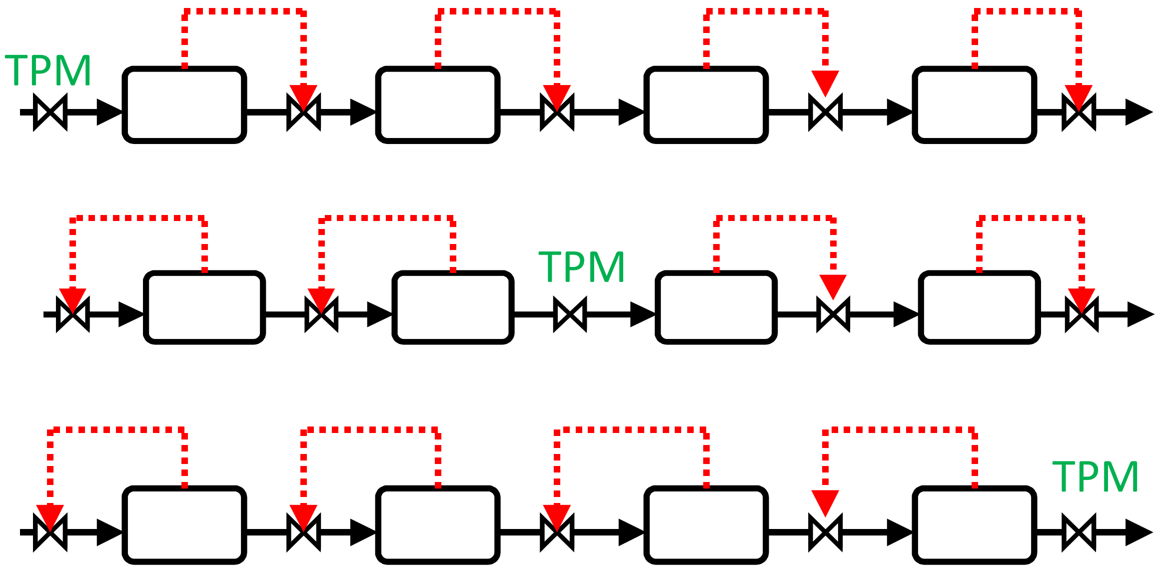
Part 4: Inventory control and optimal buffer management

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

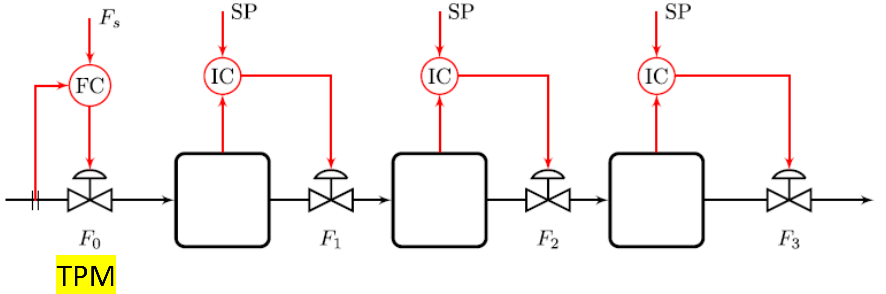
Inventory control for units in series and TPM

- *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:**

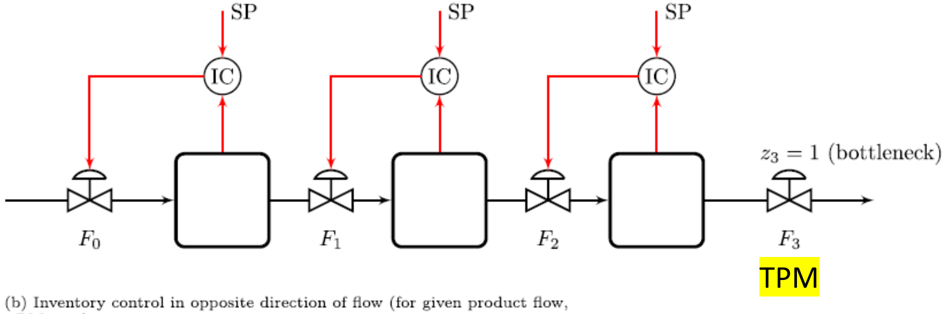


Inventory control for units in series

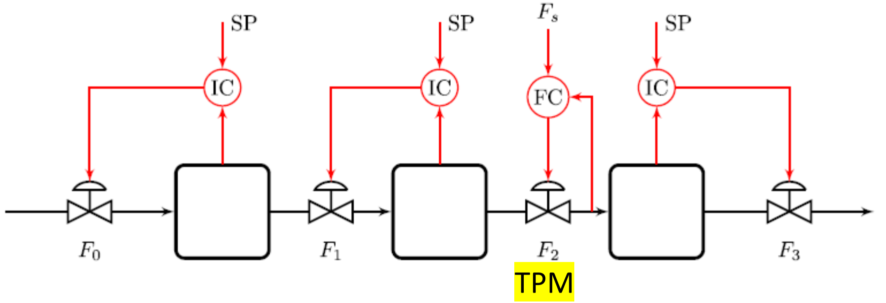
Radiating rule:
Inventory control should be "radiating" around a given flow (TPM).



