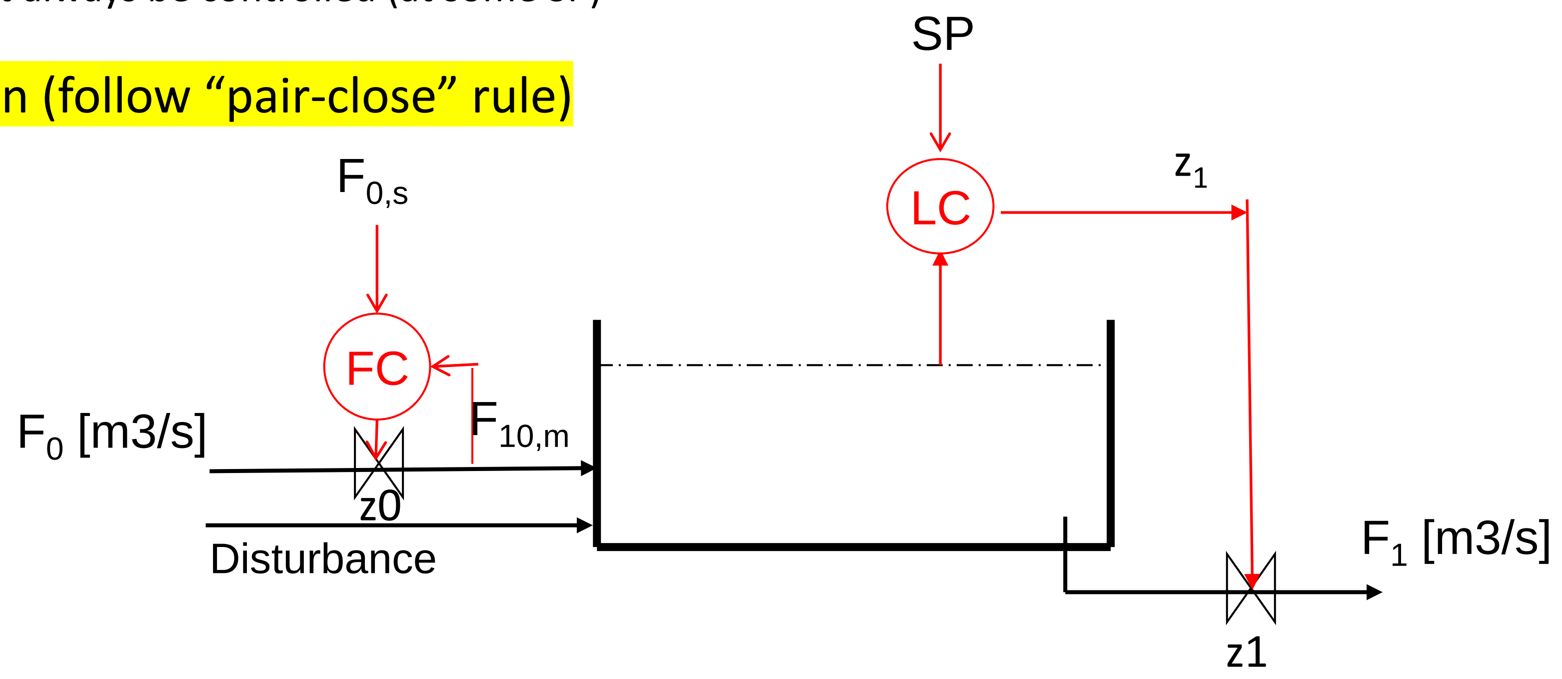
Part 5: inventory control +

Example : Level control

MV1 = z0 (inflow valve position)
MV2 = z1 (outflow valve position) (likely to saturate)
CV1 = F0 (inflow): Should be controlled at setpoint $F_{0,s}$ (if possible)
CV2 = level: must always be controlled (at some SP)

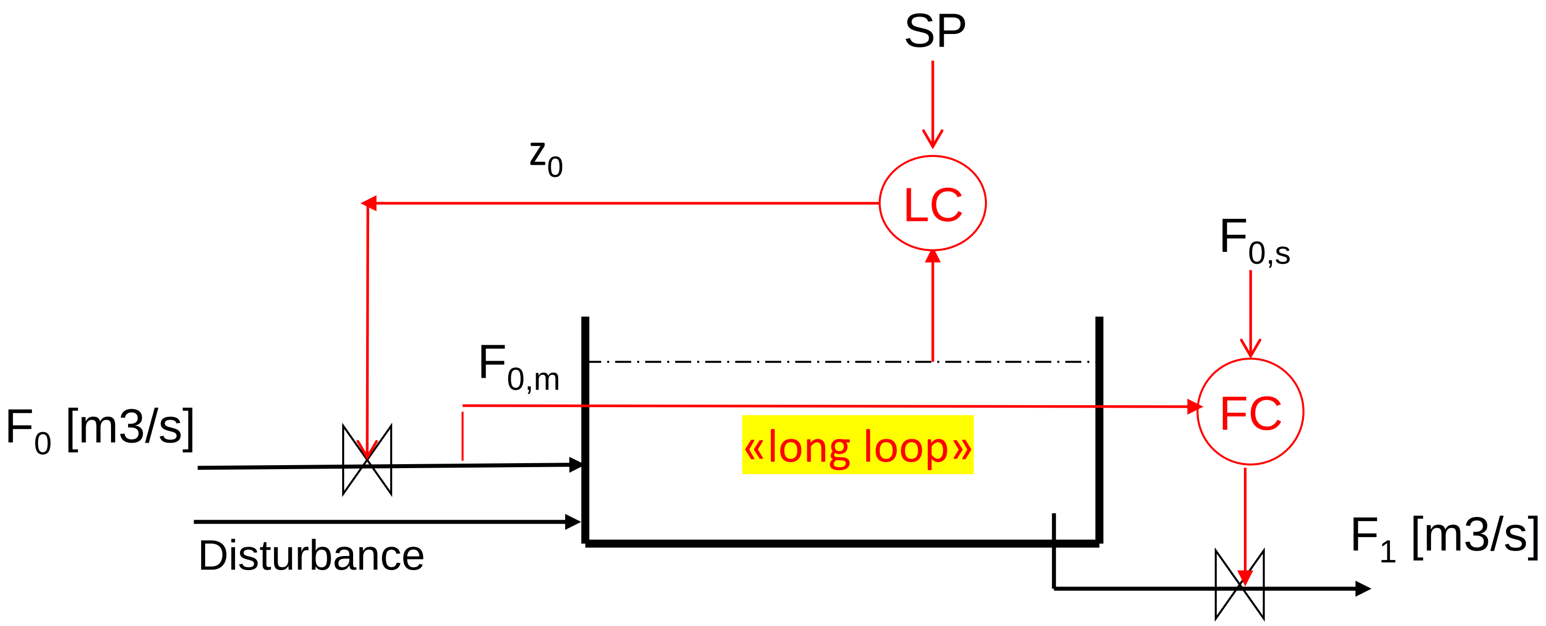
Nominal design (follow “pair-close” rule)



Problem: outflow-valve may saturate at fully open (z1=1) and then we lose level control
Note: We did not following the “input saturation rule” which says:
Pair MV that may saturate (z1) with CV that can be given up (F0)

This gives simple MV-CV switching (if z2 saturates at fully open)

Reverse pairing (follows "input saturation rule"):



BUT with Reverse pairing: Get "long loop" for F0
In addition: loose control of y2= level if z0 (F0-valve) saturates

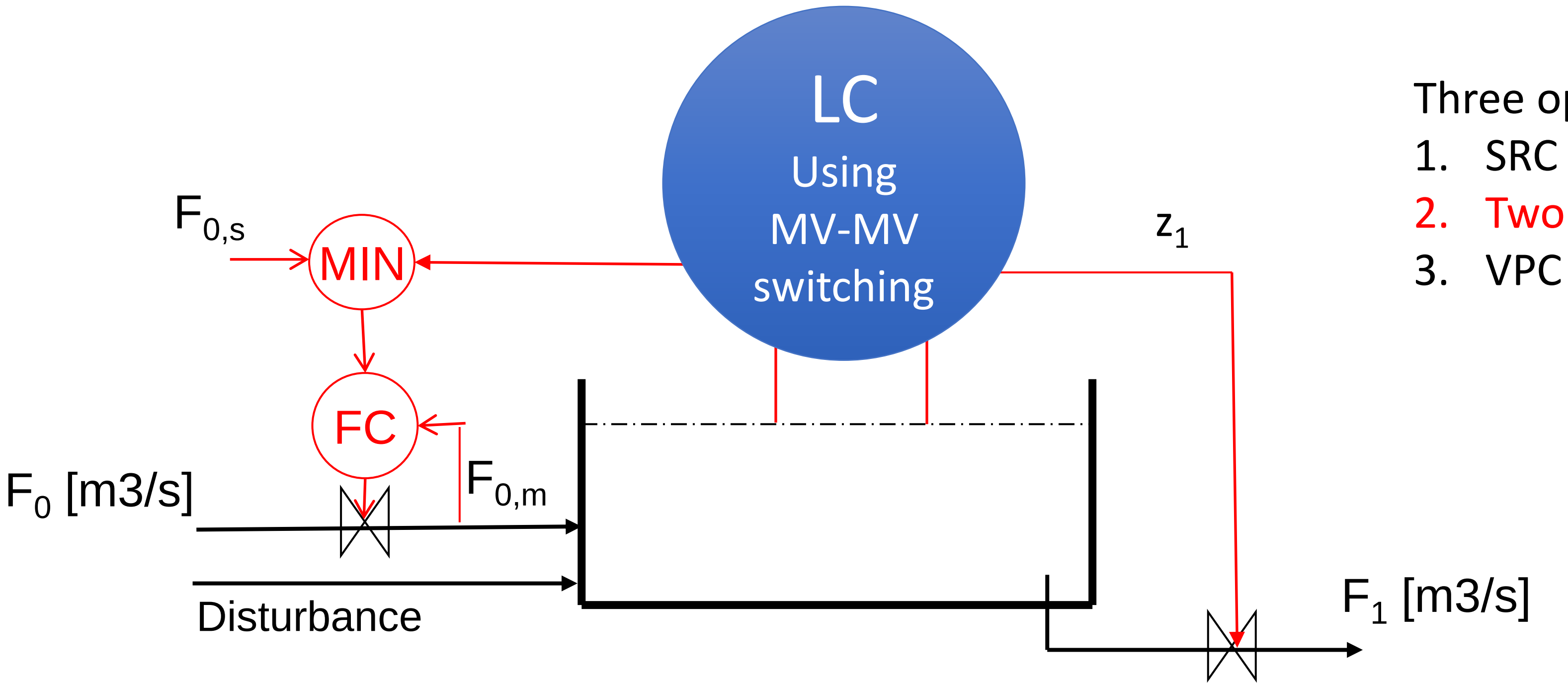
«Long loop» = Works through other loops

This is complex MV-CV switching

Alternative solution: Follow "Pair close"-rule and use Complex MV-CV switching.

