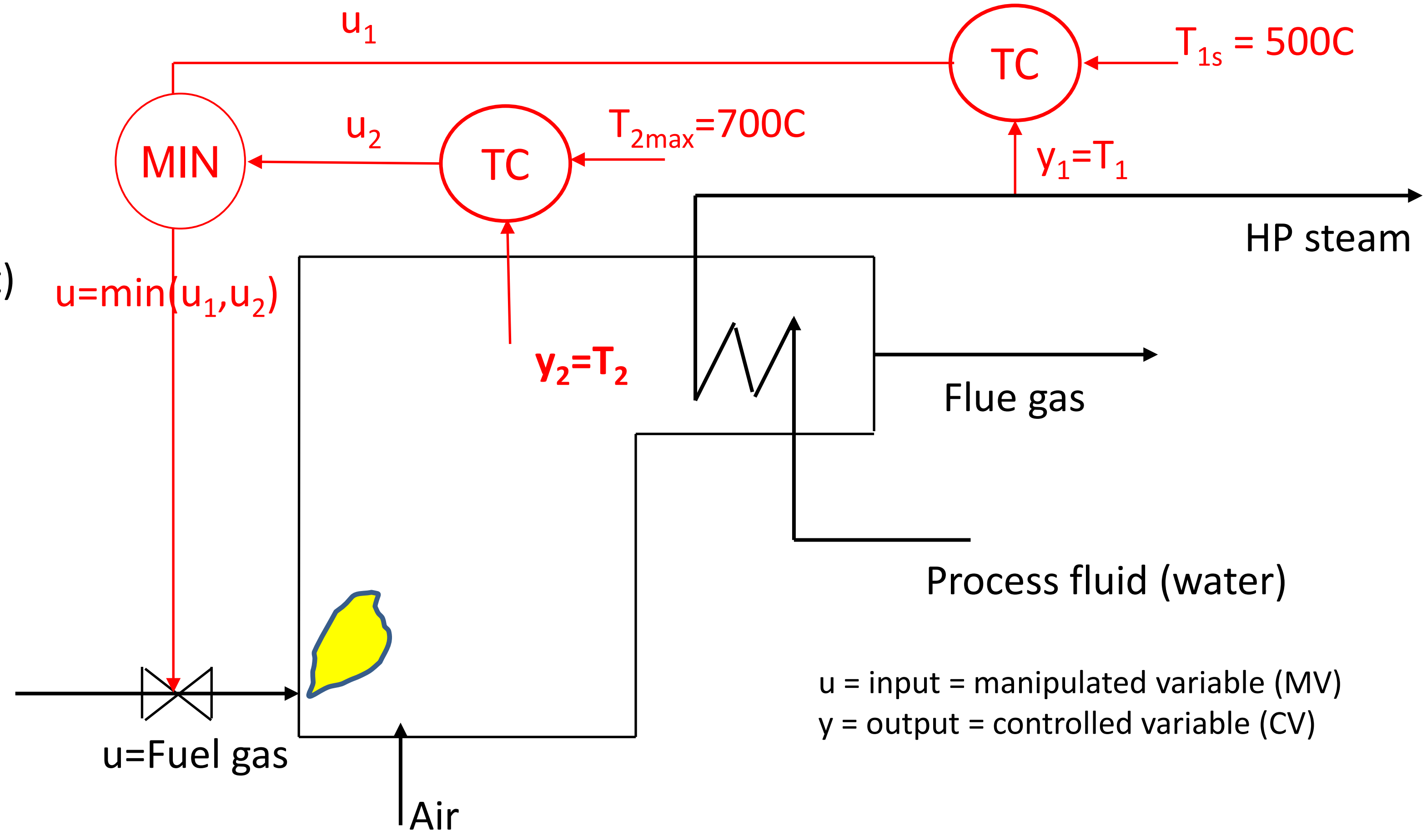
*Only option for CV-CV switching. Well, not quite true: Selectors may be implemented in other ways, for example, using «if-then»-logic.
** I prefer to use the term «override» for undesirable temporary (dynamic) switches, for example, to avoid overflowing a tank dynamically. Otherwise, it’s CV-CV switching

NTNU

Furnace control with safety constraint

Input (MV)
  u = Fuel gas flowrate
Output (CV)
  y_1 = process temperature T_1
    (desired setpoint or max constraint)
  y_2 = furnace temperature T_2
    (T2max= 700C)

Rule: Use min-selector for constraints that are satisfied with a small input



u = input = manipulated variable (MV)
y = output = controlled variable (CV)

Design of selector structure

Rule 1 (max or min selector)

- Use max-selector for constraints that are satisfied with a large input
- Use min-selector for constraints that are satisfied with a small input

Rule 2 (order of max and min selectors):

- If need both max and min selector: Potential infeasibility (conflict)
- Order does not matter if problem is feasible
- If infeasible: Put highest priority constraint at the end

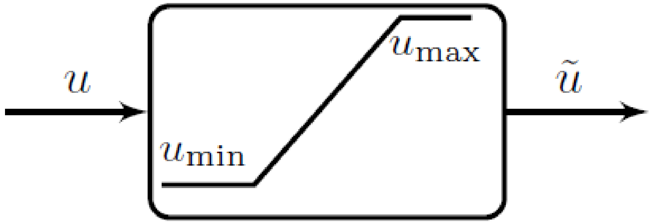
“Systematic design of active constraint switching using selectors.” Dinesh Krishnamoorthy , Sigurd Skogestad. Computers & Chemical Engineering, Volume 143, (2020)
“Advanced control using decomposition and simple elements”. Sigurd Skogestad. Annual Reviews in Control, Volume 56, 100903 (2023)

132

Valves have "built-in" selectors

Rule 3 (a bit opposite of what you may guess)

- A closed valve (u_{min}=0) gives a "built-in" max-selector (to avoid negative flow)
- An open valve (u_{max}=1) gives a "built-in" min-selector
  - So: Not necessary to add these as selector blocks (but it will not be wrong).
  - The "built-in" selectors are never conflicting because cannot have closed and open at the same time
  - Another way to see this is to note that a valve works as a saturation element



Saturation element may be implemented in three other ways (equivalent because never conflict)

1. Min-selector followed by max-selector
2. Max-selector followed by min-selector
3. Mid-selector

$$\tilde{u} = \max(u_{min}, \min(u_{max}, u)) = \min(u_{max}, \max(u_{min}, u)) = \mathrm{mid}(u_{min}, u, u_{max})$$

MV-CV switching (because reach constraint on MV)

- **Simple CV-MV switching**
  - Don't need to do anything if we followed the *Input saturation rule:*
  - *"Pair a MV that may saturate with a CV that can be given up (when the MV saturates)"*

Anti-surge control (= min-constraint on F)

Minimize recycle (MV=z) subject to

CV= $F \geq F_{min}$

MV= z ≥ 0

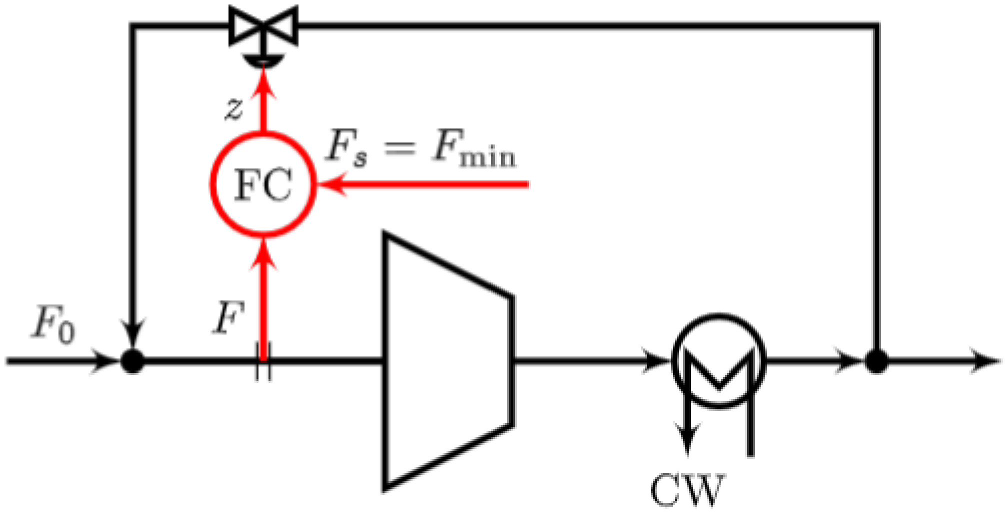
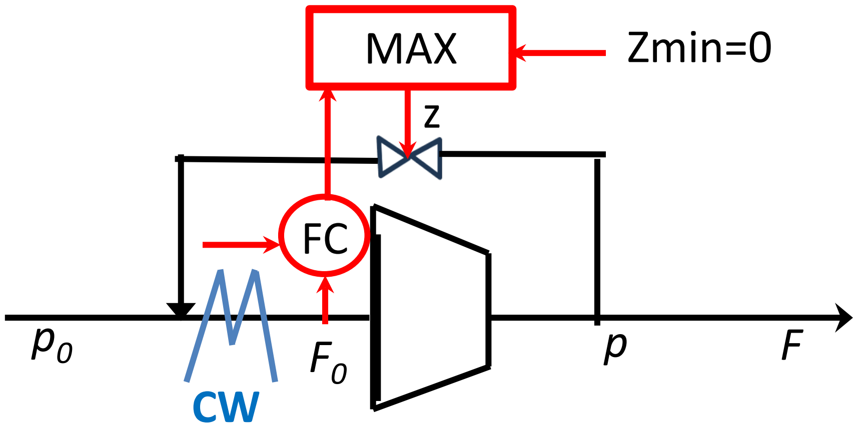


Fig. 32. Flowsheet of anti-surge control of compressor or pump (CW = cooling water). This is an example of simple MV-CV switching: When MV=z (valve position) reaches its minimum constraint ($z = 0$) we can stop controlling CV=F at $F_s = F_{min}$, that is, we do not need to do anything except for adding anti-windup to the controller. Note that the valve has a "built in" max selector.

- No selector required, because MV=z has a «built-in» max-selector at z=0.
- Generally: «Simple» MV-CV switching (with no selector) can be used if we satisfy the input saturation rule: «*Pair a MV that may saturate with a CV that can be given up (when the MV saturates at z=0)"*

QUIZ Compressor control

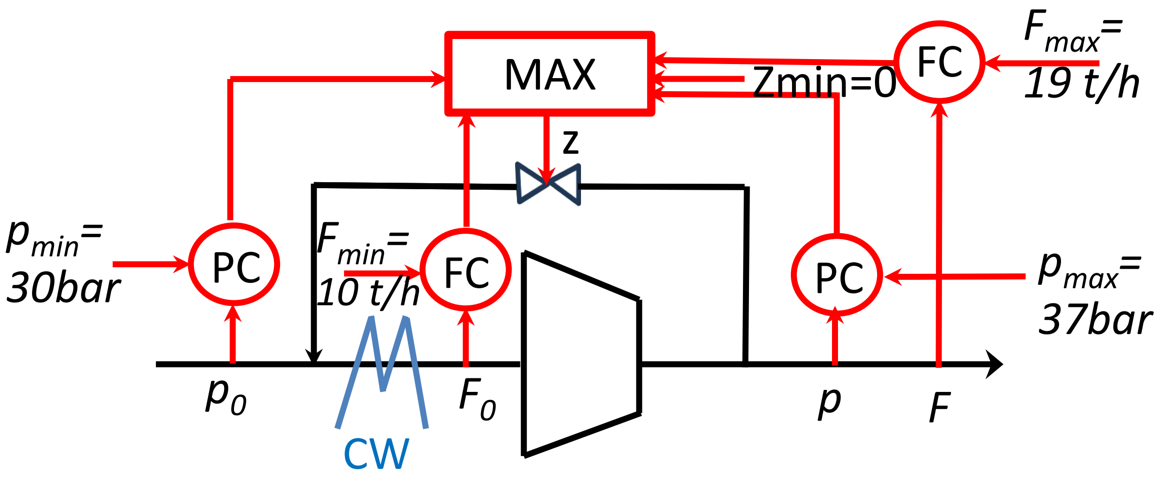


Suggest a solution which achieves

- $p < p_{max} = 37$ *bar* *(max delivery pressure)*
- $P_0 > p_{min} = 30$ *bar* *(min. suction pressure)*
- $F < F_{max} = 19$ *t/h* *(max. production rate)*
- $F_0 > F_{min} = 10$ *t/h* *(min. through compressor to avoid surge)*