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

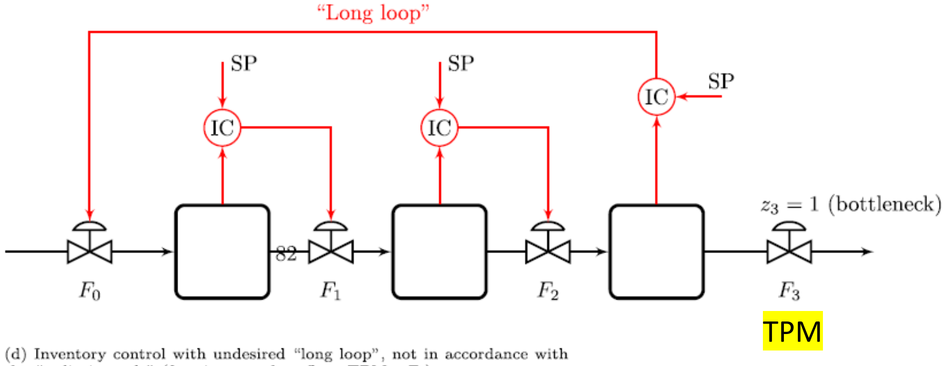


(b) Inventory control in opposite direction of flow (for given product flow, TPM= F_3)



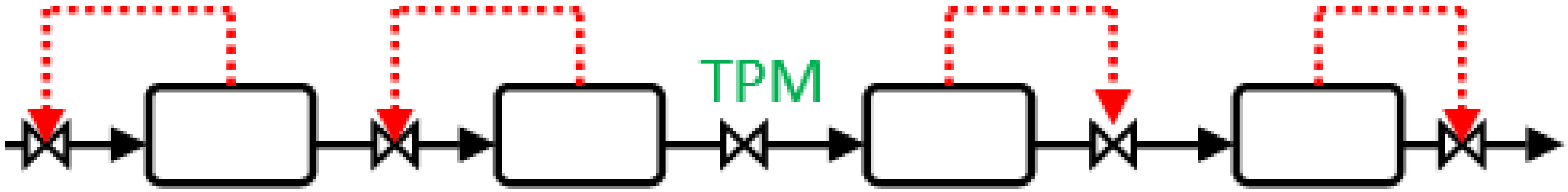
(c) Radiating inventory control for TPM in the middle of the process (shown for TPM = F_2)