When z1 saturates at max, use the other MV (z0) for level control and give up controlling F0

Get: "Bidirectional inventory control"



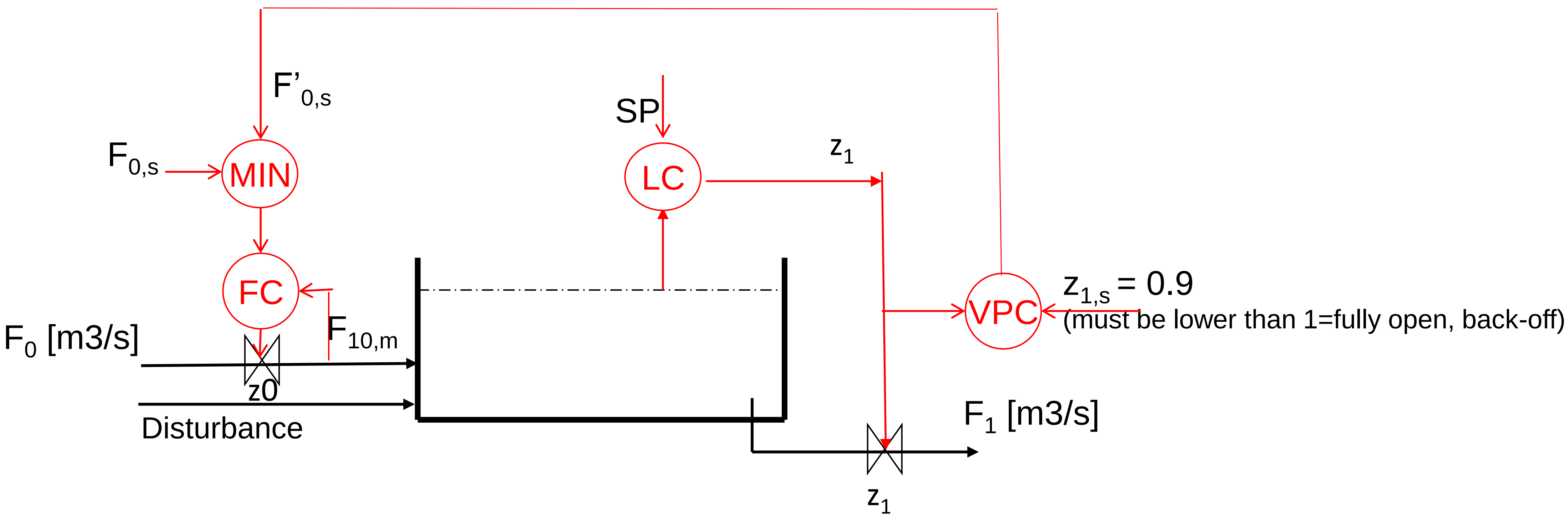
Three options for MV-MV switching

1. SRC (problem since F_{0s} varies)
2. Two controllers
3. VPC ("Long loop" for z1, backoff)

- Avoid long loop for control of F0
- Works both when F0-valve or F1-valve saturate at open

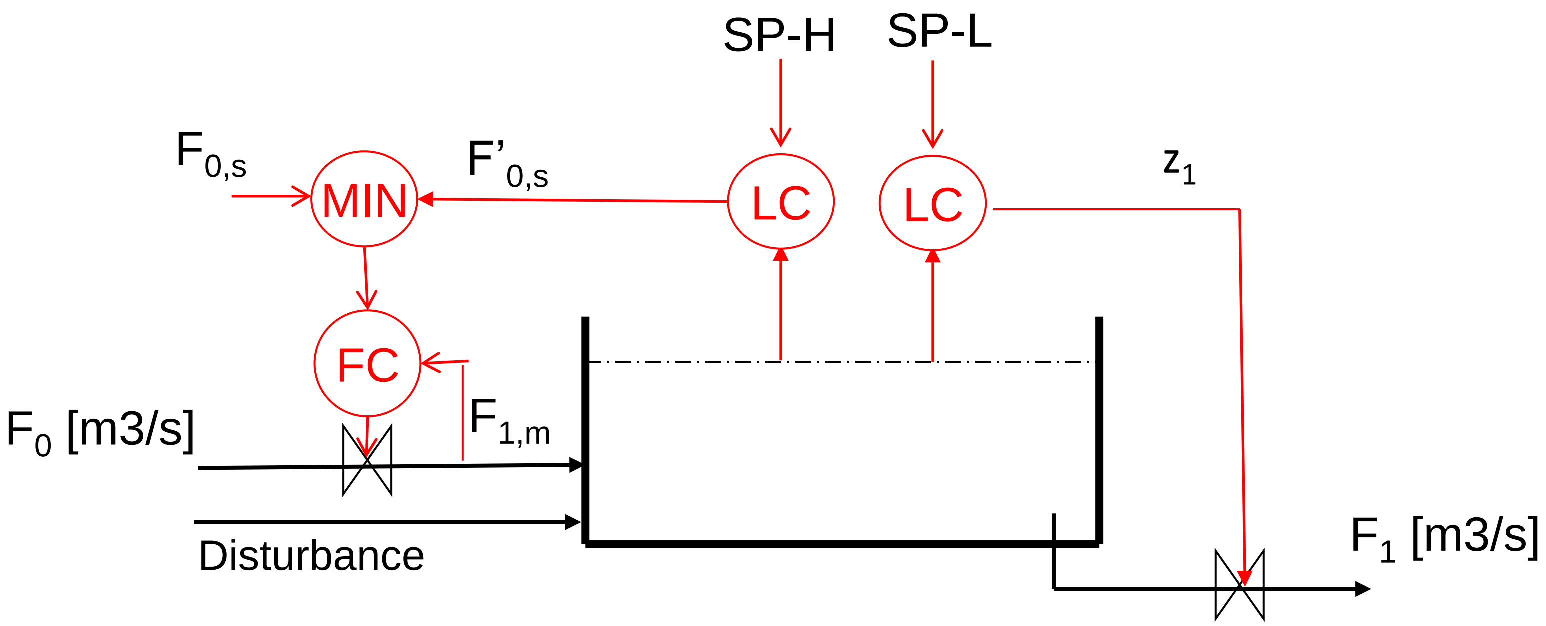
Overall: seems to be the best solution

Alt. 3. Valve position control on z_1



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

Alt. 2: Two controllers (recommended)

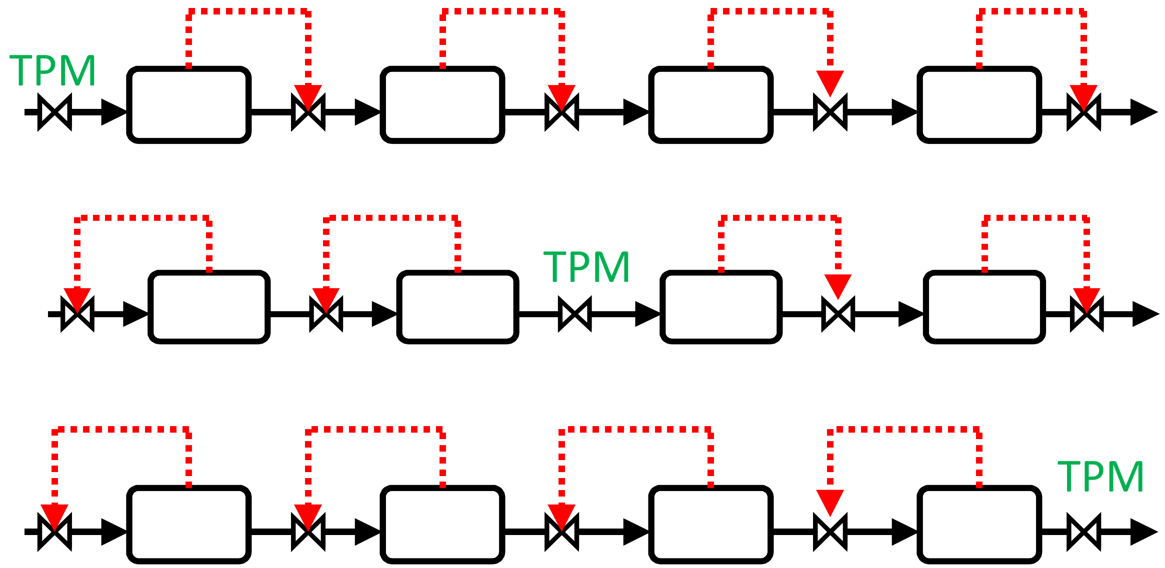


SP-L = low level setpoint
SP-H = high level setpoint

In addition: Use of two setpoints is good for using buffer dynamically!!