All these 4 constraints are satisfied by a large z
-> MAX-selector

SOLUTION



NTNU

Inventory control

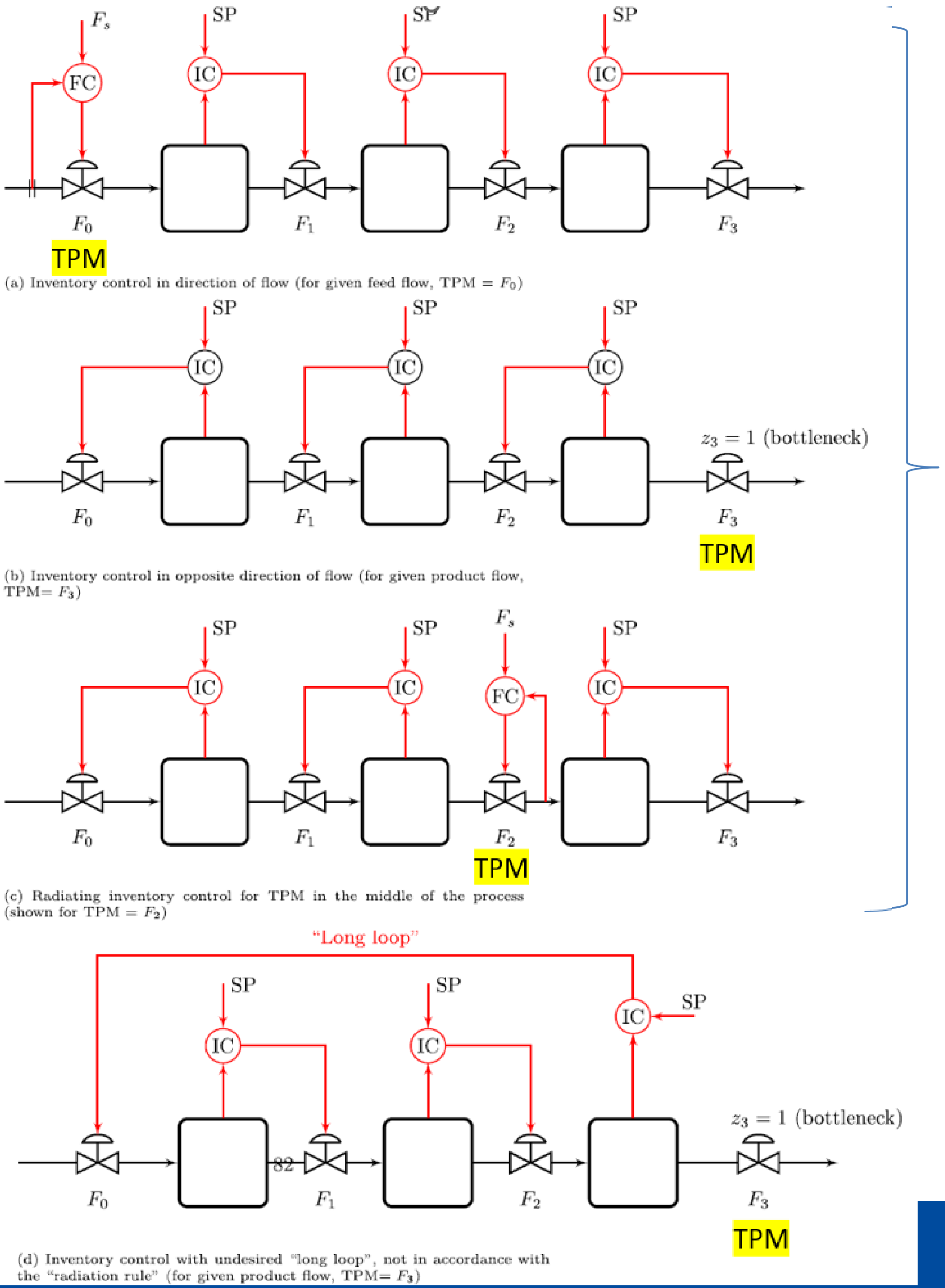
- Very important decison for plantwide control:
  - Location of TPM

- TPM = Throughput manipulator

  = *Gas Pedal = Variable used for setting the throughput/production rate (for the entire process).*

Inventory control for units in series

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



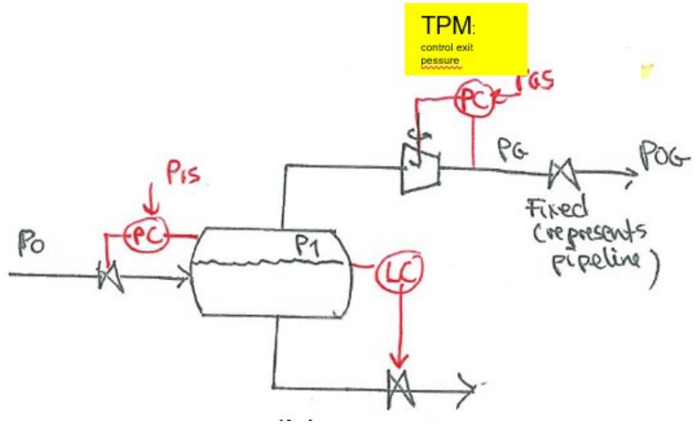
Follows radiation rule

Does NOT follow radiation rule

Rules for inventory control

Rules for inventory control

- **Rule 1.** Cannot control (set the flowrate) the same flow twice
- **Rule 2**. Follow the radiation rule whenever possible
- **Rule 3. (**which should never been broken): No inventory loop should cross the location of the TPM
- **Rule 4**. Controlling inlet or outlet pressure indirectly sets the flow (indirectly makes it a TPM)

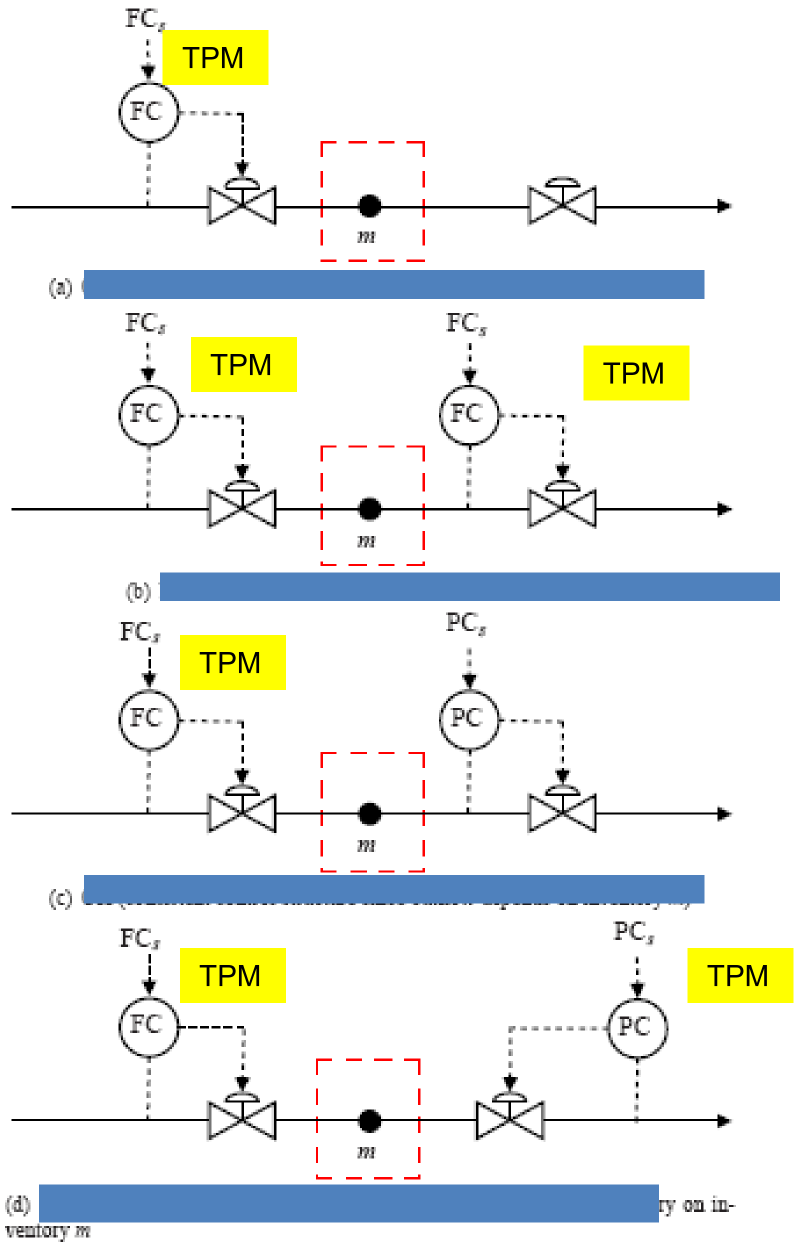


Rule 2. Controlling outlet pressure sets flow

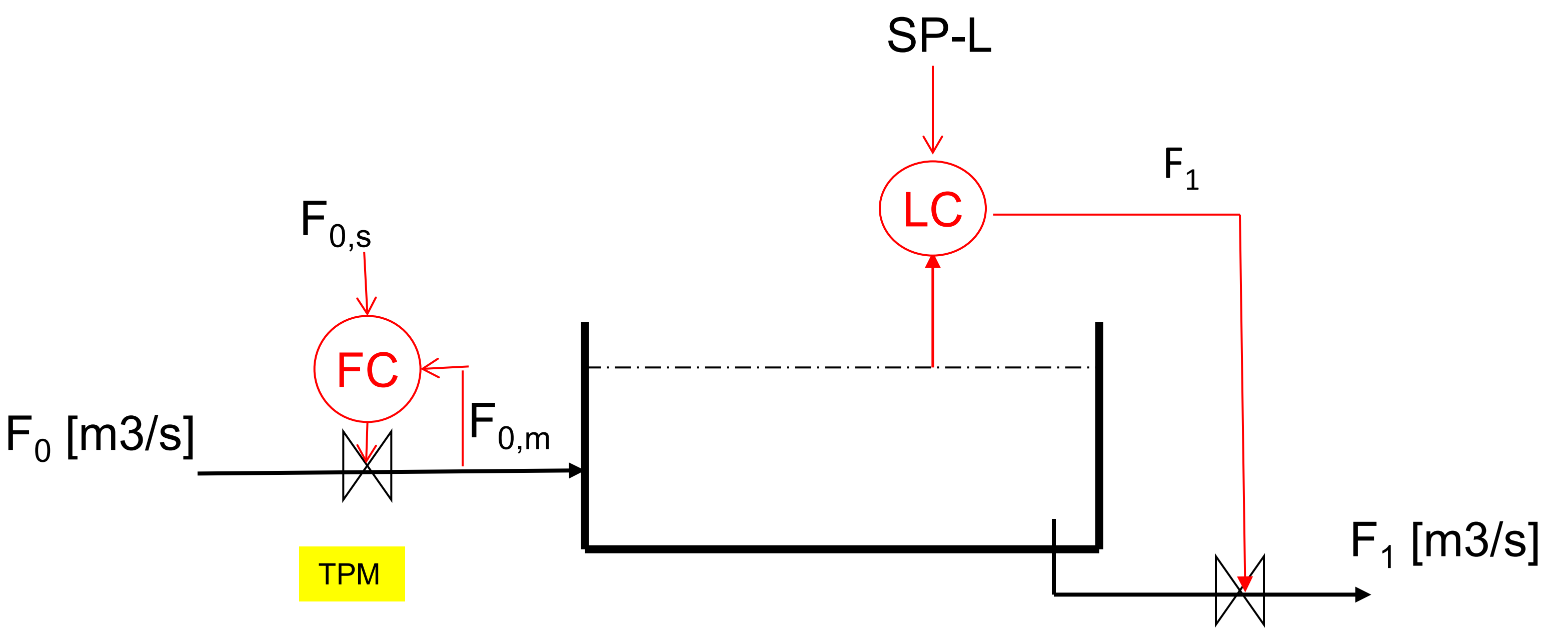
TPM = Variable used for setting the throughput/production rate (for the entire process).

QUIZ. Are these structures workable (consistent)? Yes or No?

Hint: What happens to the mass holdup inside the red box? Is it self-regulated?

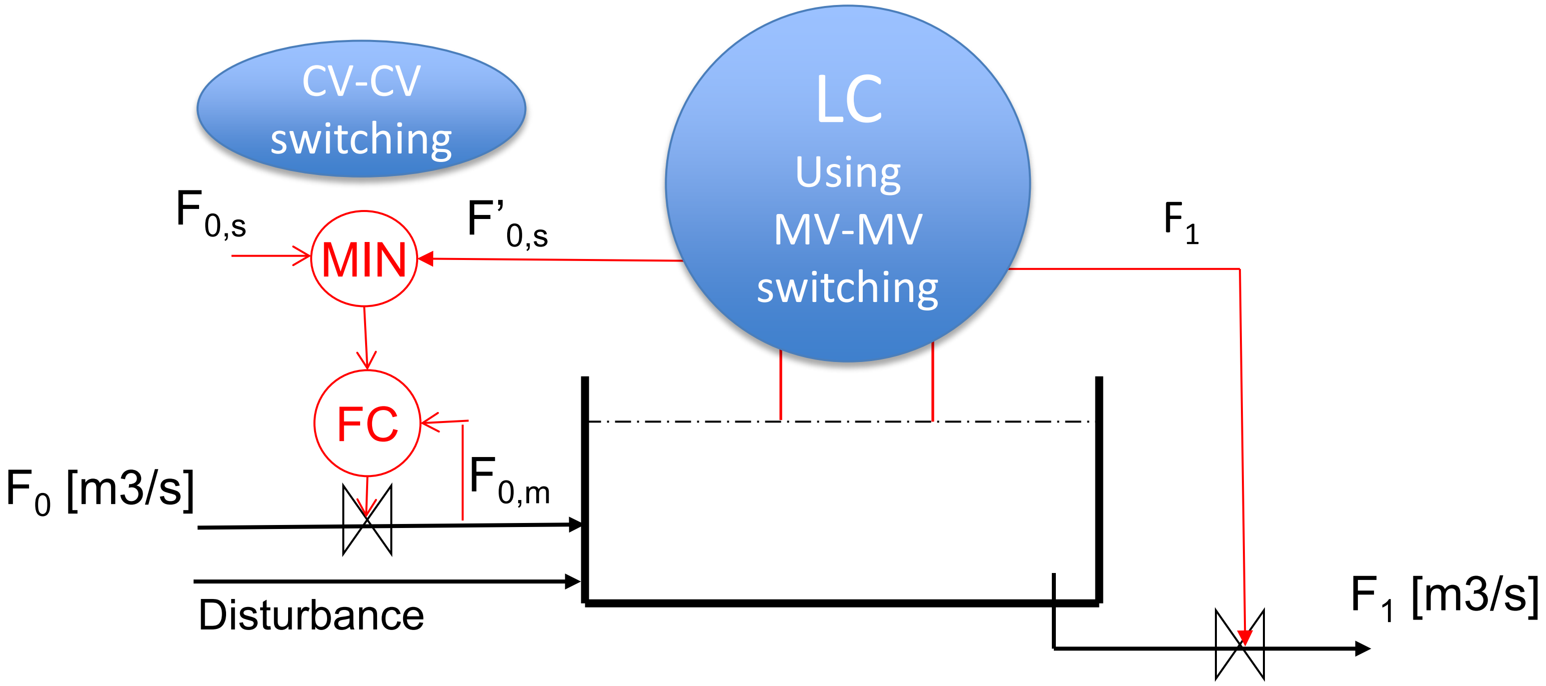


Example: Level control



What should we do if bottleneck at F1 (fully open valve, z1=1)?