Follows radiation rule



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

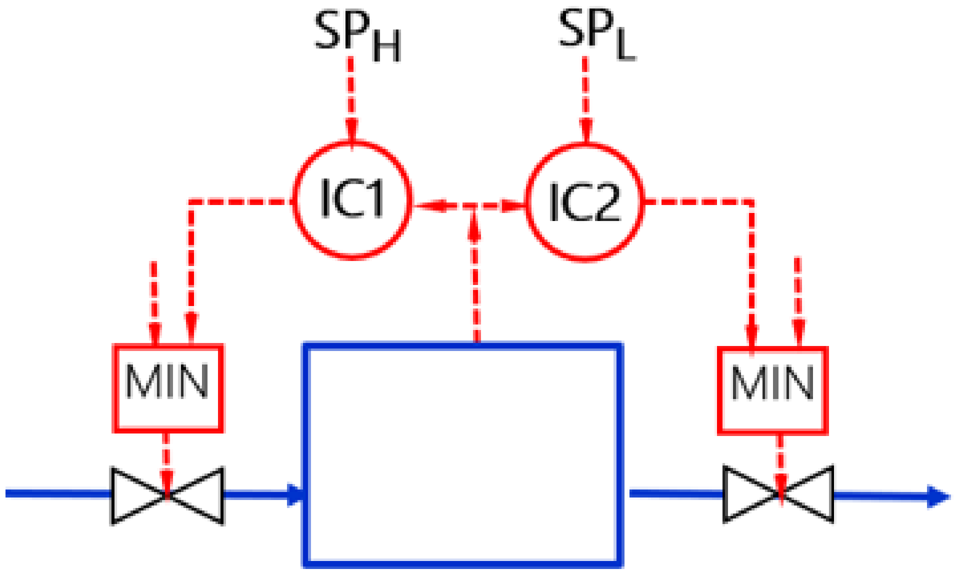
Does NOT follow radiation rule

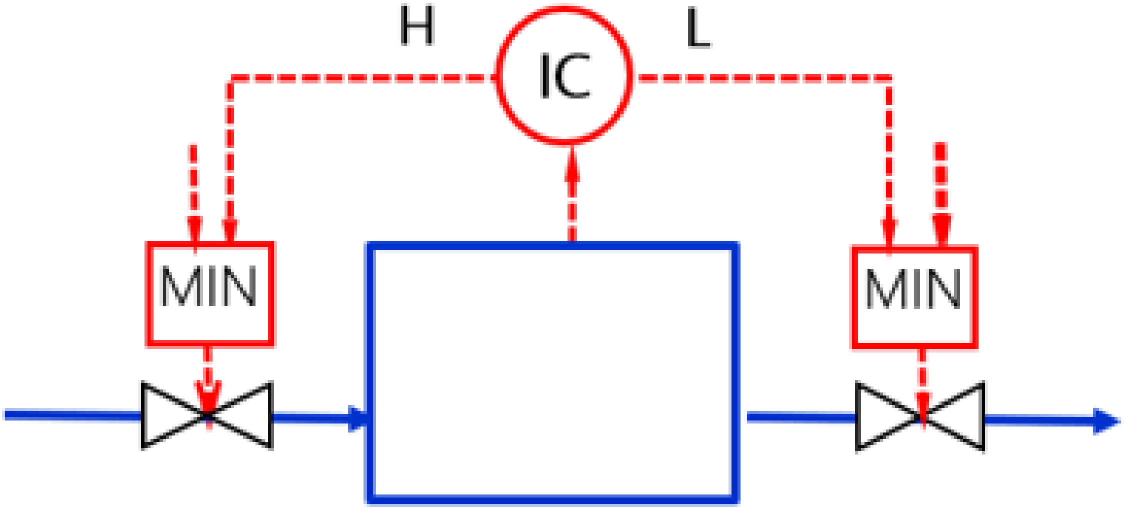


- TPM = throughput manipulator – where throughput of plant is set
- Usually only one TPM
    - To get consistent mass balance: Can only fix same flow once
    - But there are exceptions
        - Multiple feeds (they are then usually set in ratio to the “main” TPM)
        - Recycle systems often have 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

Split-parallel bidirectional inventory control





Simplified notation

Bidirectional inventory control for units in series

Reconfigures automatically with optimal buffer management!!

Maximize throughput:
$F_s = \infty$

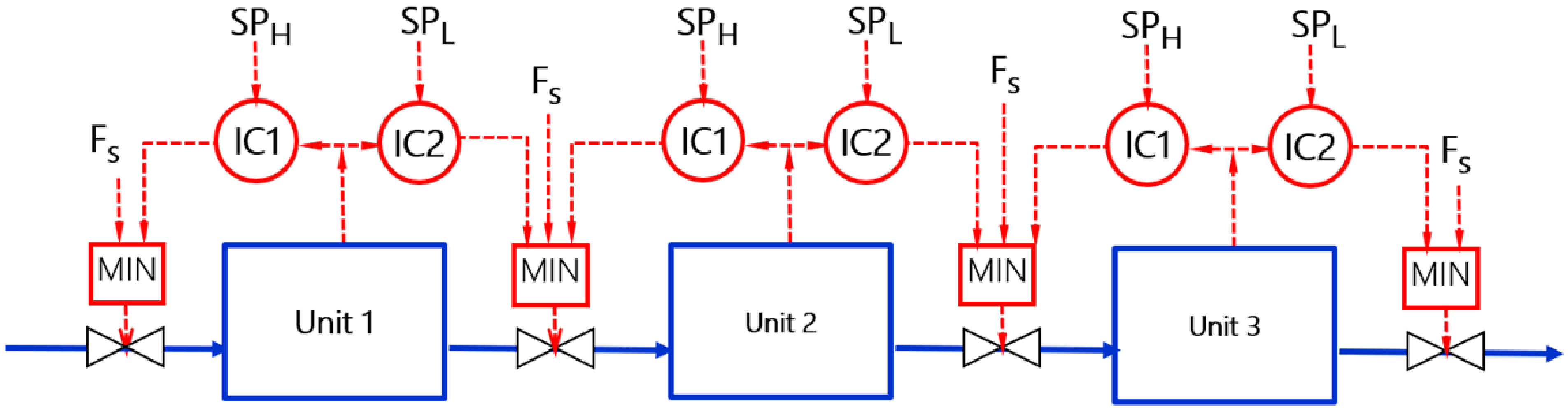


Fig. 36. Bidirectional inventory control scheme for automatic reconfiguration of loops (in accordance with the radiation rule) and maximizing throughput. Shinskey (1981) Zotică et al. (2022).
SP-H and SP-L are high and low inventory setpoints, with typical values 90% and 10%.
Strictly speaking, with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