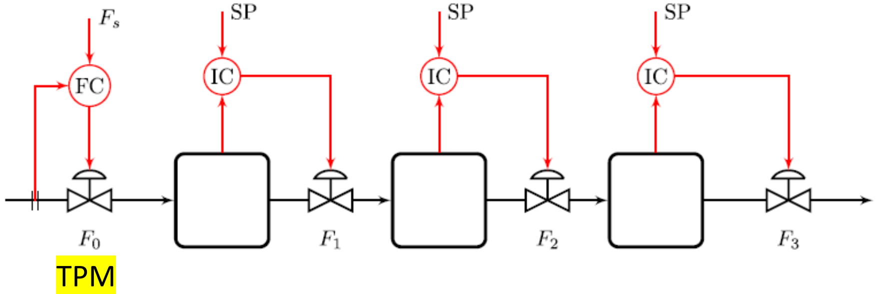
Inventory control for units in series and TPM

- *TPM* ("gas pedal") = *Variable used for setting the throughput/production rate (for the entire process).*
- Where is the TPM located for the process?
  - Usually at the feed, but not always!
  - Important for dynamics
  - Determines the inventory control structure
- **Rule (Price et al., 1994): Inventory control (Level and pressure) must be radiating around TPM:**

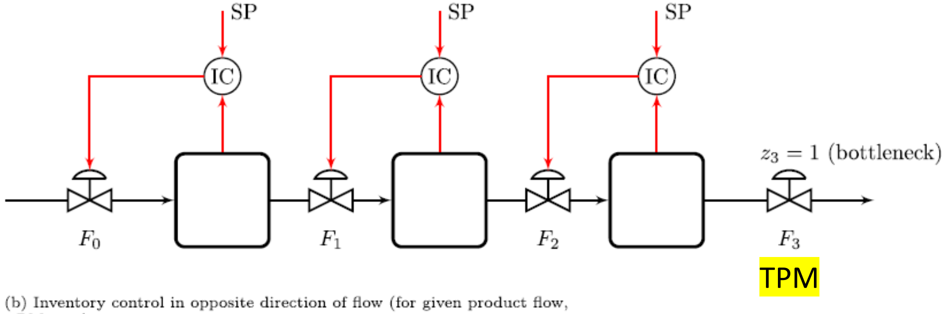


Inventory control for units in series

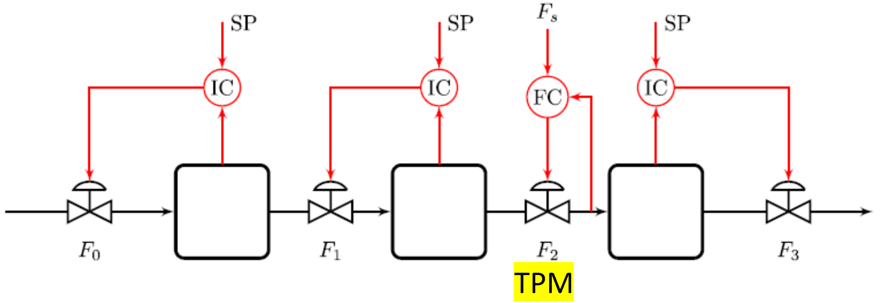
Radiating rule:
Inventory control should be "radiating" around a given flow (TPM).



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

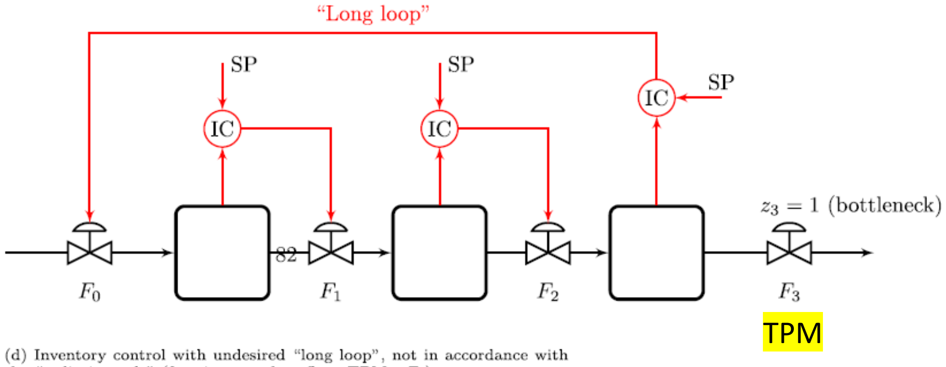


(b) Inventory control in opposite direction of flow (for given product flow, TPM= F_3)



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

Follows radiation rule



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

Does NOT follow radiation rule

Generalization of bidirectional inventory control

Reconfigures automatically with optimal buffer management!!

Maximize throughput:
$F_s=\infty$

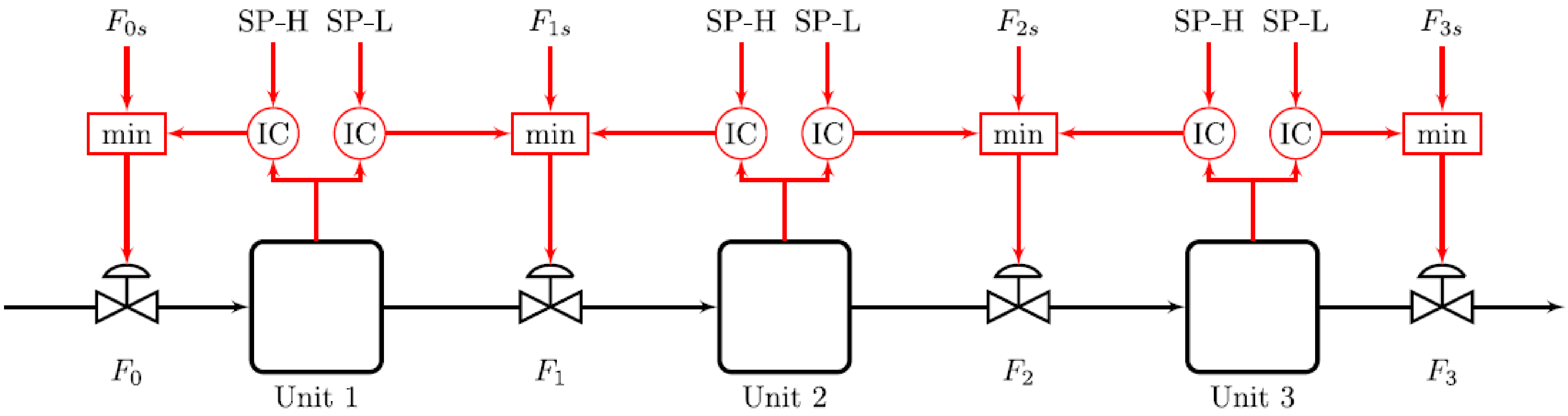
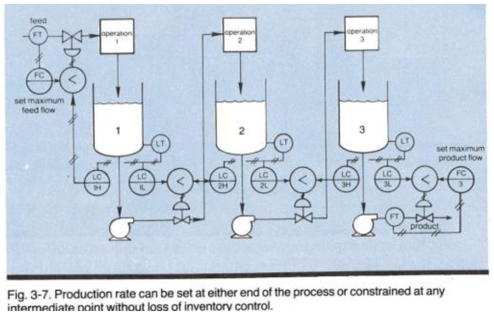
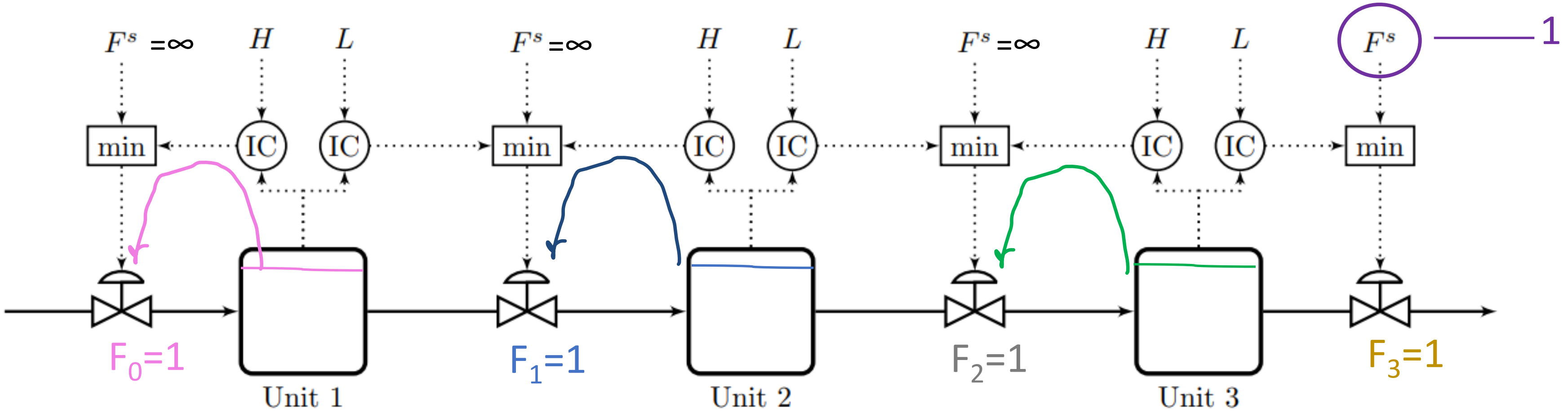


Fig. 36. Bidirectional inventory control scheme for automatic reconfiguration of loops (in accordance with the radiation rule) and maximizing throughput. Shinskey (1981) Zotică et al. (2022).
SP-H and SP-L are high and low inventory setpoints, with typical values 90% and 10%.
Strictly speaking, with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

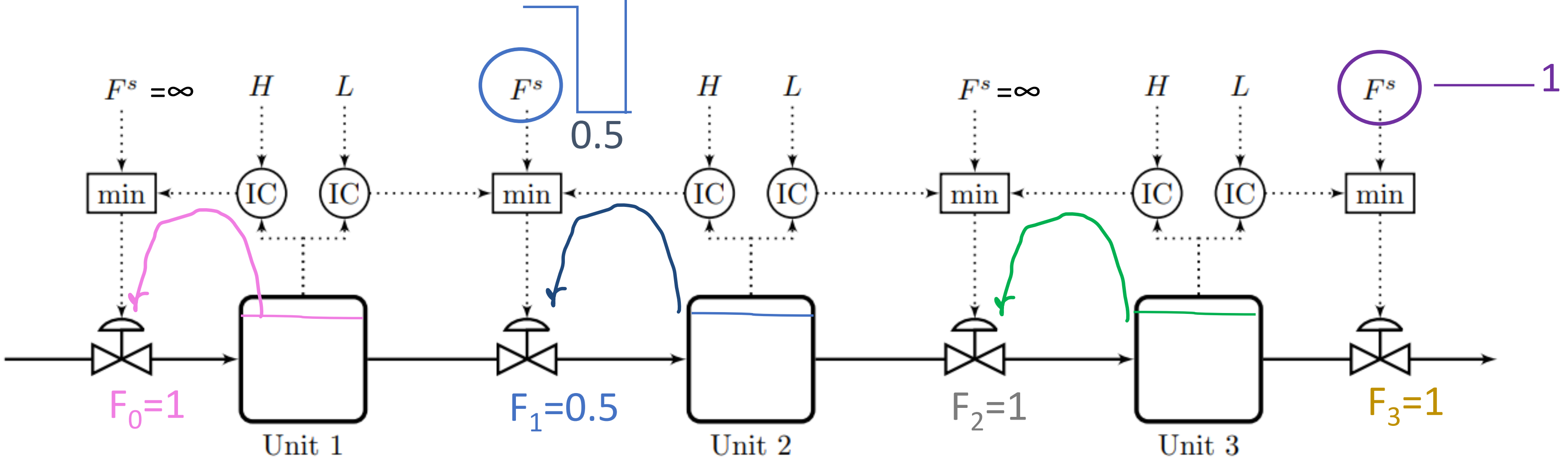
F.G. Shinskey, «Controlling multivariable processes», ISA, 1981, Ch.3