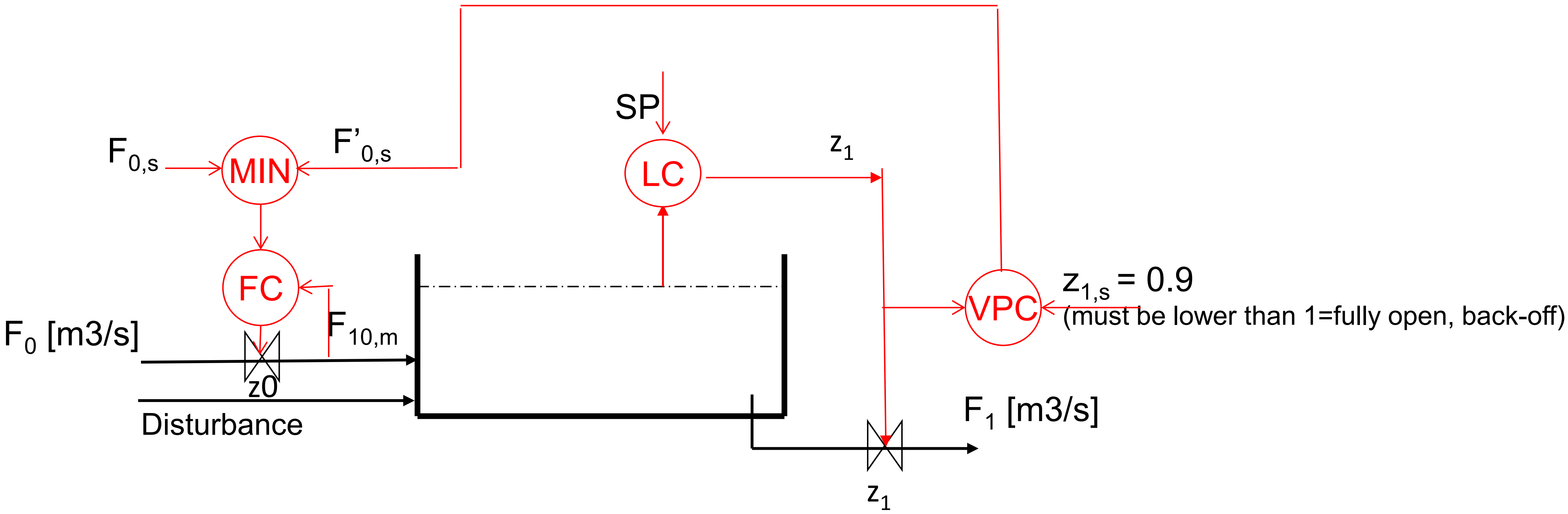
"Bidirectional inventory control" (Shinskey, 1981)



Three alternsatives for MV-MV switching

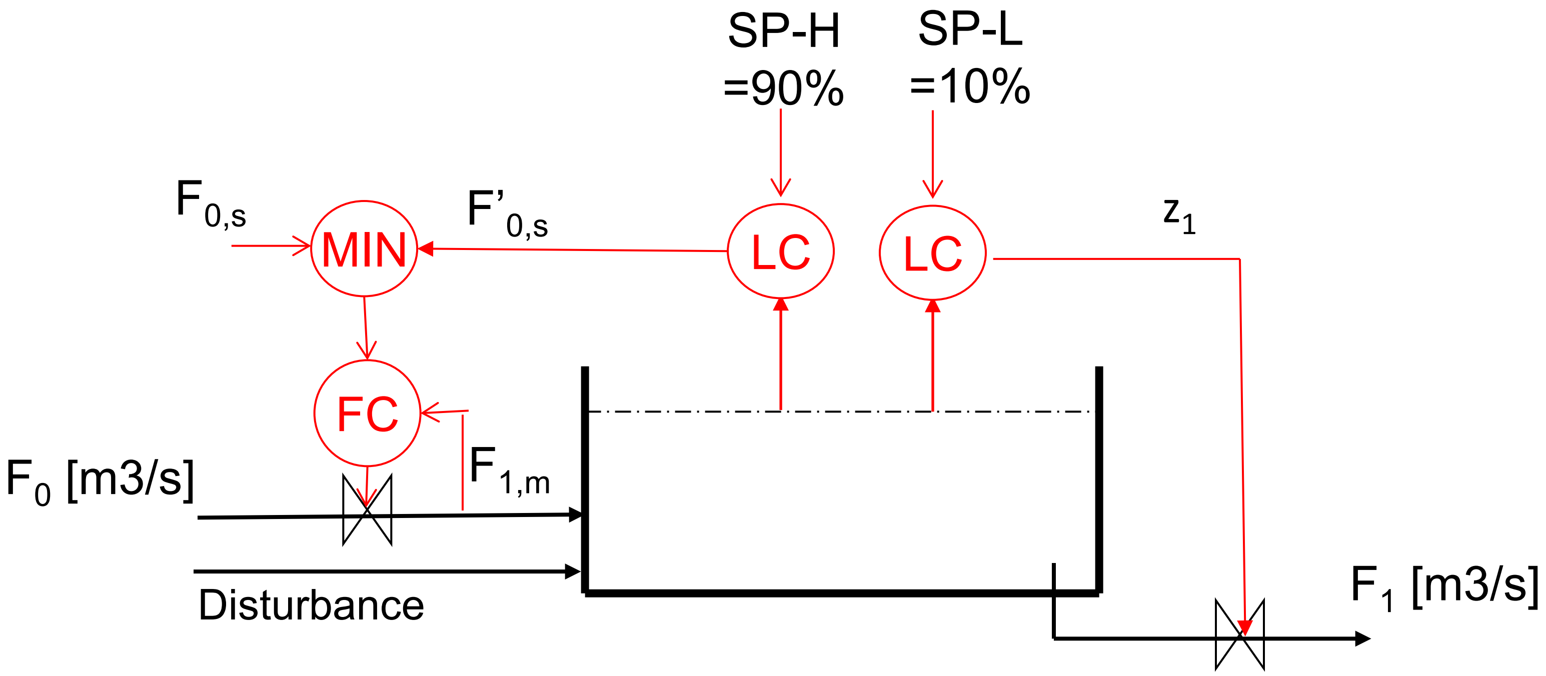
1. SRC (problem since F_{0s} varies)
2. Two controllers
3. VPC ("Long loop" for F1)

Alt. 3 MV-MV switching: VPC



VPC: “reduce inflow (F_0) if outflow valve (z_1) approaches fully open”

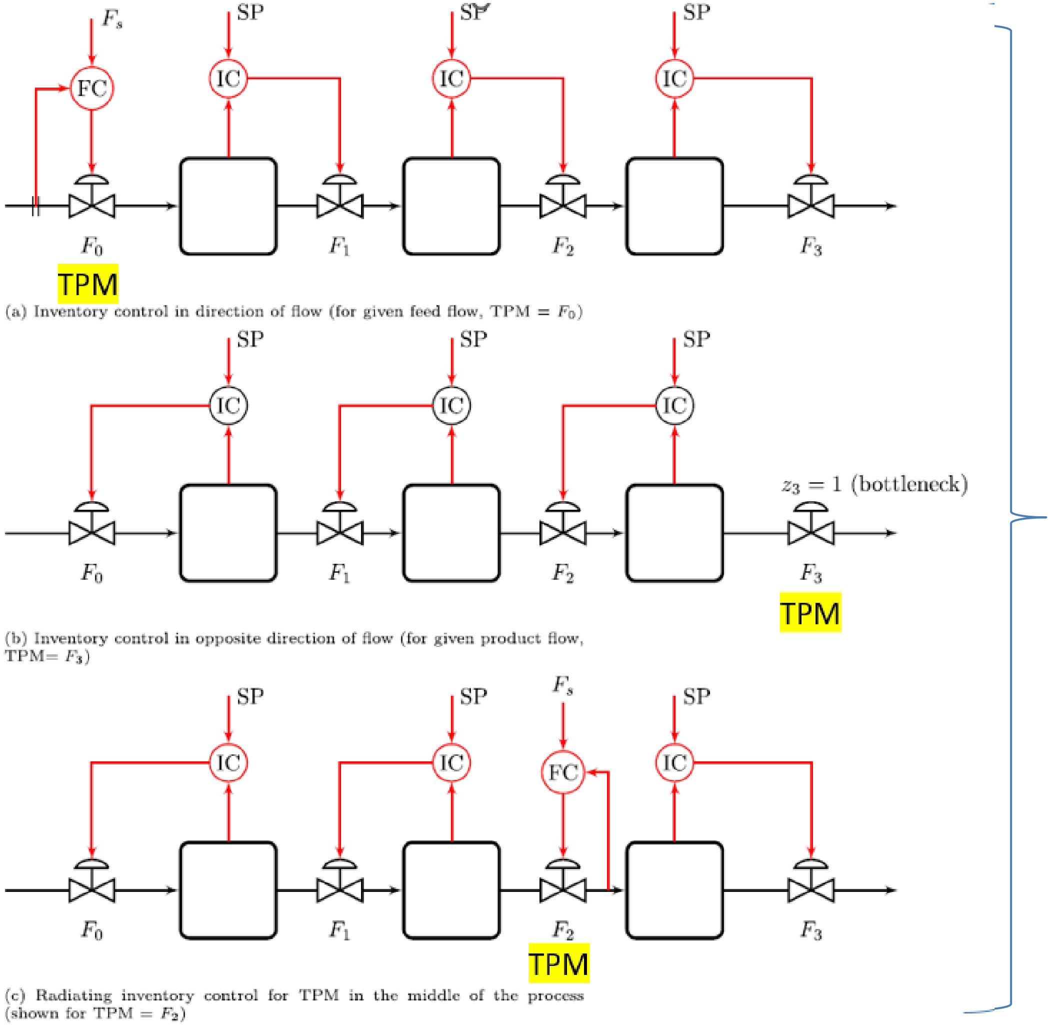
Alt. 2 MV-MV switching: Two controllers (recommended)



SP-L = low level setpoint
SP-H = high level setpoint

Extra benefit: Use of two setpoints is good for using buffer dynamically!!

Inventory control for units in series



Follows radiation rule

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

Need to reconfigure inventory loops if TPM moves

Generalization of bidirectional inventory control

Reconfigures TPM automatically with optimal buffer management!!

Maximize throughput:

$F_s = \infty$

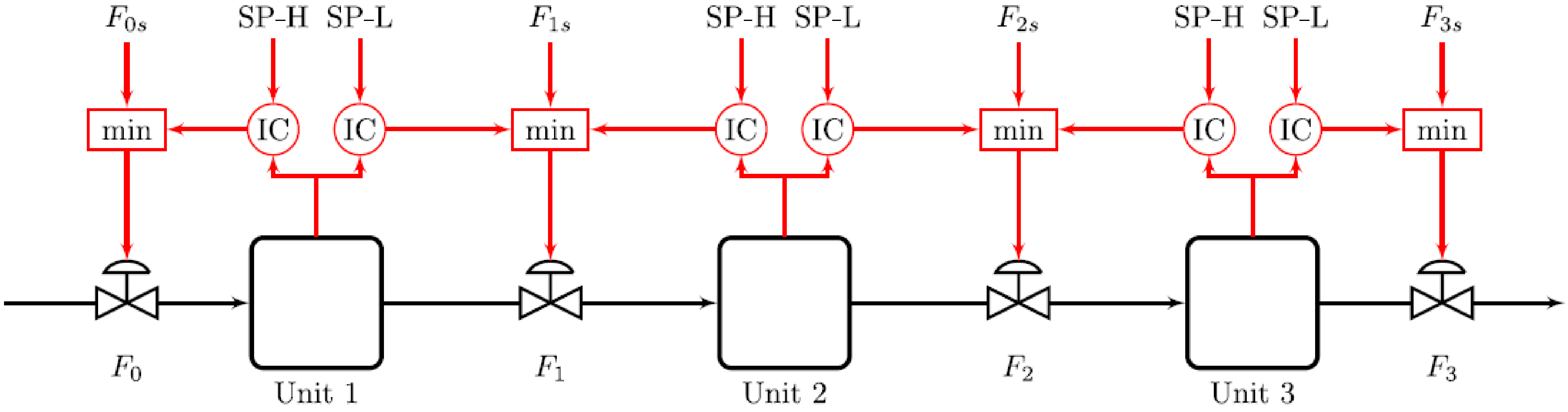


Fig. 36. Bidirectional inventory control scheme for automatic reconfiguration of loops (in accordance with the radiation rule) and maximizing throughput. Shinskey (1981) Zotica et al. (2022).
SP-H and SP-L are high and low inventory setpoints, with typical values 90% and 10%.
Strictly speaking, with 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