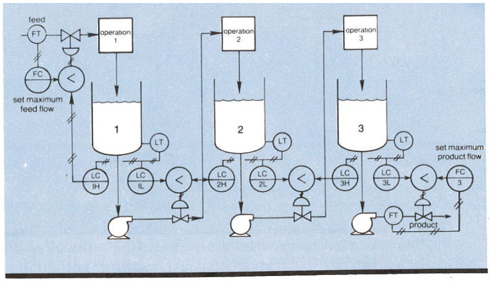
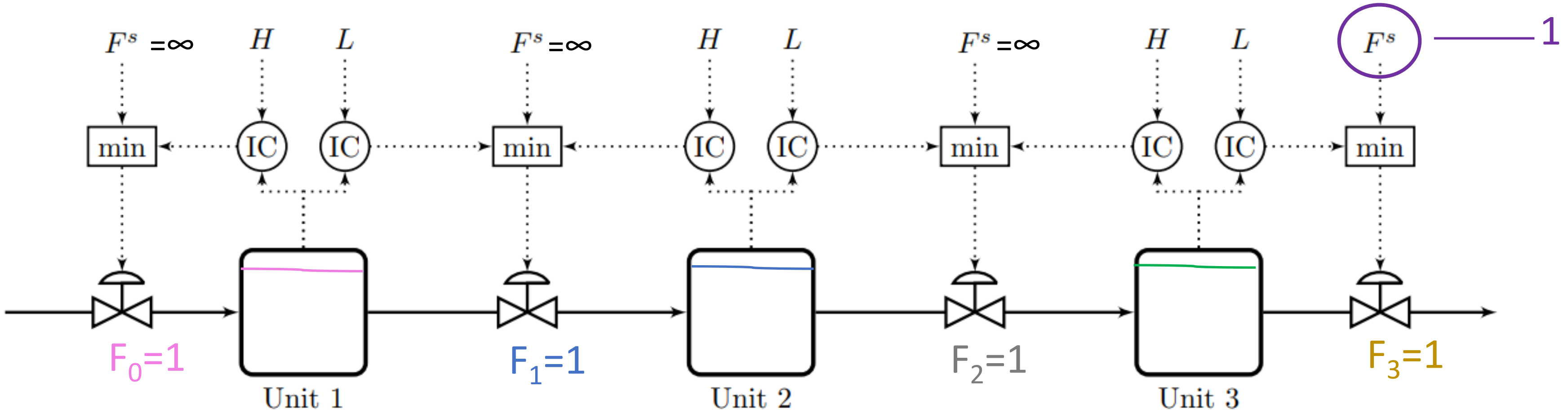
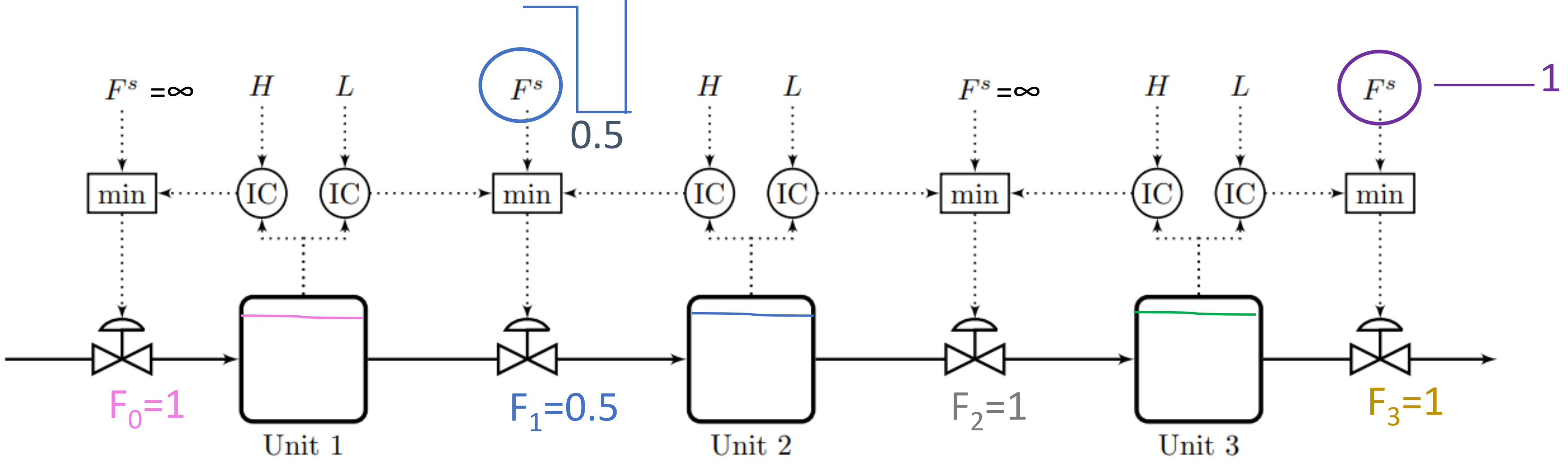


Fig. 3-7. Production rate can be set at either end of the process or constrained at any intermediate point without loss of inventory control.