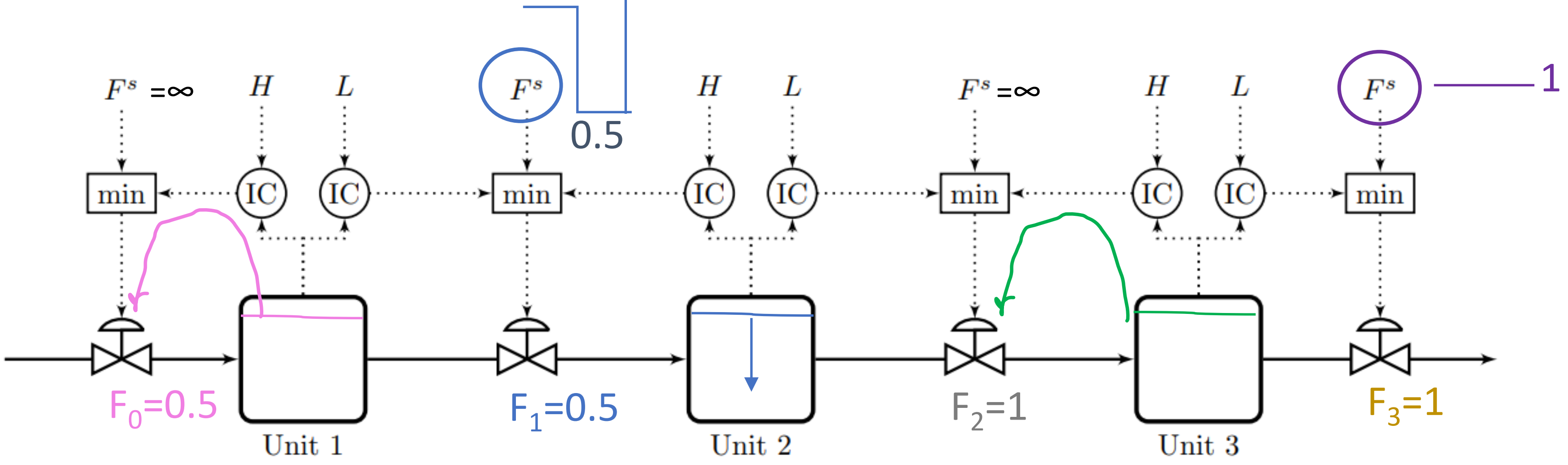
Cristina Zotica, Krister Forsman, Sigurd Skogestad ,»Bidirectional inventory control with optimal use of intermediate storage», Computers and chemical engineering, 2022



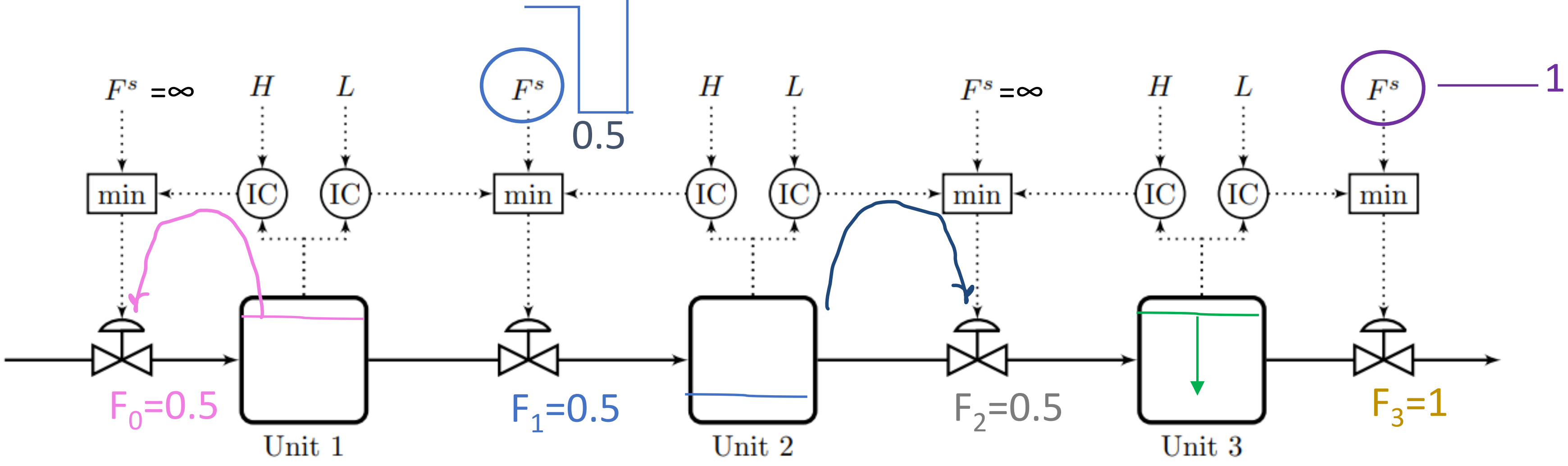
F^s =∞
H
L
F^s =∞
H
L
F^s =∞
H
L
F^s
1
min
IC
IC
min
IC
IC
min
IC
IC
min
F0=1
Unit 1
F1=1
Unit 2
F2=1
Unit 3
F3=1



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
F0=1
F1=0.5
F2=1
F3=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
F0=0.5
F1=0.5
F2=1
F3=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
F0=0.5
F1=0.5
F2=0.5
F3=1
Unit 1
Unit 2
Unit 3

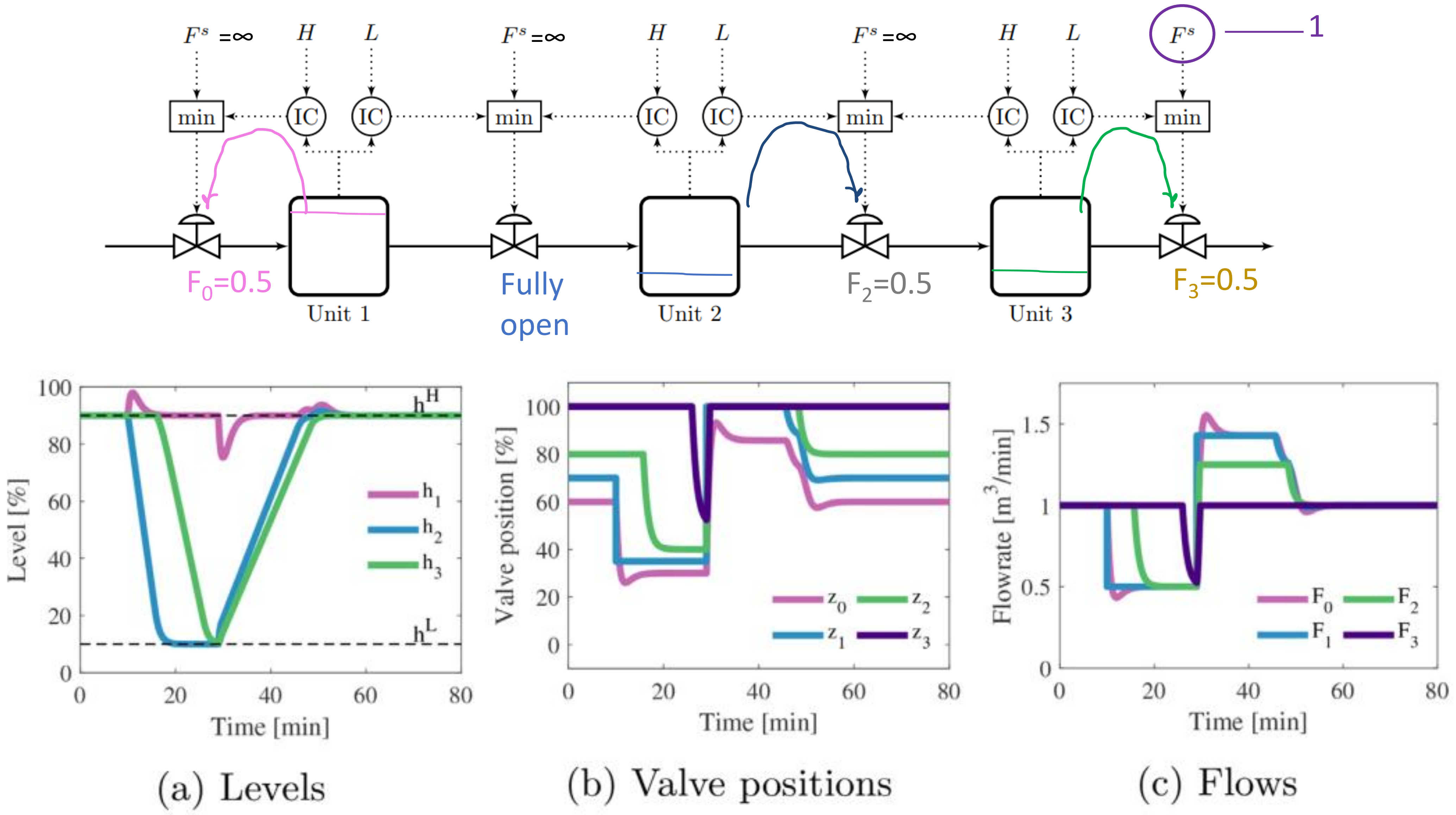
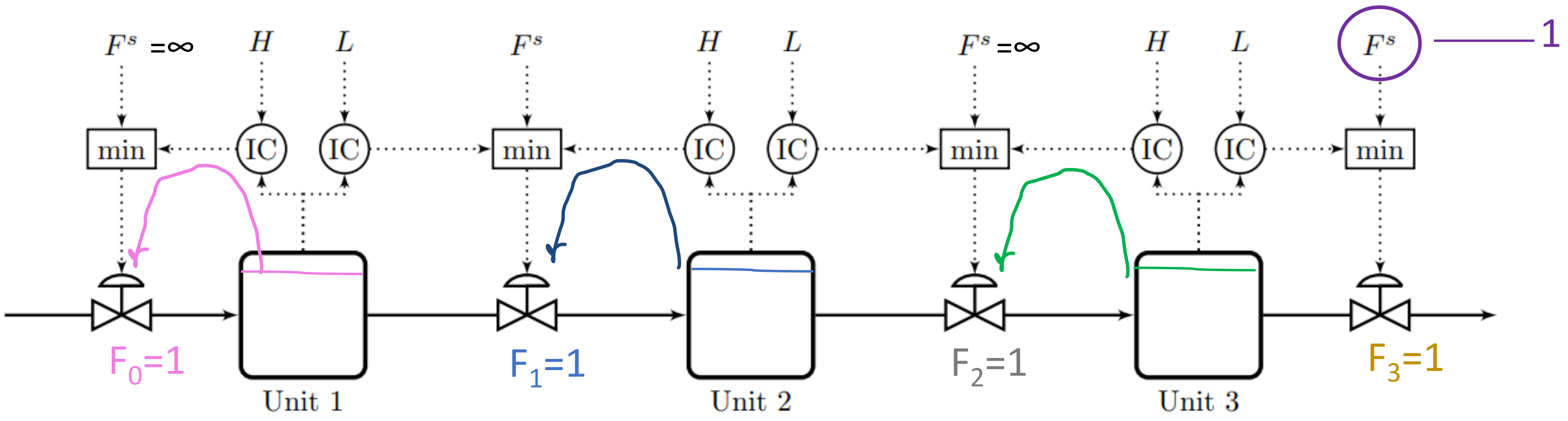