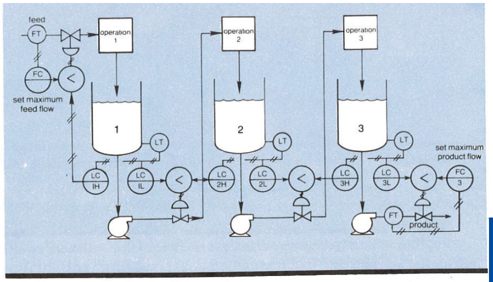
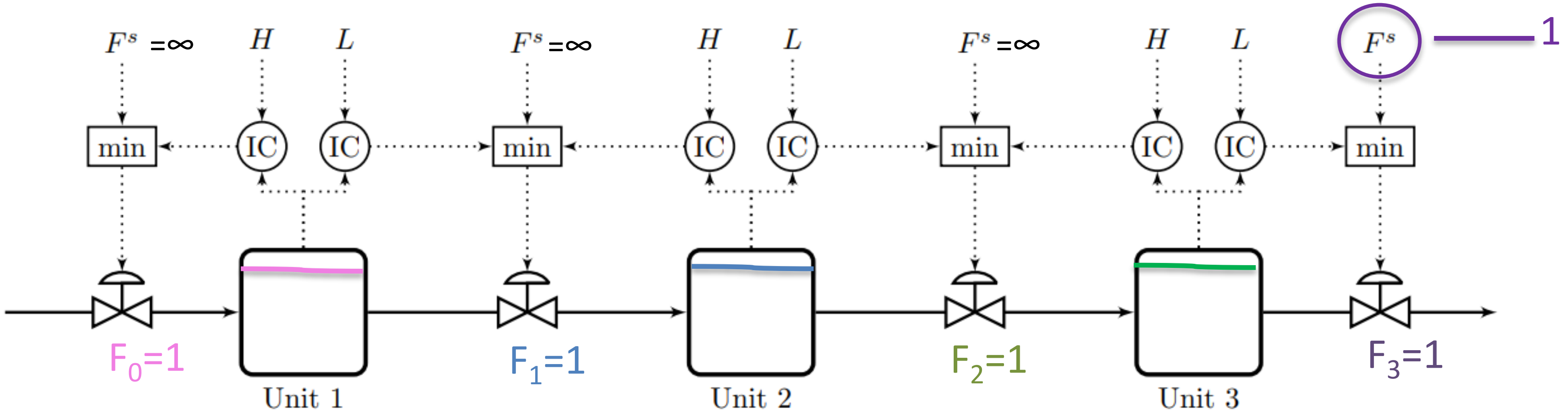
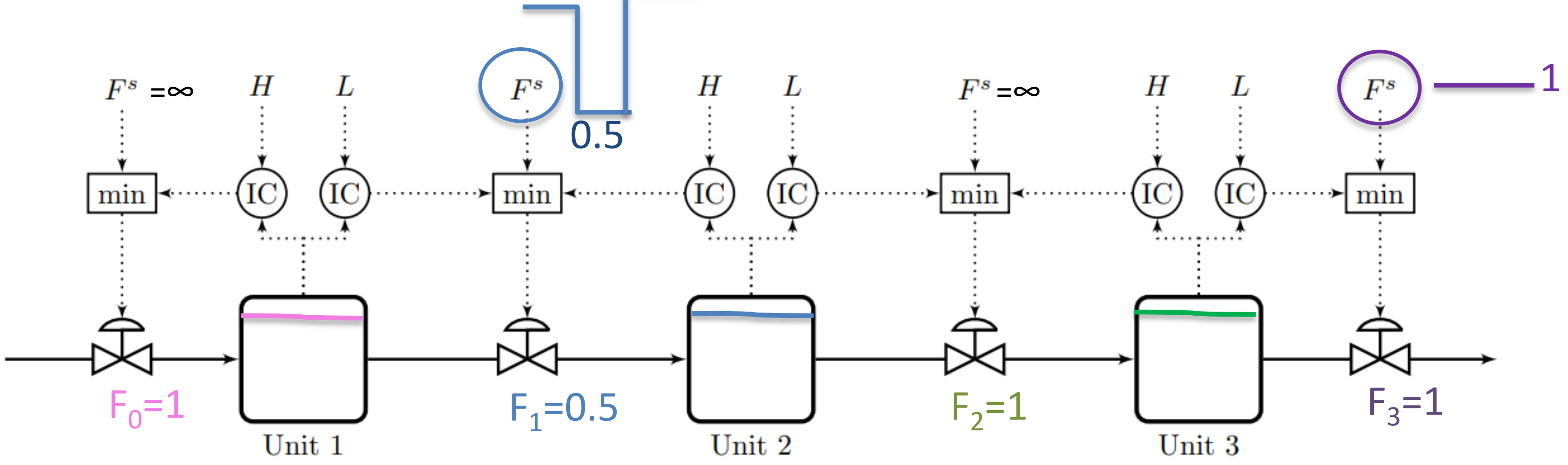


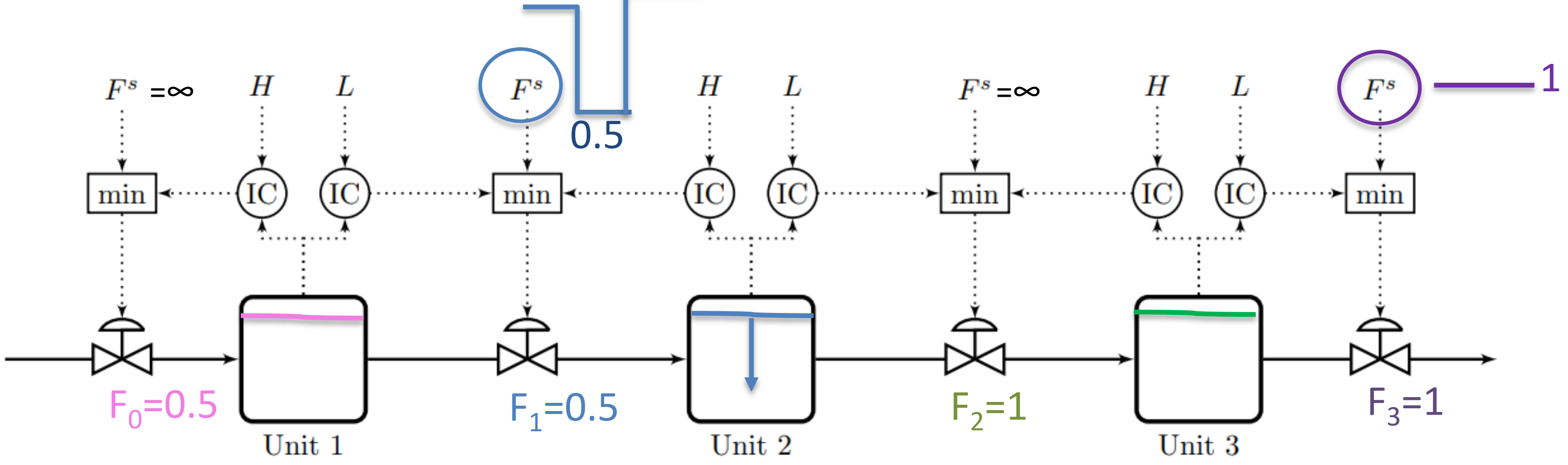
Fig. 3-7. Production rate can be set at either end of the process or constrained at any intermediate point without loss of inventory control.



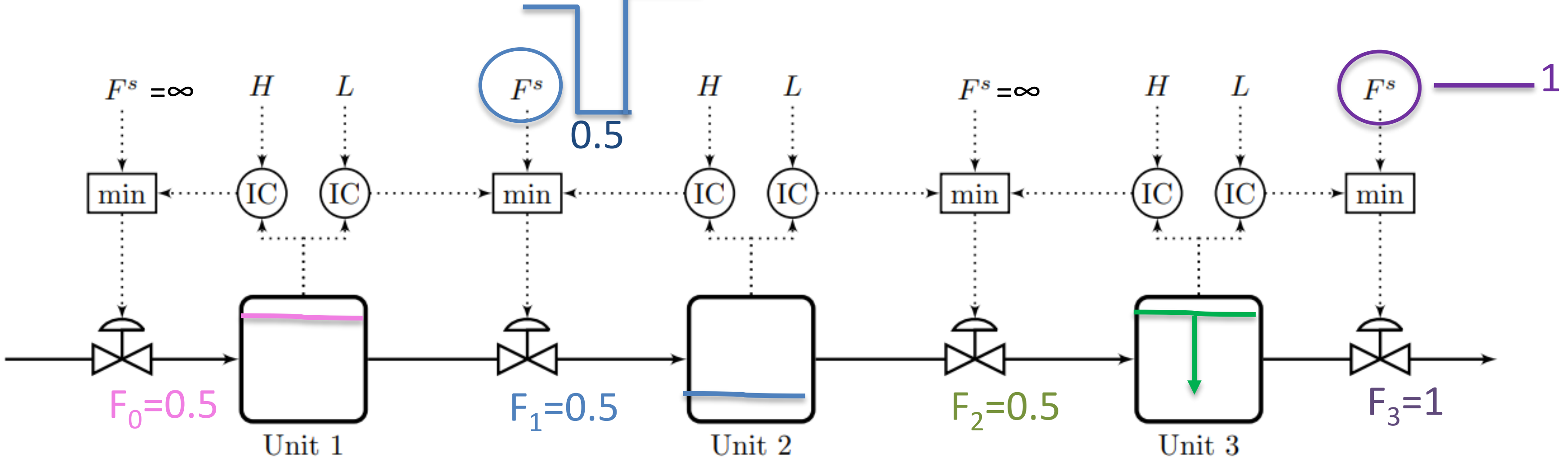
$F^s = \infty$
H
L
$F^s = \infty$
H
L
$F^s = \infty$
H
L
F^s
1
min
IC
IC
min
IC
IC
min
IC
IC
min
F_0=1
F_1=1
F_2=1
F_3=1
Unit 1
Unit 2
Unit 3



$F^s = \infty$
H
L
F^s
0.5
H
L
$F^s = \infty$
H
L
F^s
1
min
IC
IC
min
IC
IC
min
IC
IC
min
$F_0=1$
$F_1=0.5$
$F_2=1$
$F_3=1$
Unit 1
Unit 2
Unit 3



$F^s = \infty$
H
L
F^s
0.5
H
L
$F^s = \infty$
H
L
F^s
1
min
IC
IC
min
IC
IC
min
IC
IC
min
$F_0=0.5$
$F_1=0.5$
$F_2=1$
$F_3=1$
Unit 1
Unit 2
Unit 3



F^s =∞
H
L
F^s
0.5
H
L
F^s =∞
H
L
F^s
1
min
IC
IC
min
IC
IC
min
IC
IC
min
F_0=0.5
Unit 1
F_1=0.5
Unit 2
F_2=0.5
Unit 3
F_3=1

NTNU

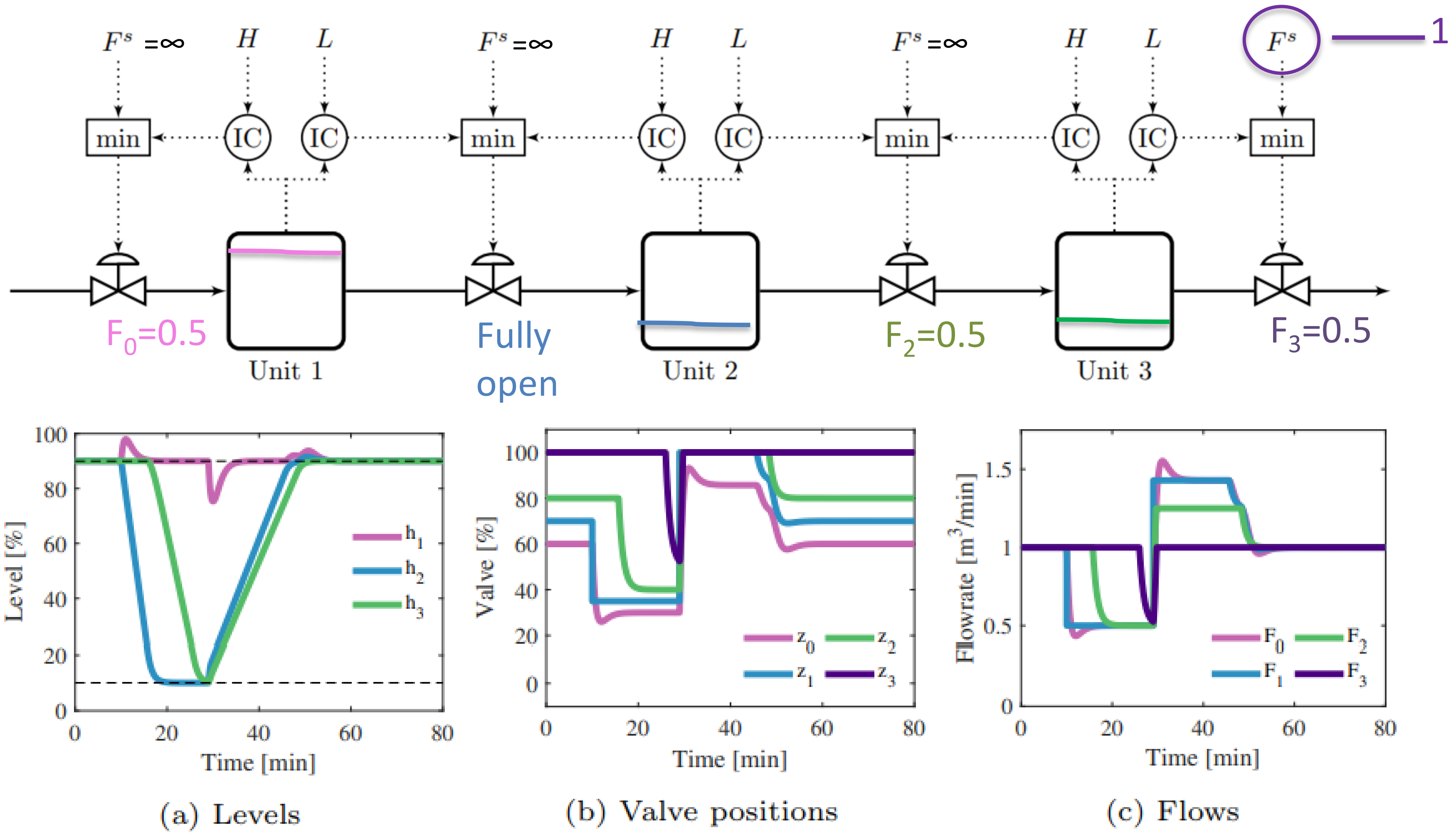
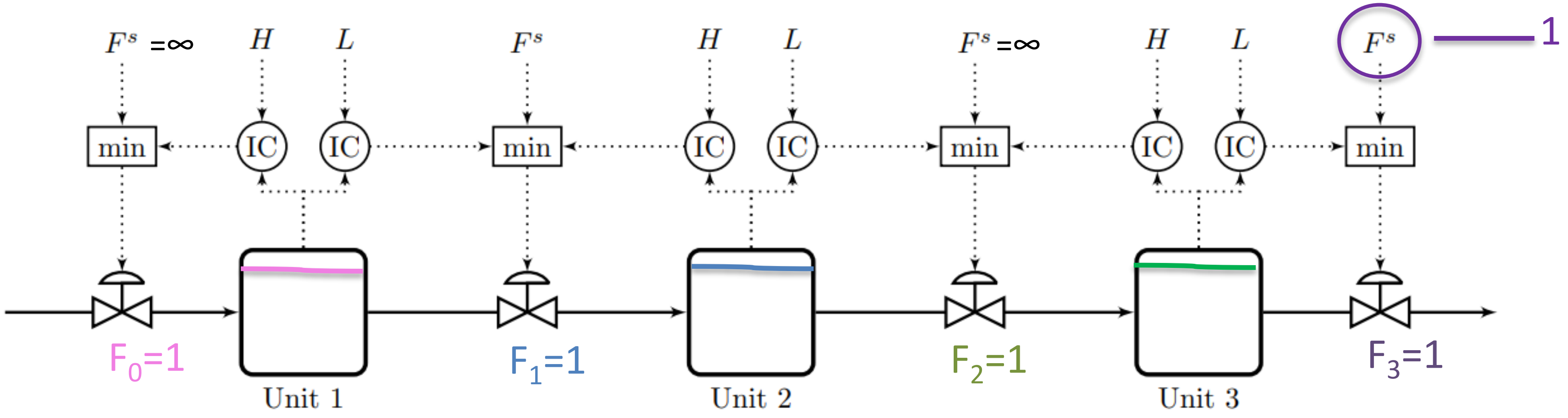