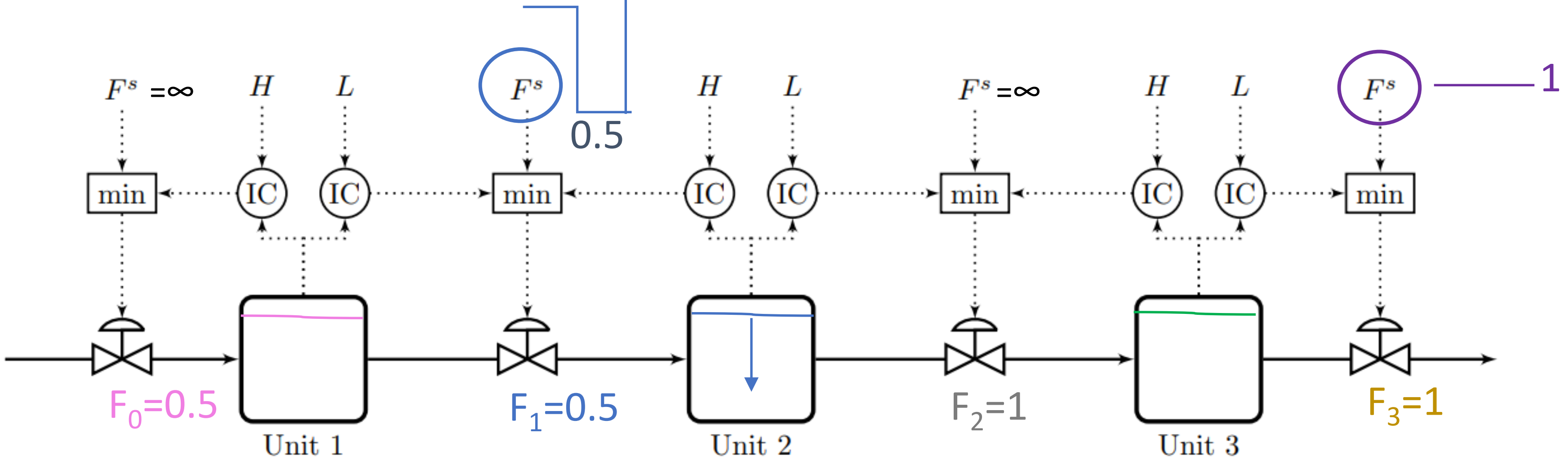
Cristina Zotica, Krister Forsman, Sigurd Skogestad ,»Bidirectional inventory control with optimal use of intermediate storage», Computers and chemical engineering, 2022



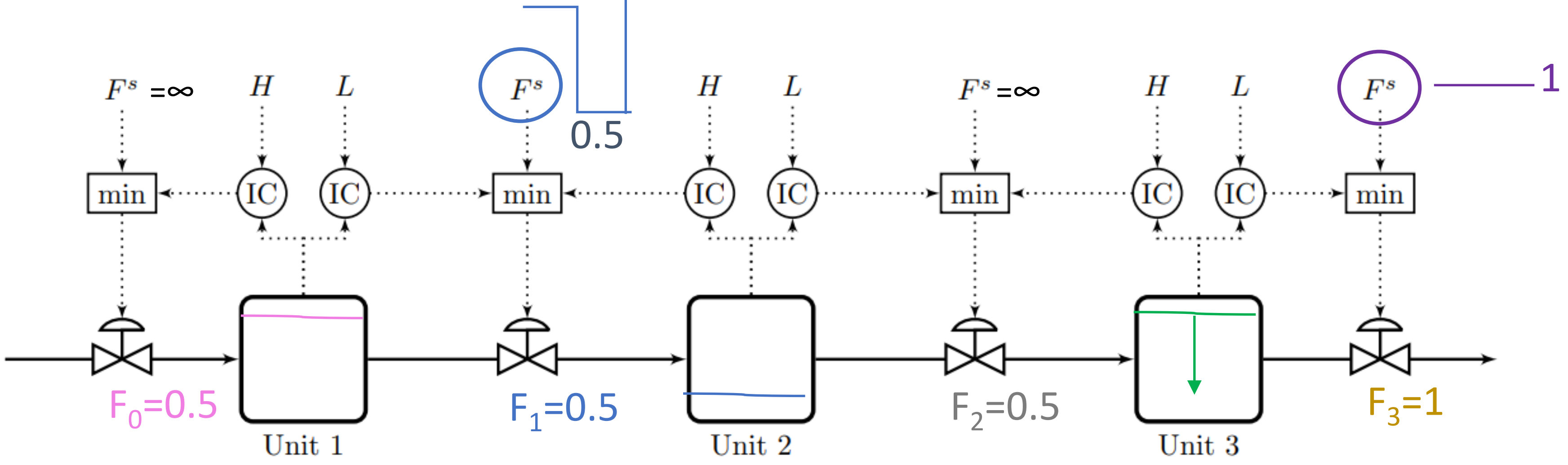
F^s =∞
H
L
F^s =∞
H
L
F^s =∞
H
L
F^s
1
min
IC
IC
min
IC
IC
min
IC
IC
min
F0=1
Unit 1
F1=1
Unit 2
F2=1
Unit 3
F3=1