Fig. 13. Simulation of a temporary (19 min) bottleneck in flowrate F_1 for the proposed control structure in Fig. 10. The TPM is initially at the product (F_3).



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

YES. Use «trick»/insight of unachievable high setpoints on all flows

Extension . **Bidirectional inventory control with minimum flow for F_2**

Max flow: $F_s=\infty$

L = 10%,

M_L = 40%,

M_H = 60%

H = 90%.

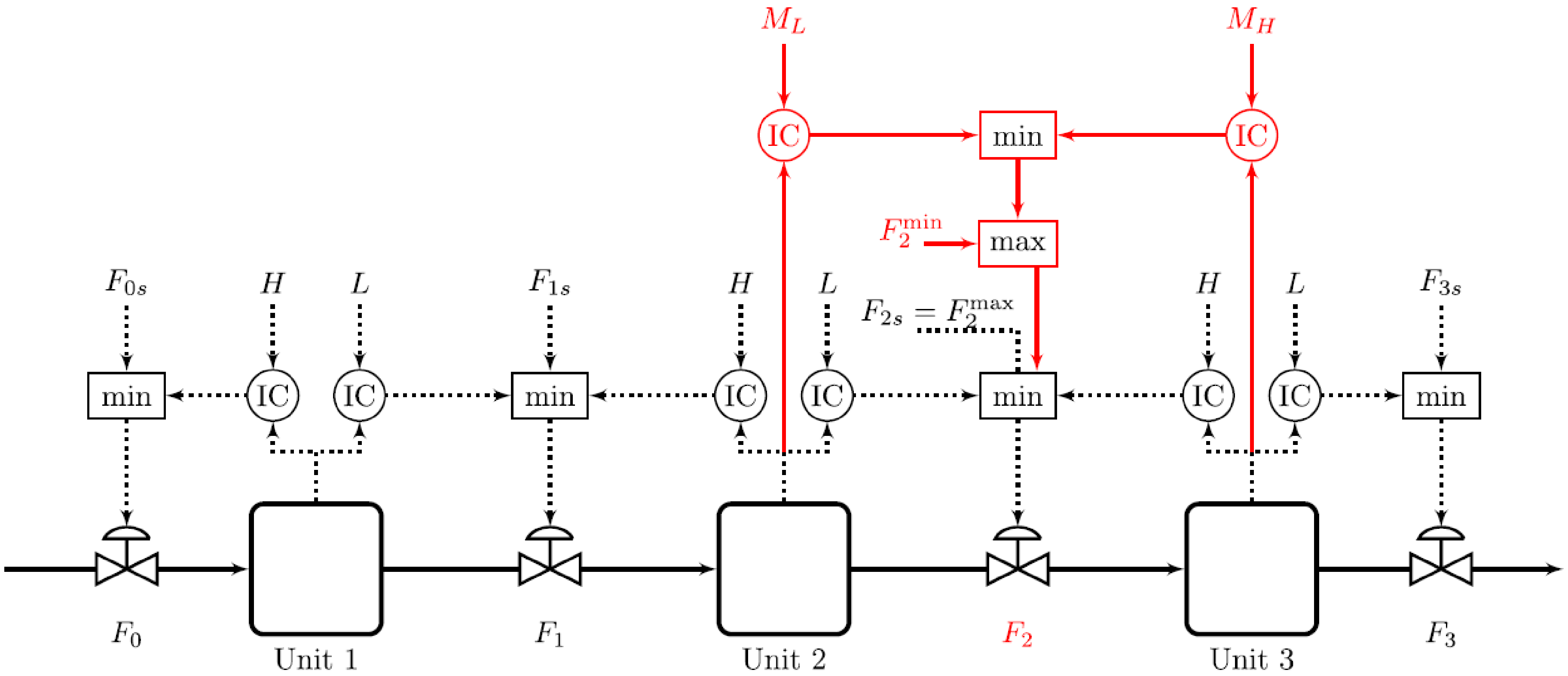


Fig. 37. Bidirectional inventory control scheme for maximizing throughput (dashed black lines) while attempting to satisfy minimum flow constraint on F_2 (red lines). H, L, M_L and M_H are inventory setpoints.

The control structure in Fig. 37 may easily be dismissed as being too complicated so MPC should be used instead. At first this seems reasonable, but a closer analysis shows that MPC may not be able to solve the problem (Bernardino & Skogestad, 2023).[8] Besides, is the control structure in Fig. 37 really that complicated? Of course, it is a matter of how much time one is willing to put into understanding and studying such structures. Traditionally, people in academia have dismissed almost any industrial structure with selectors to be ad hoc and difficult to understand, but this view should be challenged.

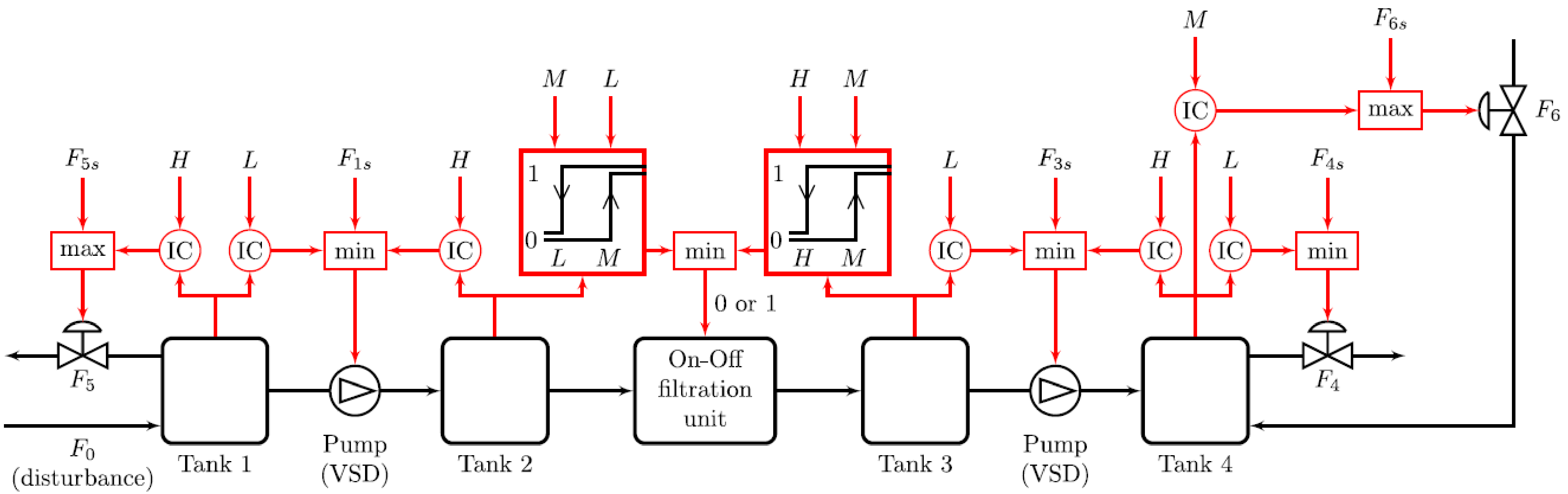


Fig. 38. Bidirectional inventory control structure for industrial plant with on/off (1/0) control of filtration unit. H, L and M are inventory setpoints with typical values 90%, 10% and 50%.
If it is desirable to set a flowrate (F_s) somewhere in the system, then flow controllers must be added at this location.