Figure 12: Simulation of a 19 min temporary bottleneck in flow F_1 for the control structures in Fig. 3d with the TPM downstream of the bottleneck.

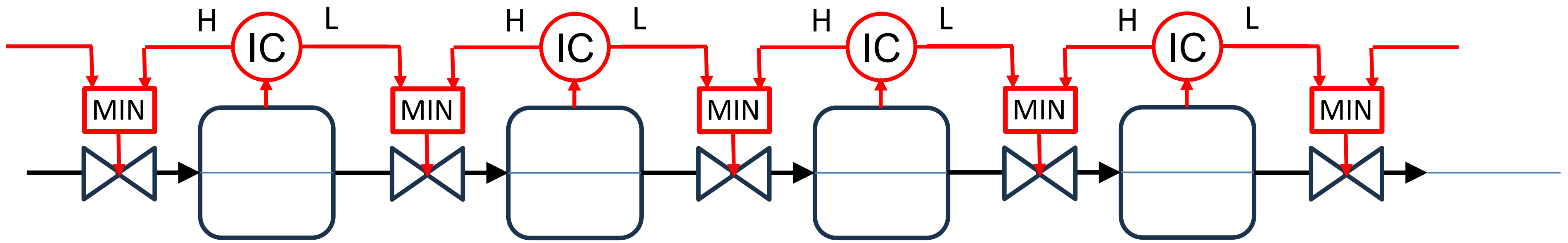


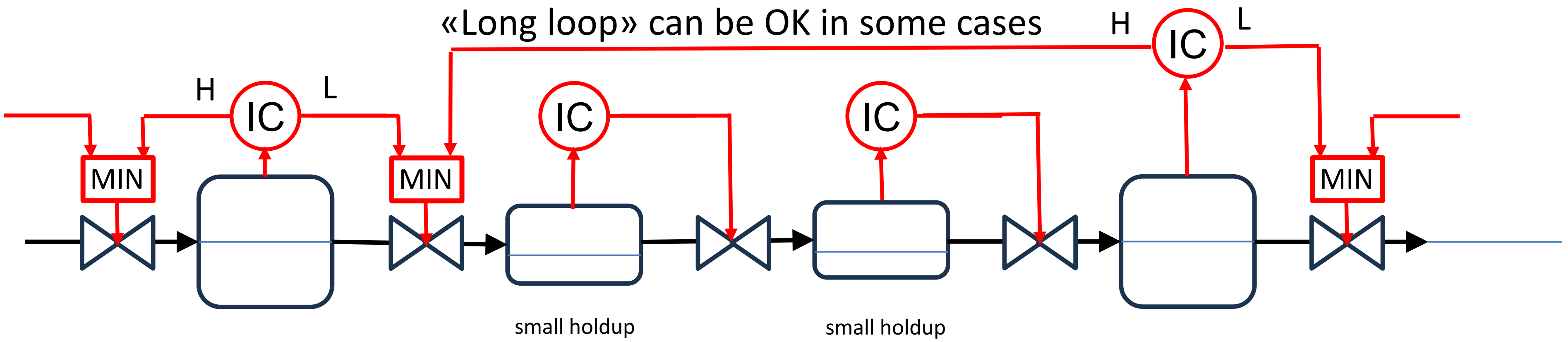
Challenge: Can MPC be made to do his? Optimally reconfigure loops and find optimal buffer?

- Yes, possible with standard setpoint-based MPC if we use
  - Trick: All flow setpoints = infinity (unachievable setpoint)
- What about Economic MPC? Cannot do it easily; may try scenario-MPC

NTNU

Don’t need bidirectional control on all units



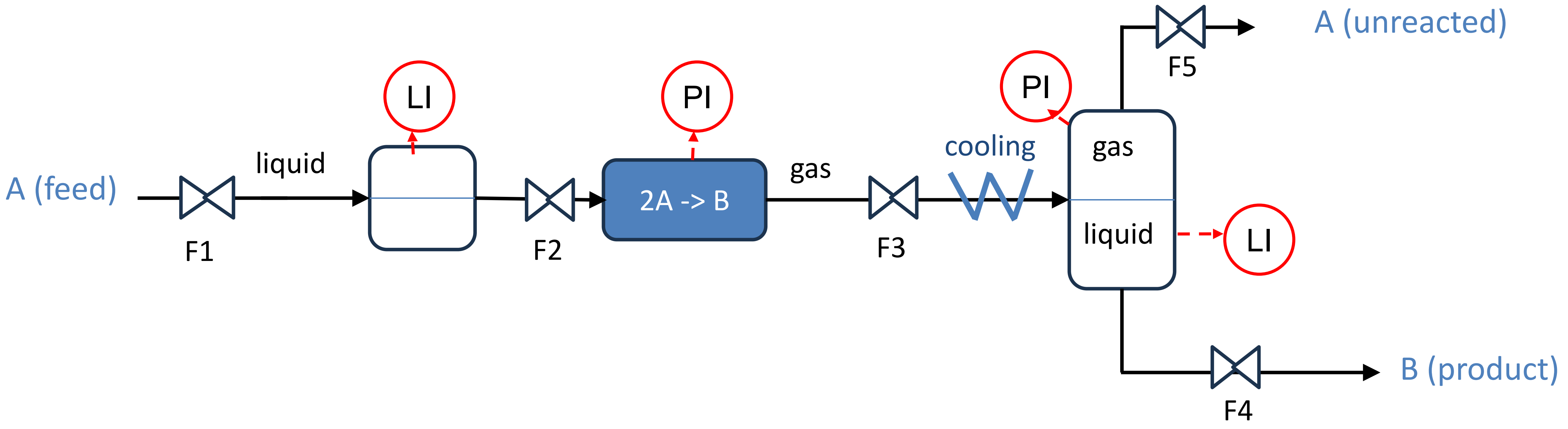


Important insight

- Many problems: Optimal steady-state solution always at constraints
- In this case optimization layer may not be needed
  - if we can identify the active constraints and control them using selectors

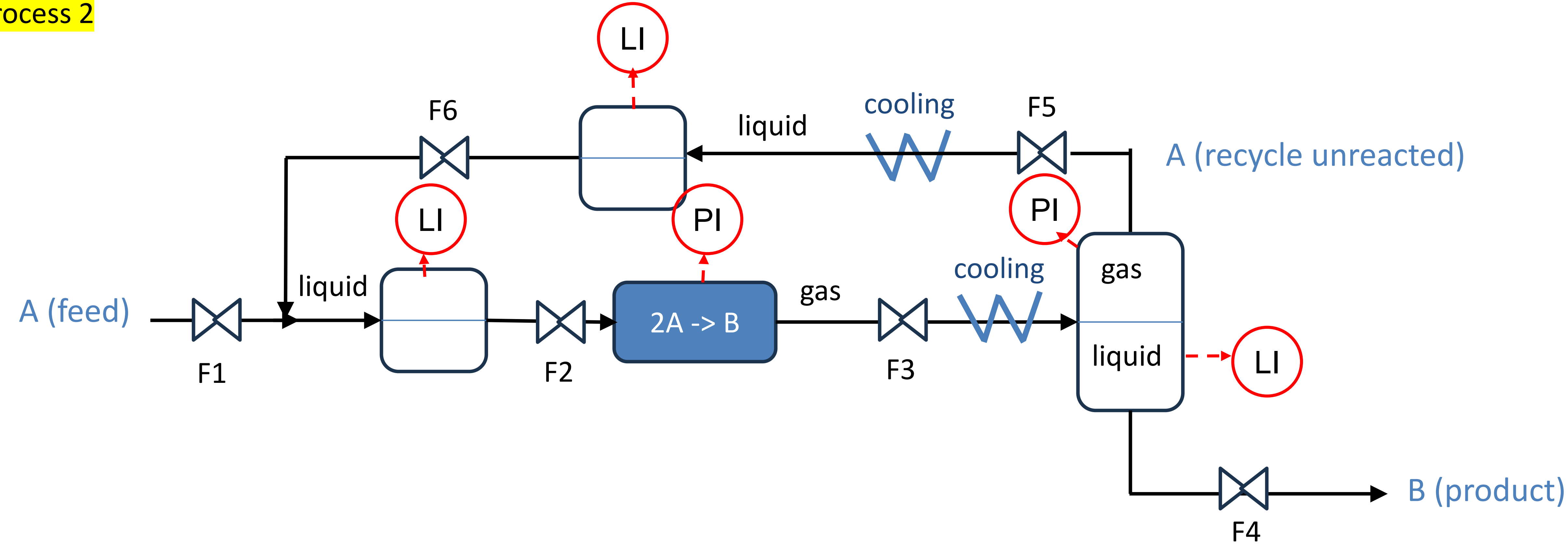
Control of chemical processes with recycle