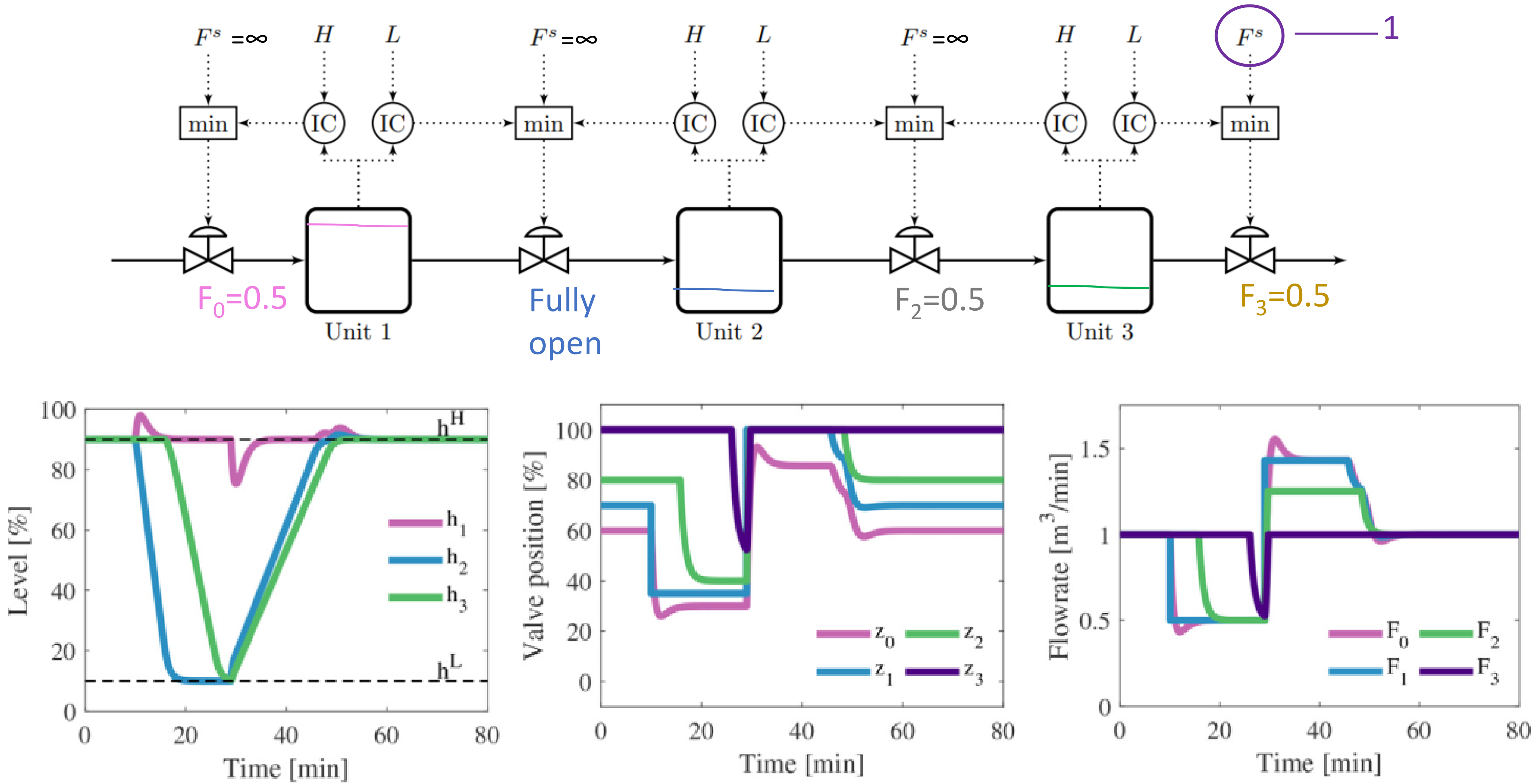
F^s =∞
H
L
F^s
0.5
H
L
F^s =∞
H
L
F^s
1
min
IC
IC
min
IC
IC
min
IC
IC
min
F0=1
Unit 1
F1=0.5
Unit 2
F2=1
Unit 3
F3=1