Inventory control (level, pressure)

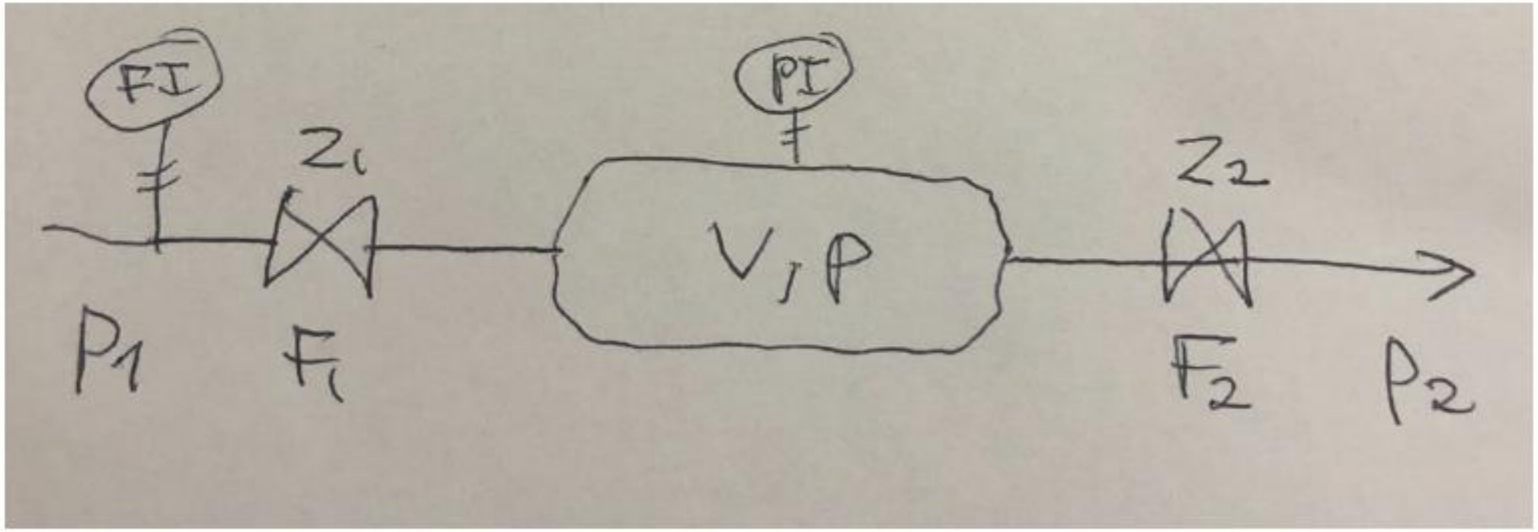
- All inventories (level, pressure) must be regulated by
  - Controller, or
  - “self-regulated” (e.g., overflow for level, open valve for pressure)
  - Exception closed system: Must leave one inventory (level) uncontrolled

- Usually only one TPM
  - To get consistent mass balance: Can only fix same flow once
  - But there are exceptions
    - Multiple feeds (they are then usually set in ratio to the “main” TPM)
    - Recycle systems often have a flow that can be set freely

- Rule for maximizing production for cases where we cannot rearrange inventory loops: Locate TPM at expected bottleneck
  - Otherwise you will need a “long loop” and you get loss in production because of backoff from constraint

QUIZ
Exam 2022

Problem 5 (25%). Modelling and control of flow and pressure



Consider a gas pipeline with two valves. We have measurements of the inflow F_1 and the intermediate pressure p and these should be controlled. The volume of the pipeline can be represented as a tank with volume V as shown in the figure above.

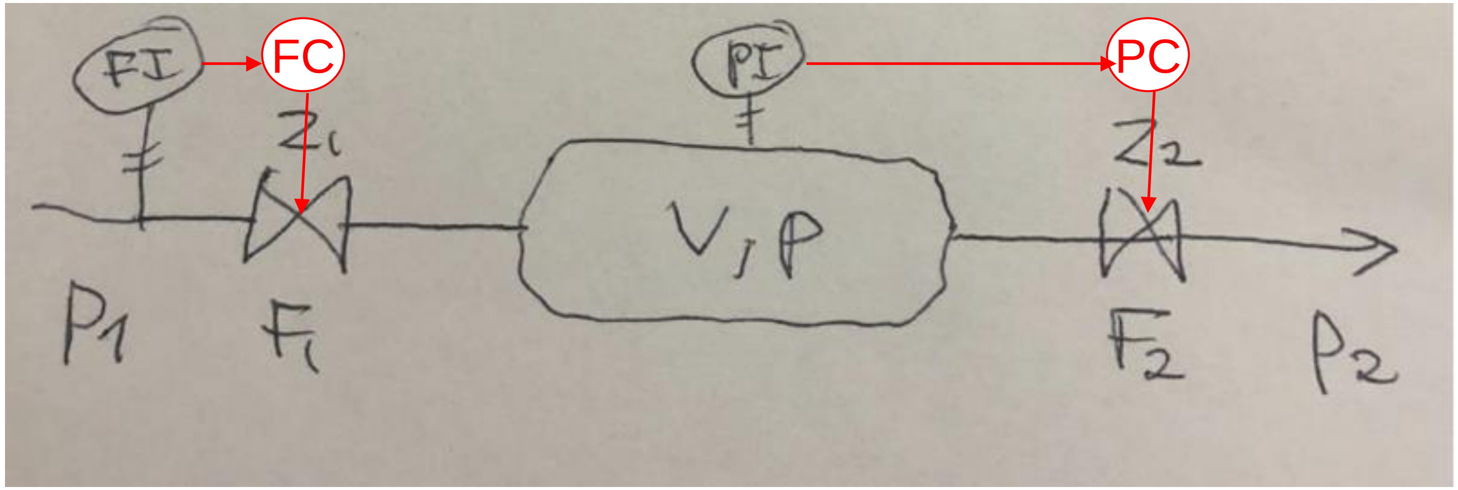
Steady-state data: F_1=1 kg/s, $z_1=z_2=0.5$, p_1=2 bar, p=1.88 bar, p_2=1.8 bar, V=130 m^3, T=300 K,
Parameters: R=8.31 J/K.mol, M_W=18e-3 kg/mol (so the gas is steam).

The following model equations are suggested to describe the system.

(1) $dm/dt = F_1 - F_2$
(2) $m = k_p\, p$ where $k_p = VM_w/(RT)$
(3) $F_1 = C_1 z_1 \sqrt{p_1 - p}$
(4) $F_2 = C_2 z_2 \sqrt{p - p_2}$

(a) Suggest a control structure
(b) What if we want to control p2 instead of p?

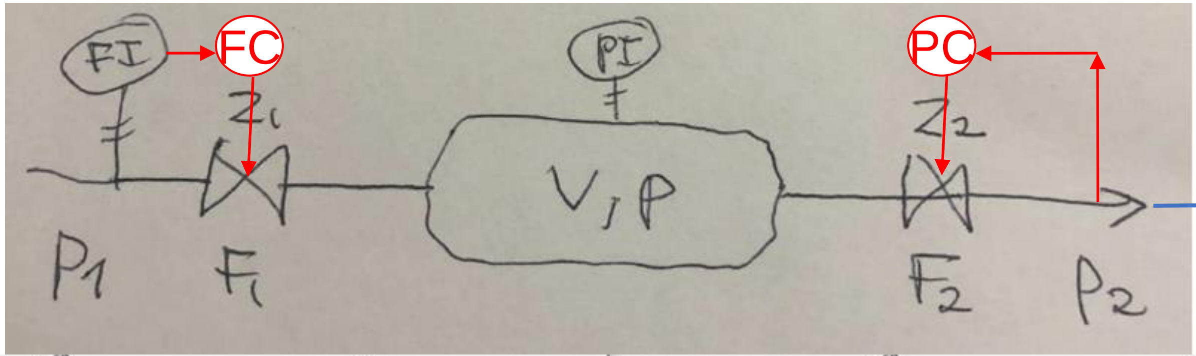
(a) The «obvious» pair-close pairing os OK. However, interactions between loops may be severe. Suggest tuning the FC first, and the PC about 5 times slower.



(b)

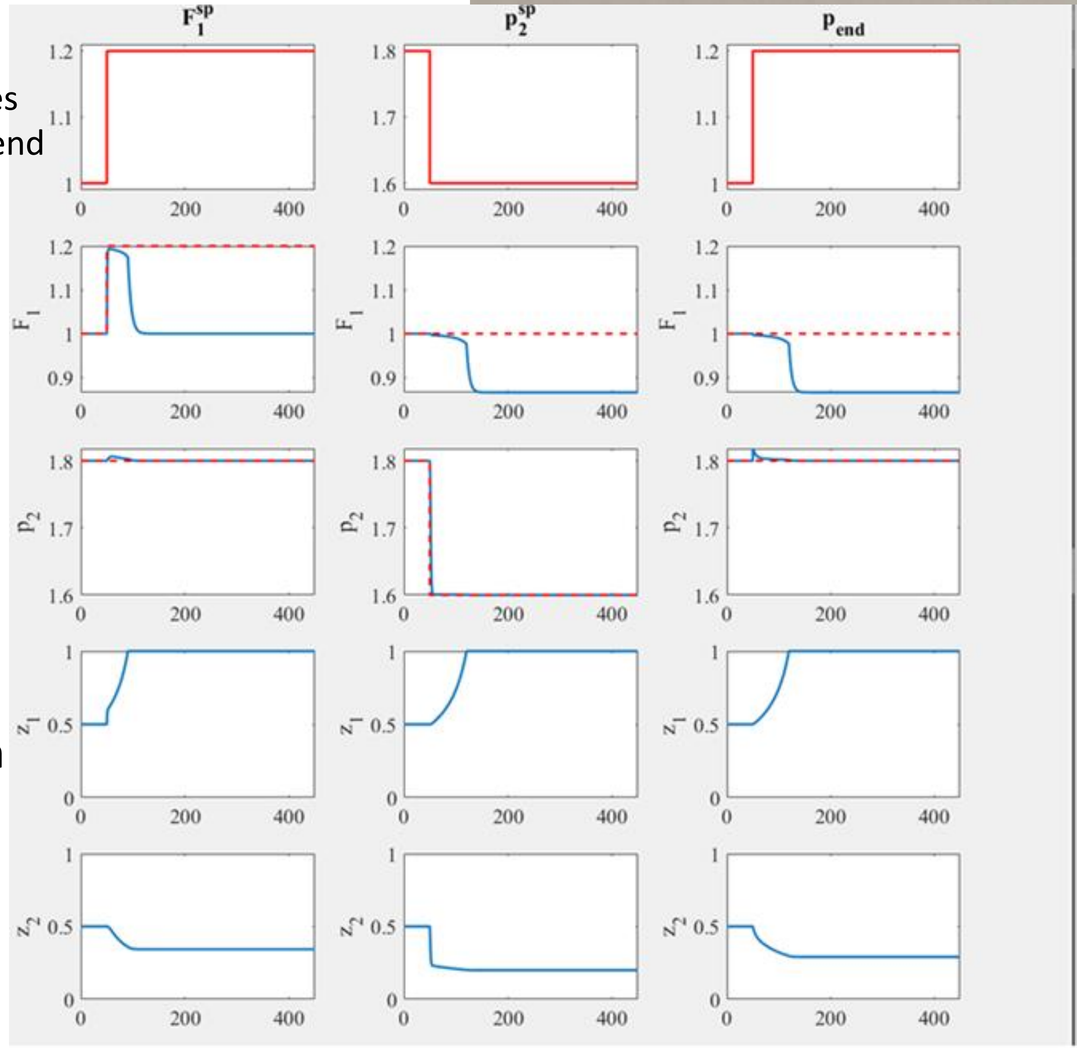
NO!