Process 1



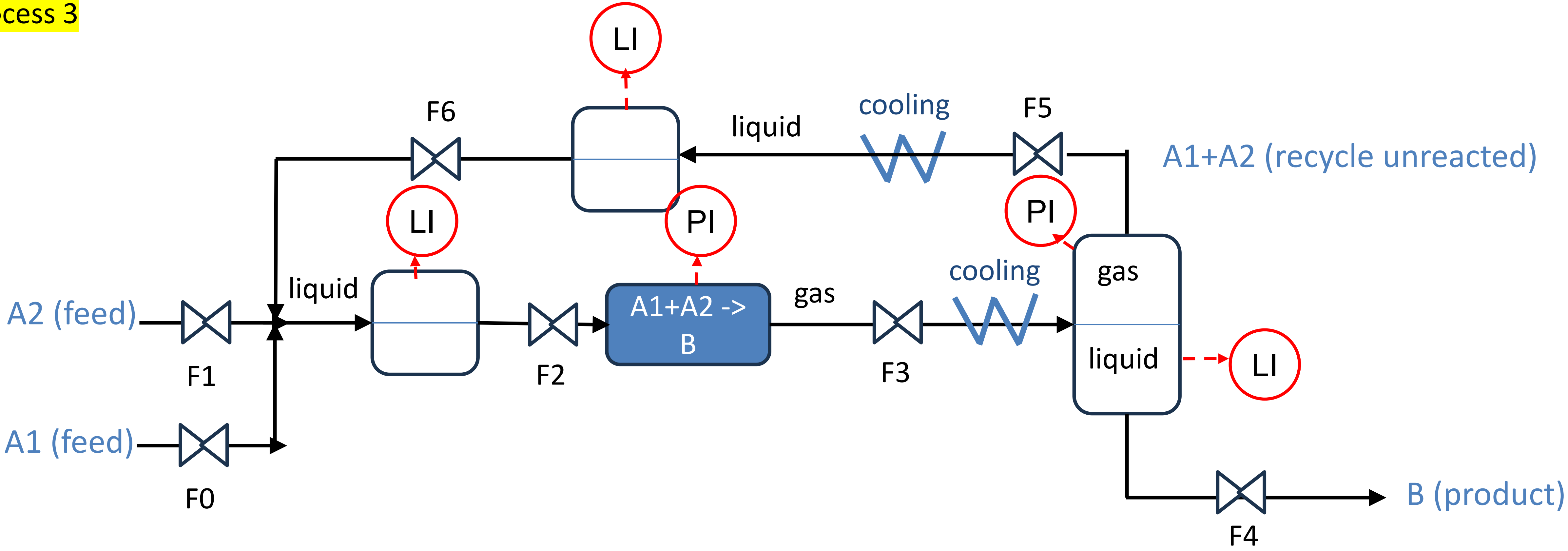
Exothermic reaction

Process 2



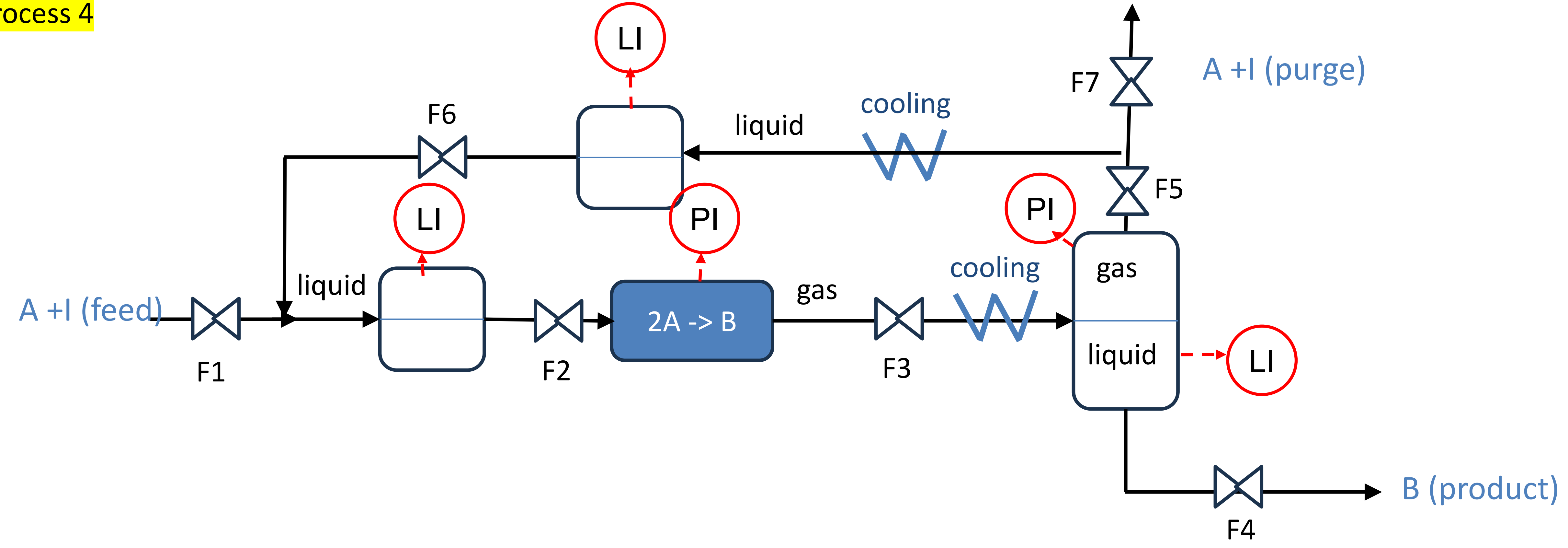
Comment: Valve F5 may not be necessary. Could use valve on cooling instead

Process 3



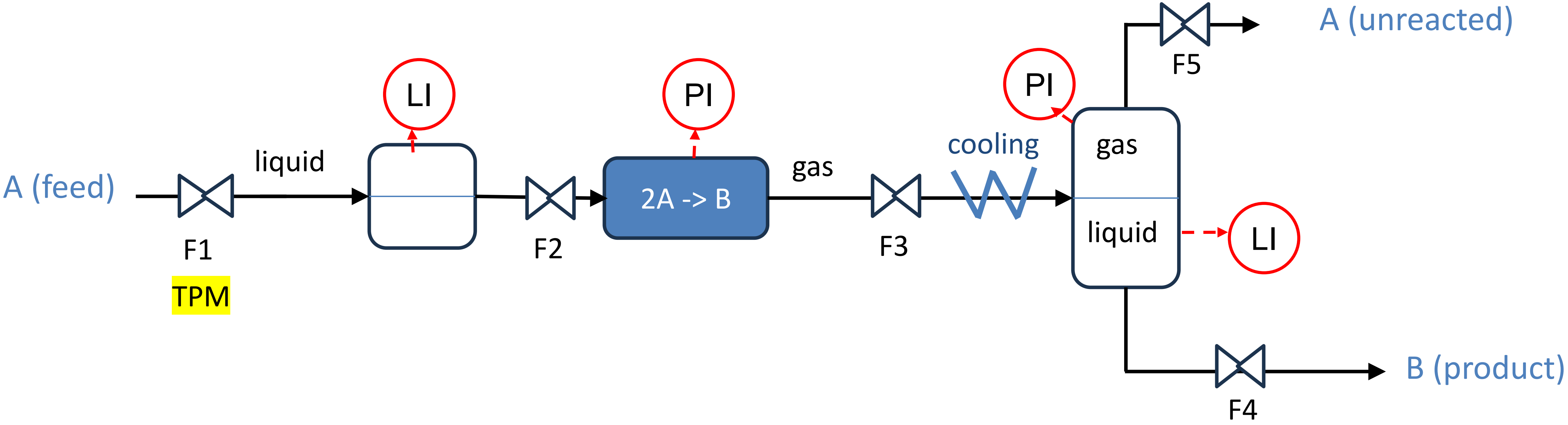
LI
F6
liquid
cooling
F5
A1+A2 (recycle unreacted)
LI
PI
PI
liquid
A2 (feed)
F1
A1+A2 -> B
gas
cooling
gas
liquid
LI
F2
F3
A1 (feed)
F0
B (product)
F4

Process 4


LI
F6
liquid
cooling
F7
A +I (purge)
F5
LI
PI
PI
liquid
gas
cooling
gas
A +I (feed)
2A -> B
liquid
LI
F1
F2
F3
B (product)
F4