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

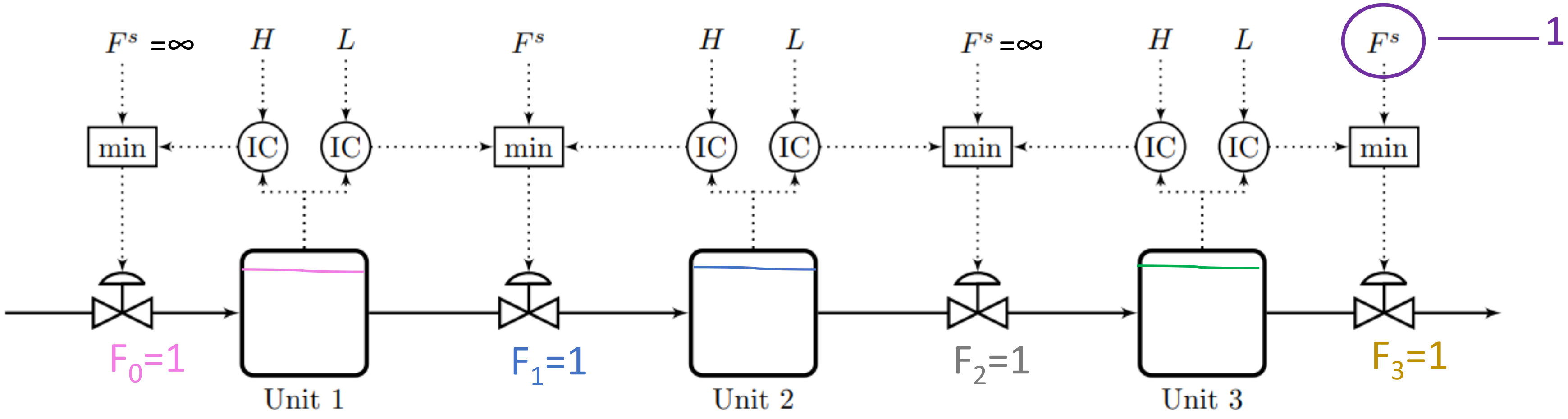


$F^s = \infty$
H
L
F^s
0.5
H
L
$F^s = \infty$
H
L
F^s
1
min
IC
IC
min
IC
IC
min
IC
IC
min
$F_0=0.5$
$F_1=0.5$
$F_2=0.5$
$F_3=1$
Unit 1
Unit 2
Unit 3



(a) Levels (b) Valve positions (c) Flows

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

Industrial Case (Perstorp)