Not consistent



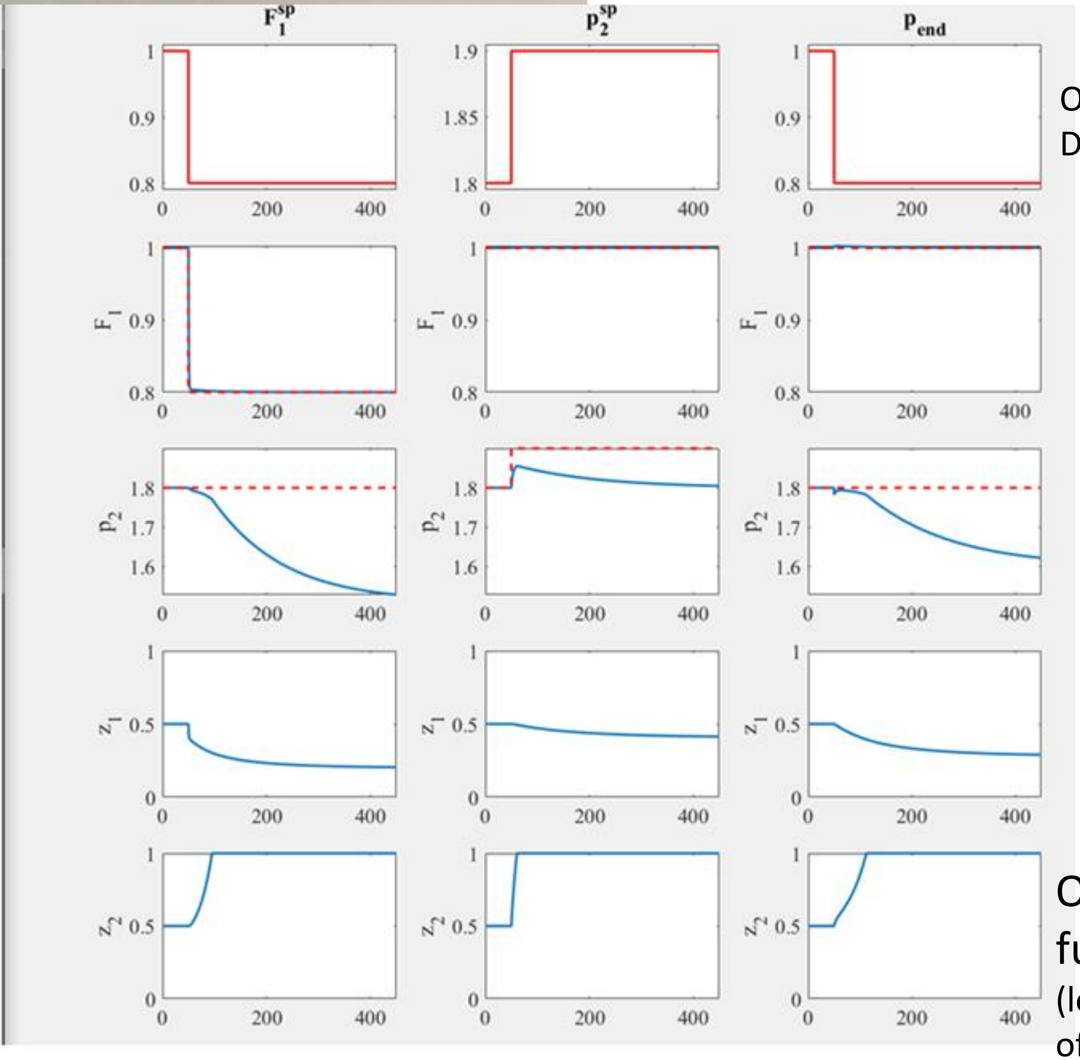
$F_3 = c_3(p_2 - p_{end})^{1/2}$

Disturbances in F1, p2, pend



z1 to fully open (lose control of F1)

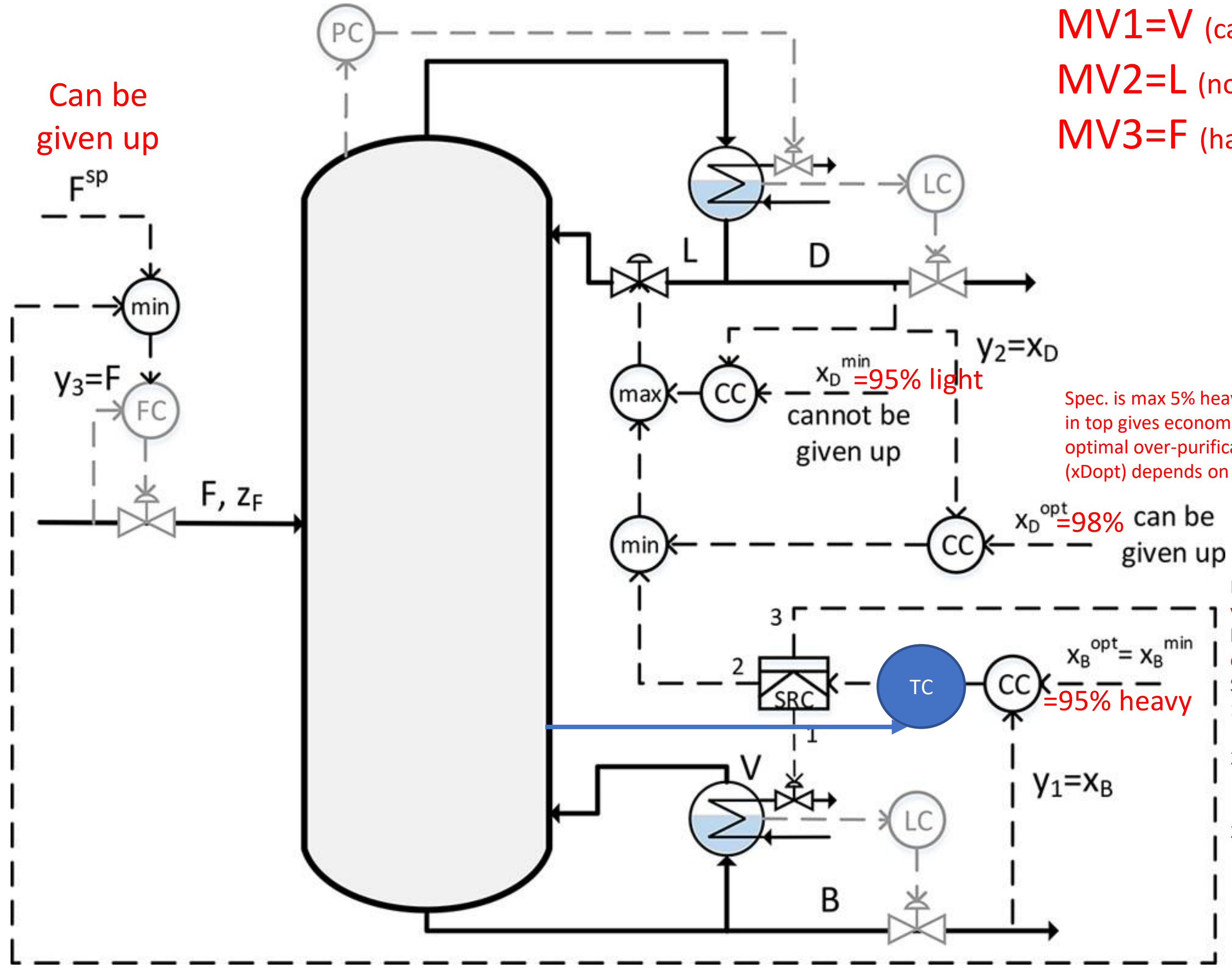
Opposite Disturbances



Or: z2 to fully open (lose control of p2)

Distillation example.

CV1=xB (cannot be given up)
MV1=V (can reach max)
MV2=L (normally used for xD)
MV3=F (has setpoint)



Spec. is max 5% heavy in top, but heavy in top gives economic loss. Economic optimal over-purification of top product (xDopt) depends on prices p

Bottom product is the most valuable and should always be kept at its purity constrant (xBmin).
Split range control*:

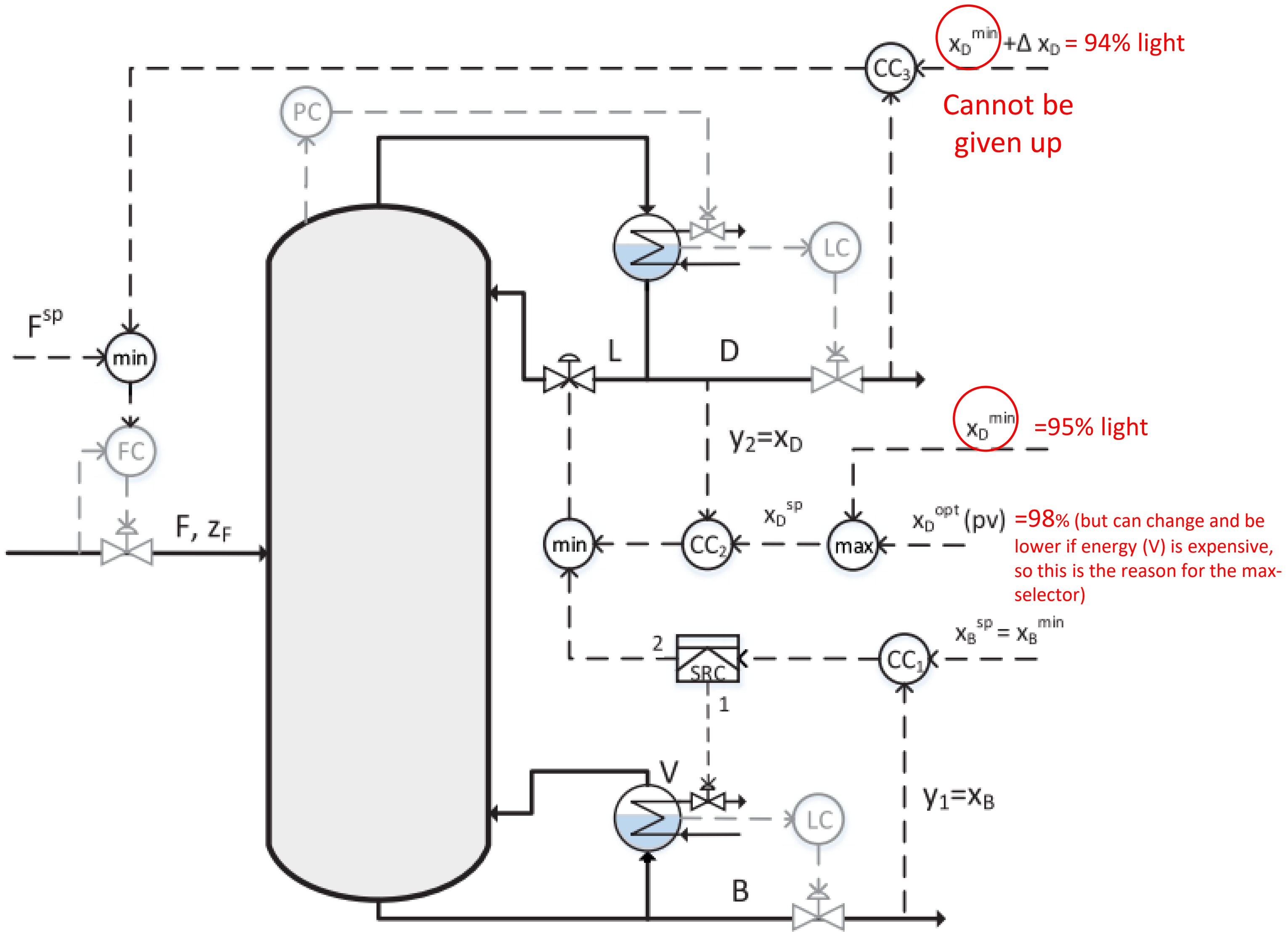
1. Normally we control xB with V
2. If V reaches Vmax then we instead use reflux L (and give up xDopt)
3. If L is not availabe (beacuse it's used to control sDmin) then we instead use the feedrate F to control xB

*Split range control can be replaced by three controllers with different setpoints for xB

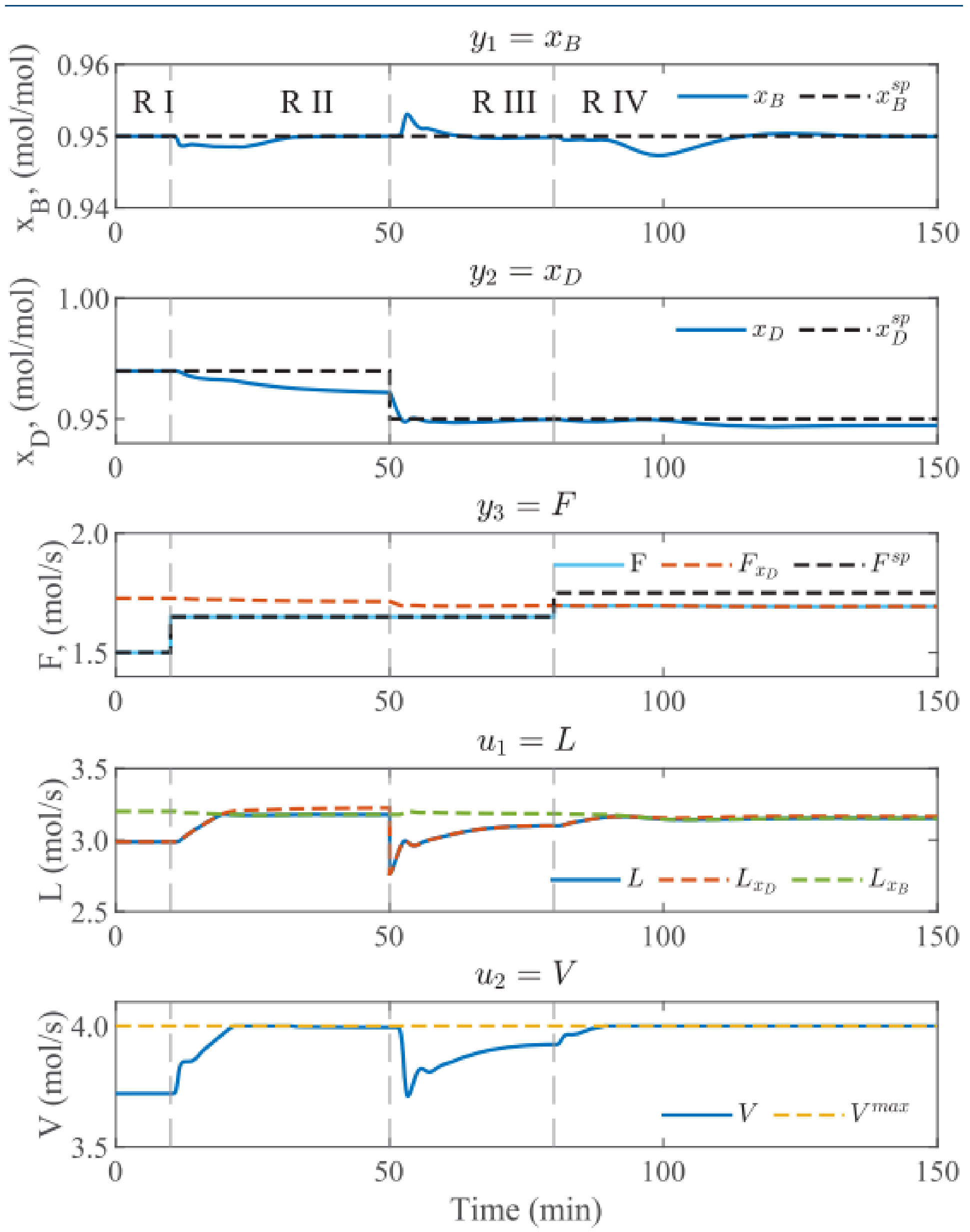
"Systematic Design of Active Constraint Switching Using Classical Advanced Control Structures"
Adriana Reyes-Lúa and Sigurd Skogestad
Industrial & Engineering Chemistry Research **2020** *59* (19), 9342-9342

This is an example where MPC may be preferred

Alternative solution with different setpoints



This solution looks simpler, but it is not as good dynamically in cases where we need to limit feed F to the column.
We then use F to control top composition, and L to control bottom compoistion.
The reverse pairing is better (which is what we get with the other solution)



Simulation of alternative solution. The problematic pairing is used toward the end (t>80), but it's not really tested because there are no disturbances

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

Here is a summary of some additional insights from this paper:

- If the industrial solution has a selector (sometimes realized using a saturation element, especially for the cascade implementation) then generally there is a CV constraint involved. Most likely, the selector is performing a steady-state CV-CV switch (E4), although there may be exceptions as seen in the cross-limiting example below.
  - A CV-CV switch can be realized in two ways, either with two (or more) independent controllers with a selector on the MV (Fig. 17), or as a cascade implementation with a selector on the CV setpoint (Fig. 19).
  - If there are several selectors (max and min) in series then we know that the constraints are potentially conflicting and that the highest priority constraint should be at the end (Fig. 18).

- If the industrial solution has a valve position controller (VPC) then there may be two quite different problems that it is addressing (see E3 and E7 in Table 1), and it may not be immediately clear which.

  1. If we have an extra MV for dynamic reasons (E3; Fig. 12) then the two controllers (and MVs) are used all the time. The MV manipulated by the VPC (MV_1 in Fig. 12) is then used on the longer time scale, whereas the MV linked to the CV (MV_2 in Fig. 12) is used for dynamic reasons (fast control). Here, an alternative is to use parallel control (Fig. 13).
  2. There is also another possibility, namely, when the VPC makes use of an extra MV to avoid that the primary MV saturates at steady-state (E7; Fig. 24). This is then a case where the VPC is used for MV-MV switching and the VPC is only active part of the time.

- For MV-MV switching there are three alternatives.

  1. A common solution is split range control (E5; Fig. 21) which is usually easy to identify.
  2. Another common solution is multiple controllers with different setpoints (E6; Fig. 23). It may be a bit more difficult to identify.
  3. Finally, there is VPC (E7), as just discussed, which is probably the least common solution for MV-MV switching

  One should have all these three alternatives in mind when choosing the best solution for MV-MV switching, as there is not one alternative which is best for all problems (see Section 5.1 for details).

Challenges selector design

- Standard approach requires pairing of each active constraint with a single input
  - May not be possible in complex cases
- Stability analysis of switched systems is still an open problem
  - Undesired switching may be avoided in many ways:
    - Filtering of measurement
    - Tuning of anti-windup scheme
    - Minimum time between switching
    - Minimum input change

Implementing optimal operation Summary

- Most people think
  - You need a detailed nonlinear model and an on-line optimizer (RTO) if you want to optimize the process
  - You need a dynamic model and model predictive control (MPC) if you want to handle constraints
  - The alternative is Machine Learning
- No! In many cases you just need to measure the constraints and use PID control
  - «Conventional advanced regulatory control (ARC)»
- How can this be possible?
  - Because optimal operation is usually at constraints
  - Feedback with PID-controllers can be used to identify and control the active constraints
  - For unconstrained degrees of freedom, one often have «self-optimizing» variables

- This fact is not well known, even to control professors
  - Because most ARC-applications are *ad hoc*
  - Few systematic design methods exists

- Today ARC and MPC are in parallel universes
  - Both are needed in the control engineer's toolbox