Process 1

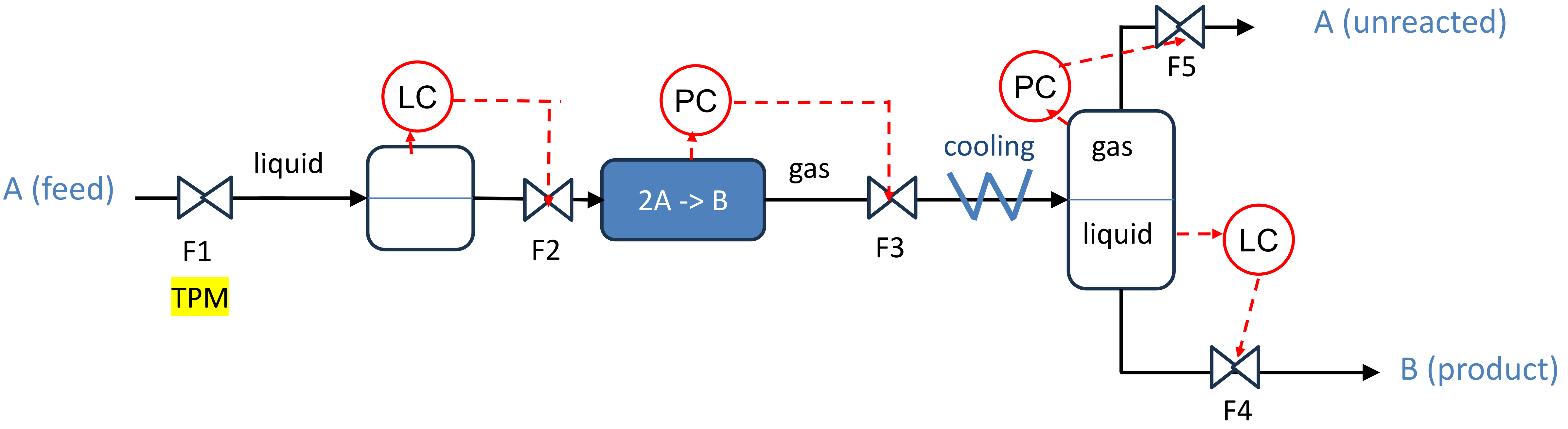
Control



Exothermic reaction

Process 1

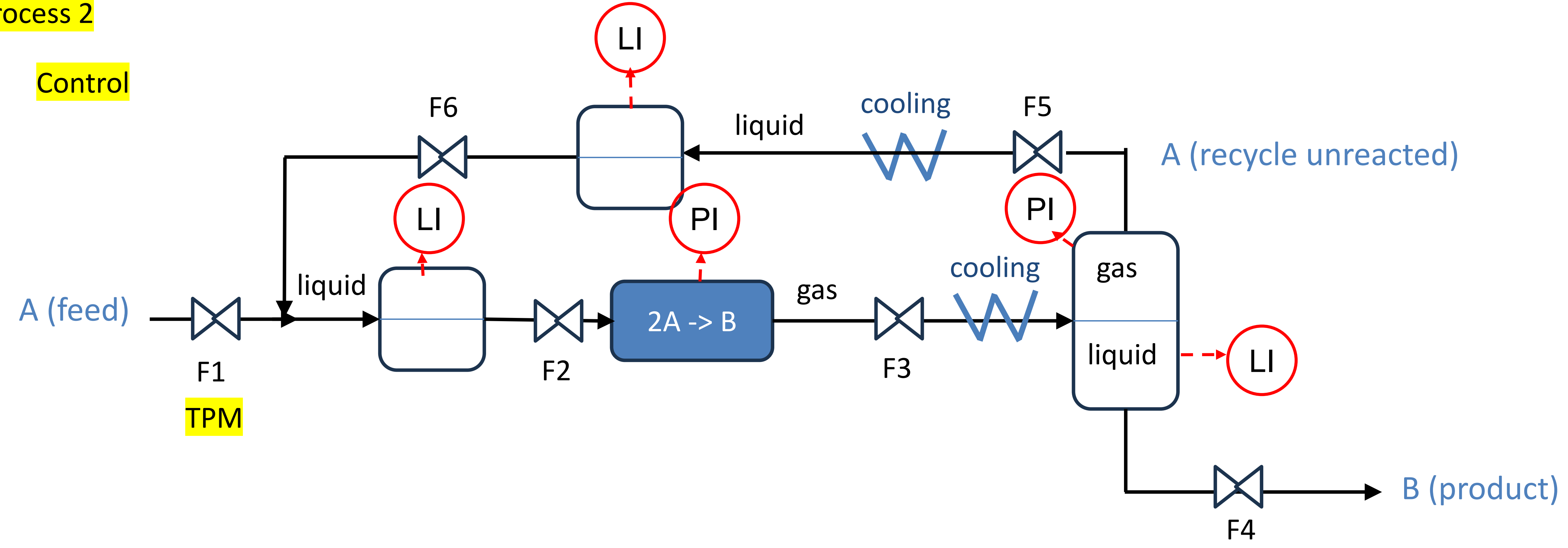
Control



Exothermic reaction

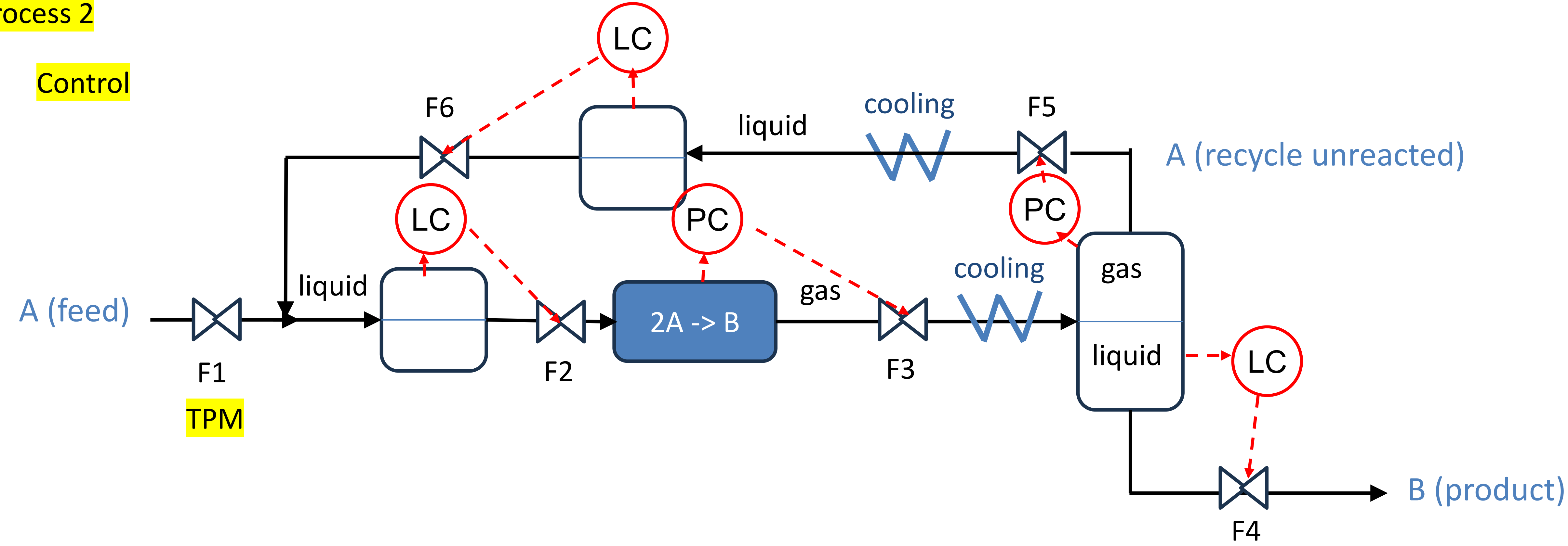
Process 2

Control



Process 2

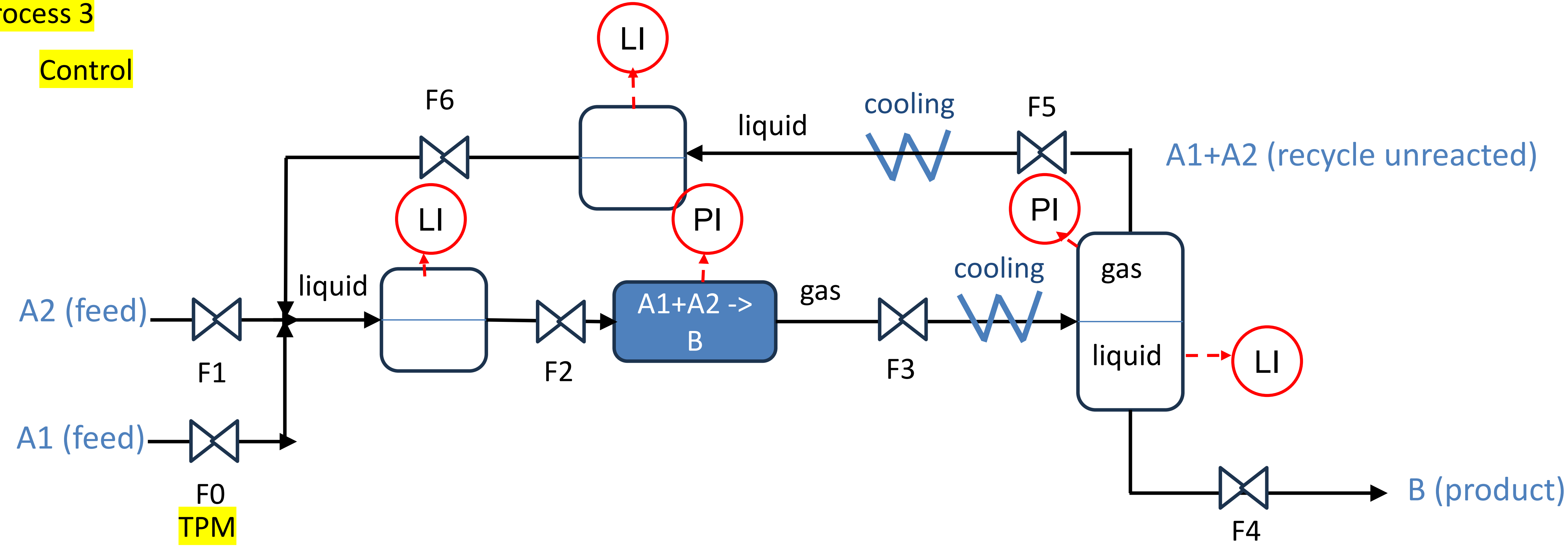
Control



TPM

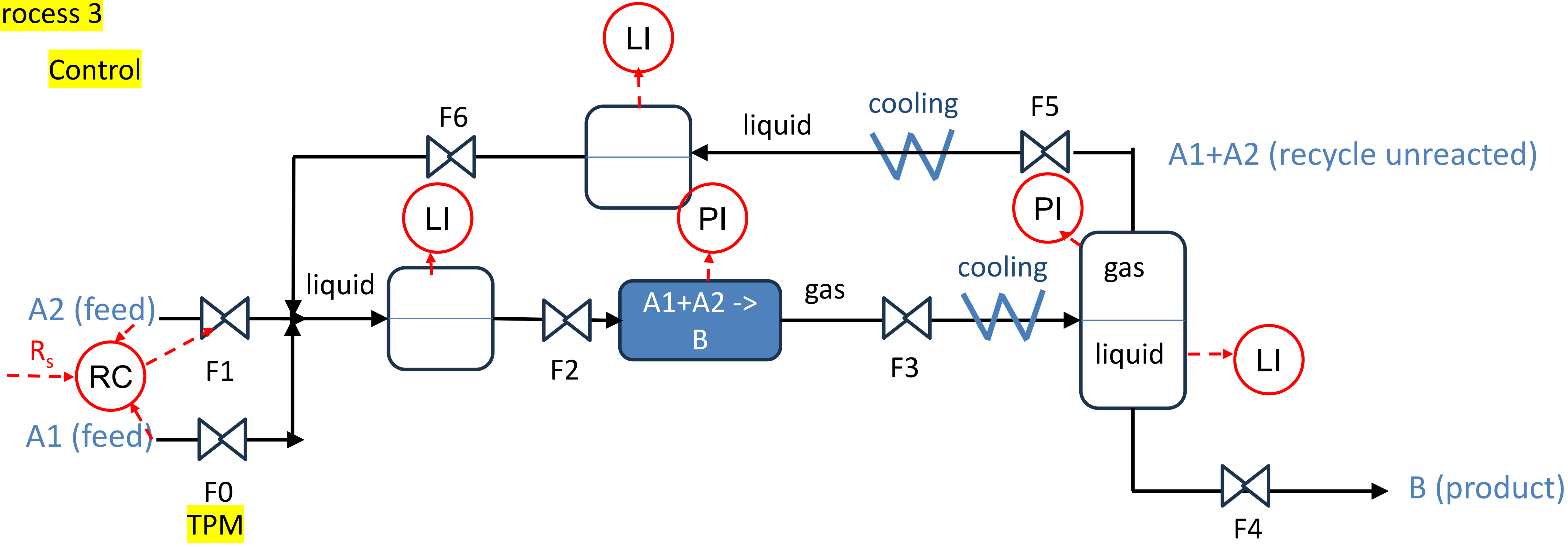
Process 3

Control



Process 3

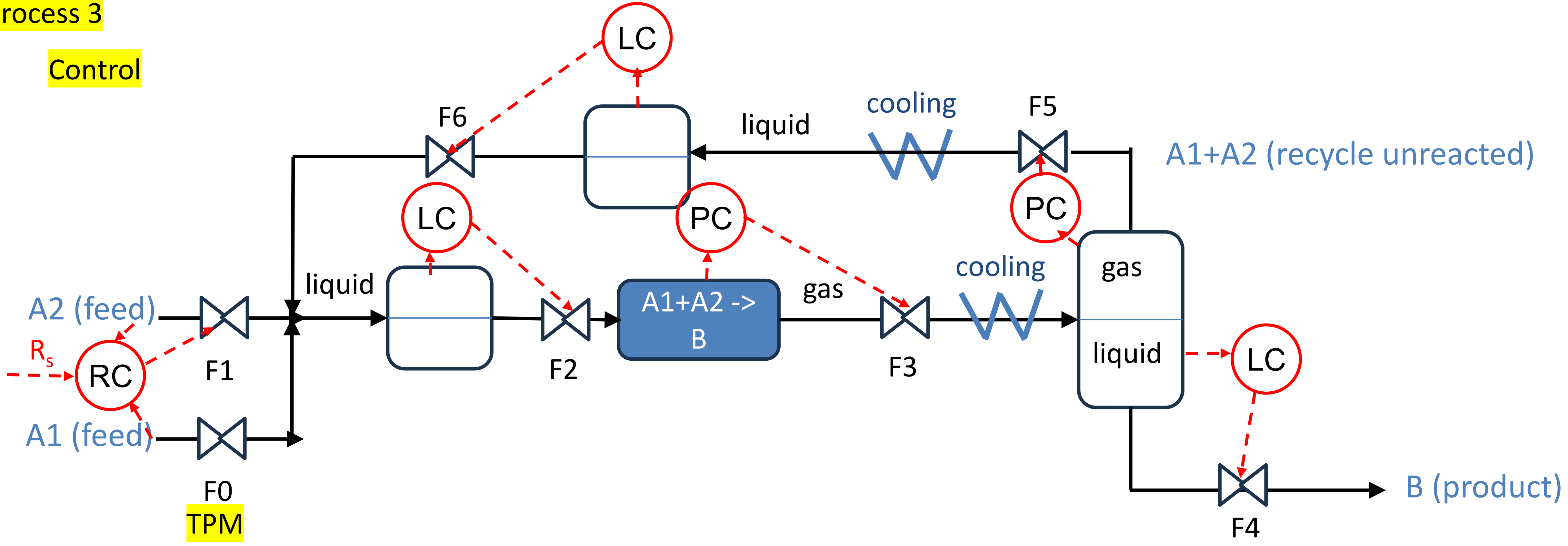
Control



The ratio control can be done in different ways.
It requires two flow measurements (F0, F1)
One of the flows is the TPM

Process 3

Control

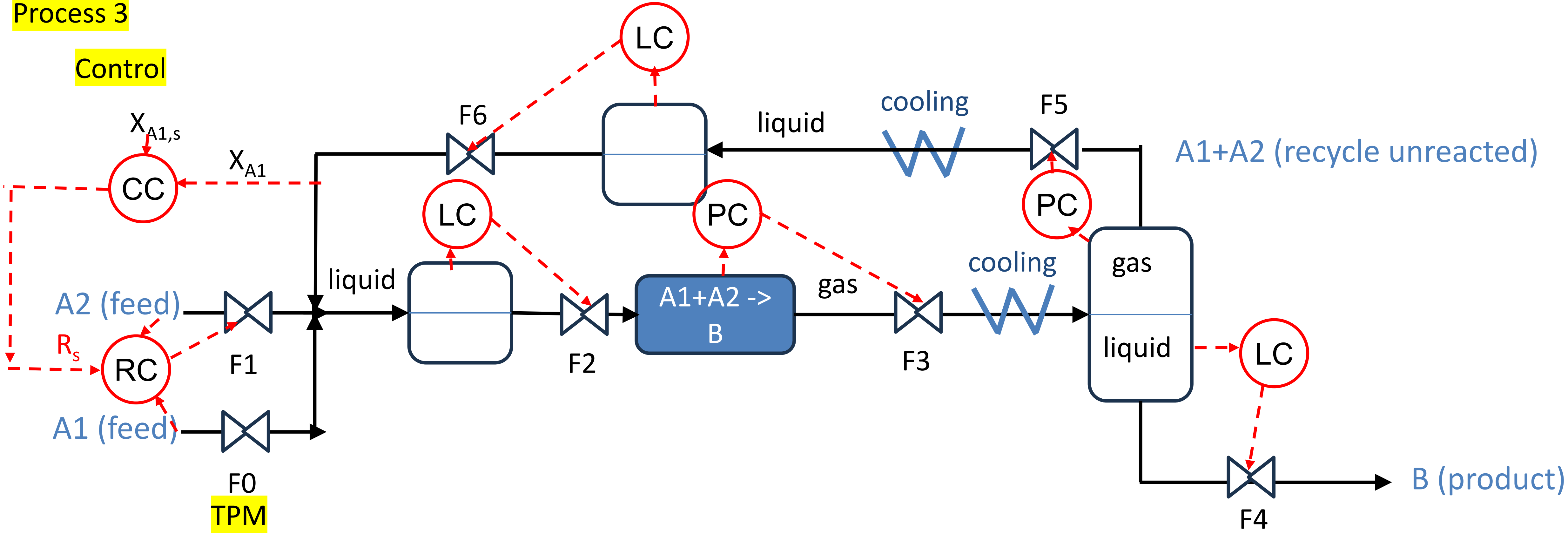


Will this work?

No, it's not possible to feed exactly the same amount of A1 and A2 without feedback correction

Process 3

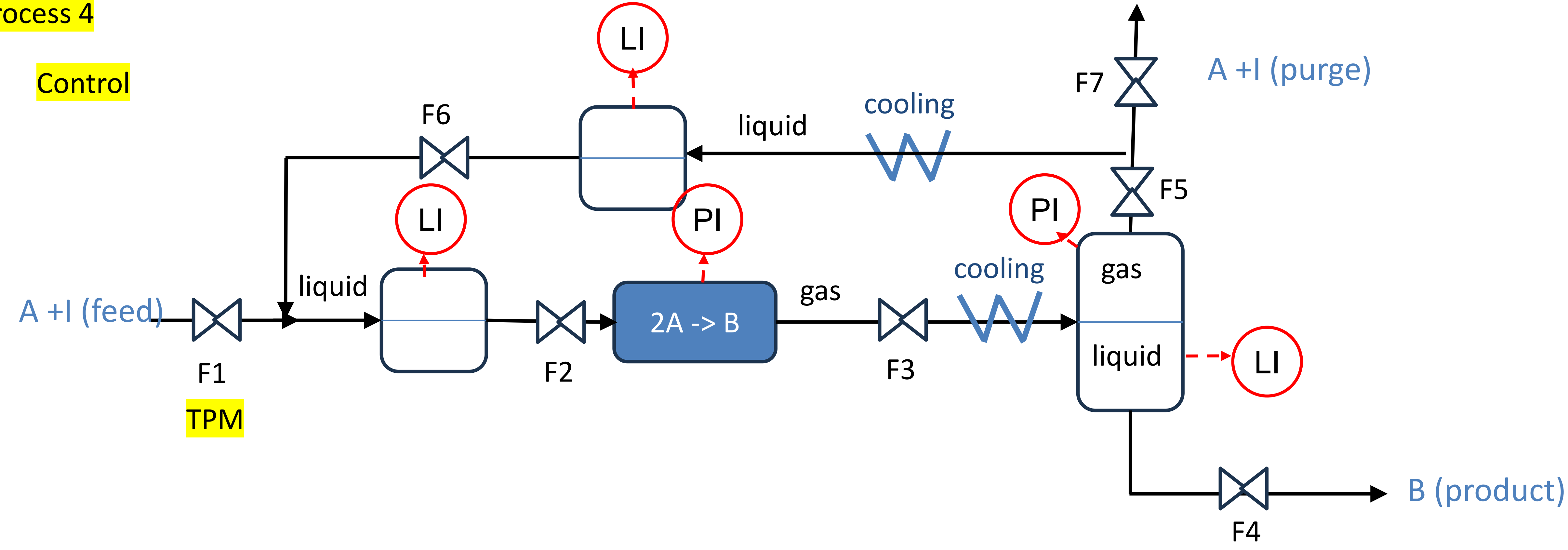
Control



With composition control of A1 (or A2).
This works!

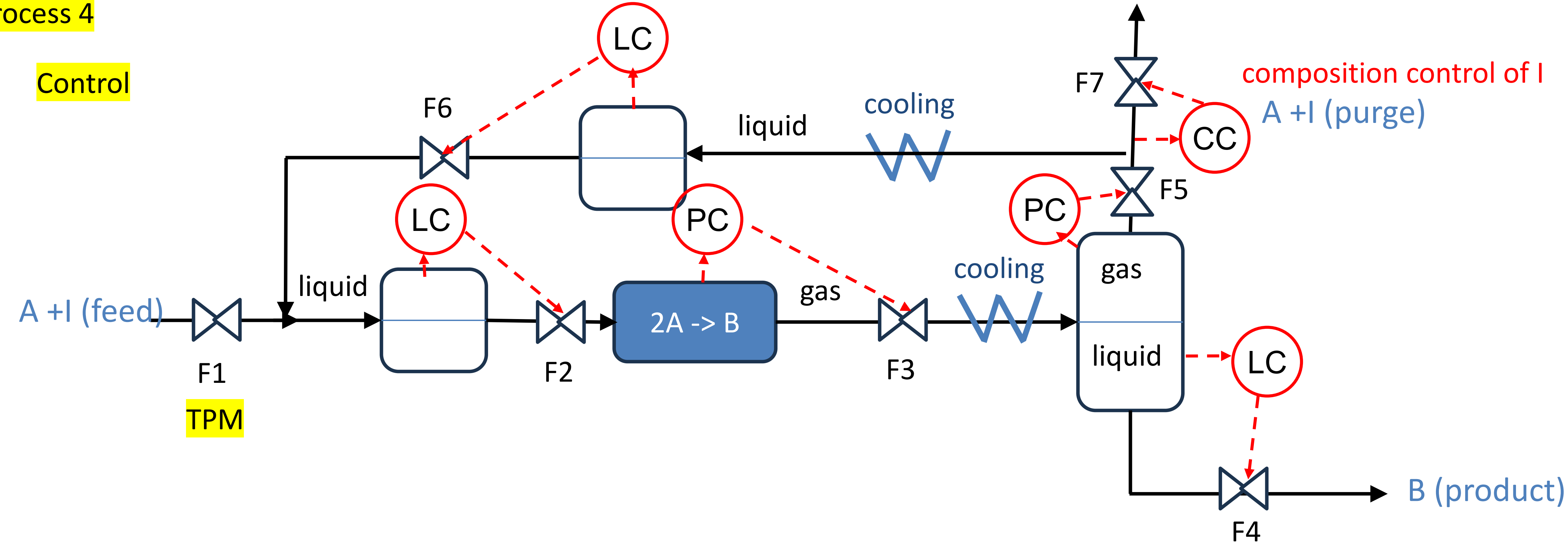
Process 4

Control



Process 4

Control

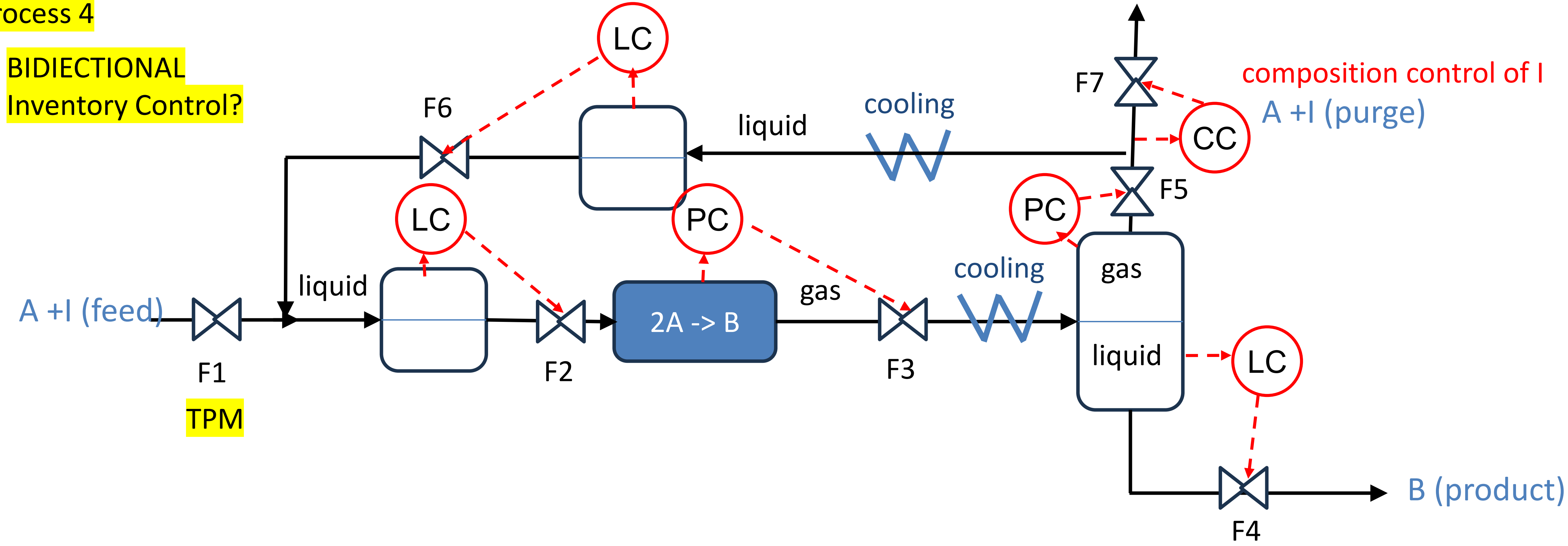


TPM

NTNU

Process 4

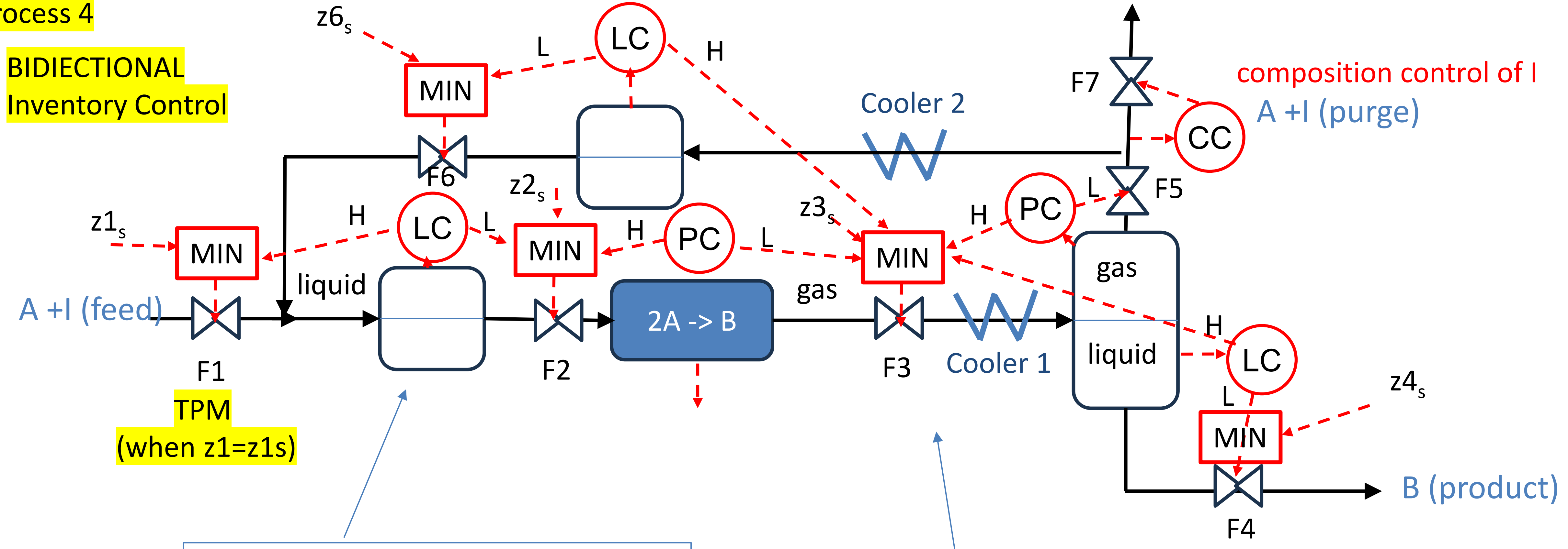
BIDIECTIONAL
Inventory Control?



TPM

Process 4

BIDIECTIONAL Inventory Control



This LC is two controllers which both control level.

- The one with outflow F2 as the MV has a Low level setpoint
- The one with inflow F1 as the MV has a High level setpoint

3 H-setpoints go to this MIN-selector:
Reduce F3 if

1. too high pressure (cooler 2 max),
2. too much gas (high level) (F6 limiting)
3. too much liquid (high level) (F4 limiting)

Example bidirectional inventory control

AIChE

Economic Plantwide Control of the Ethyl Benzene Process

Rahul Jagtap, Ashok S Pathak, and Nitin Kaistha
Dept. of Chemical Engineering, Indian Institute of Technology Kanpur, Kanpur 208016,Uttar Pradesh, India

DOI 10.1002/aic.13964
Published online December 10, 2012 in Wiley Online Library (wileyonlinelibrary.com).

A1: Benzene
A2: Ethylene
B: Ethylbenzene (product)
C: Diethylbenzene (undersired)
A1+A2 → B
B + A2 → C
C + A1 → 2B

LS = MIN

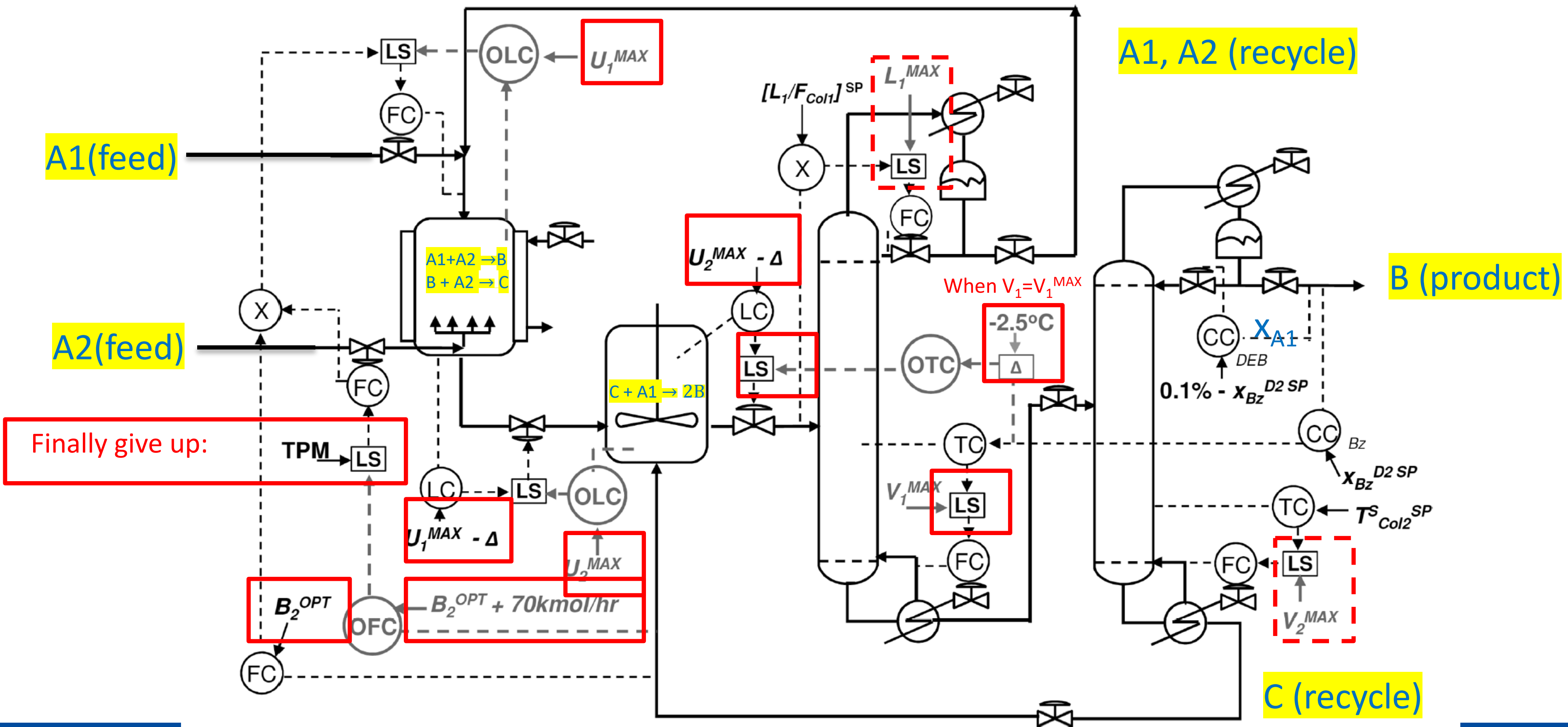


Figure 7. CS2 with overrides for handling equipment capacity constraints.