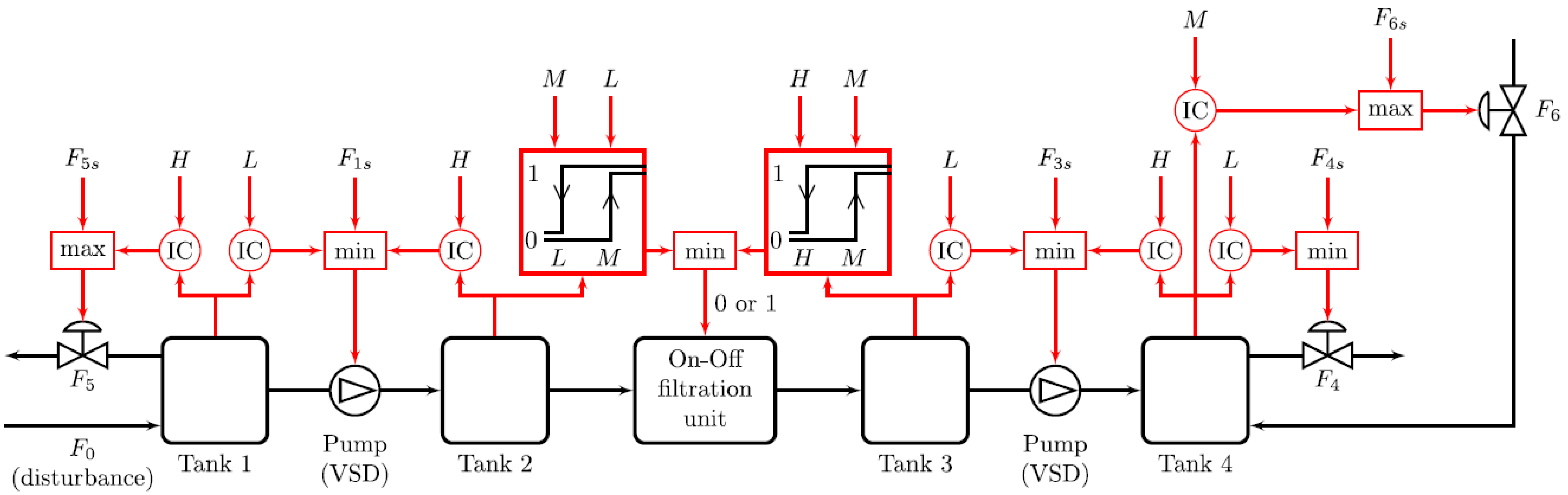


Fig. 38. Bidirectional inventory control structure for industrial plant with on/off (1/0) control of filtration unit. H, L and M are inventory setpoints with typical values 90%, 10% and 50%.

If it is desirable to set a flowrate (F_s) somewhere in the system, then flow controllers must be added at this location.

I made this example to find a case where MPC does not work;

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

Max flow: $F_s = \infty$

L = 10%,

M_L = 40%,

M_H = 60%

H = 90%.

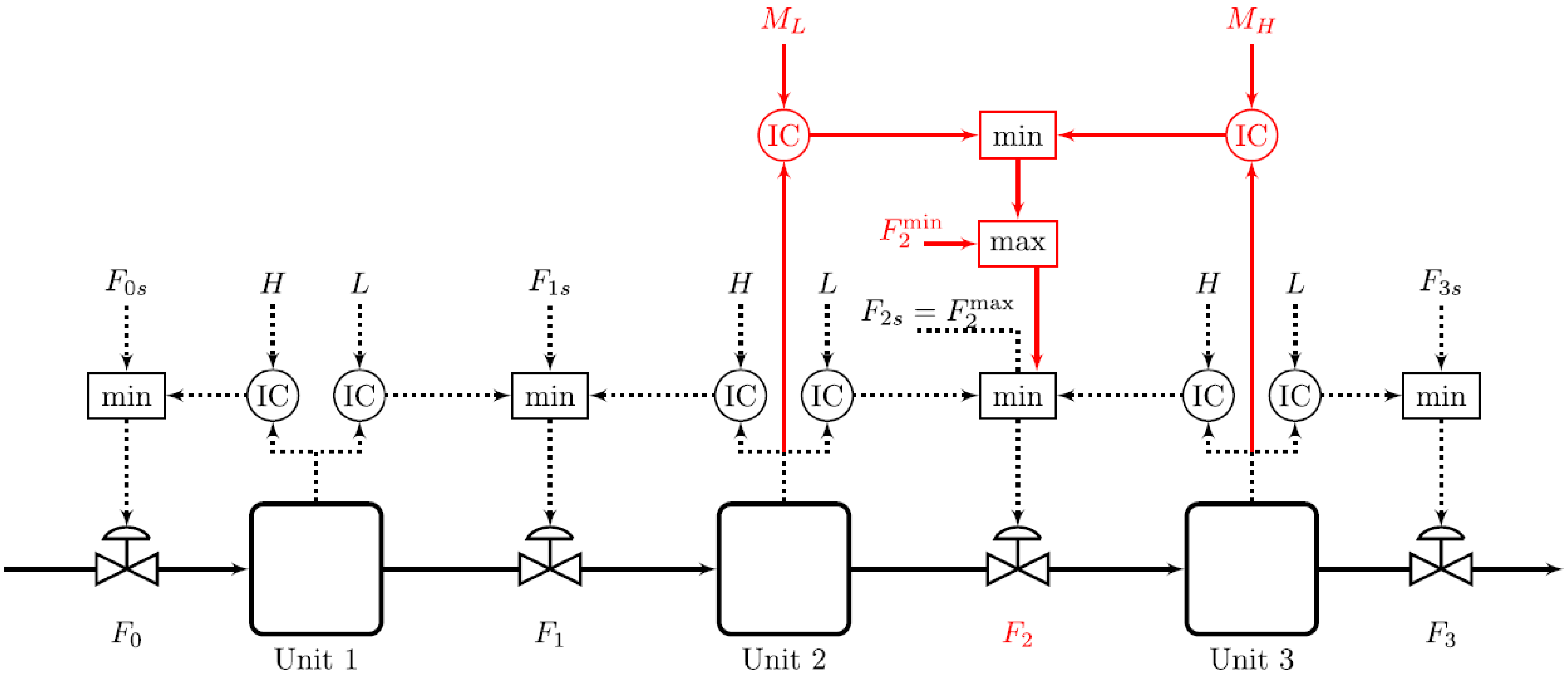
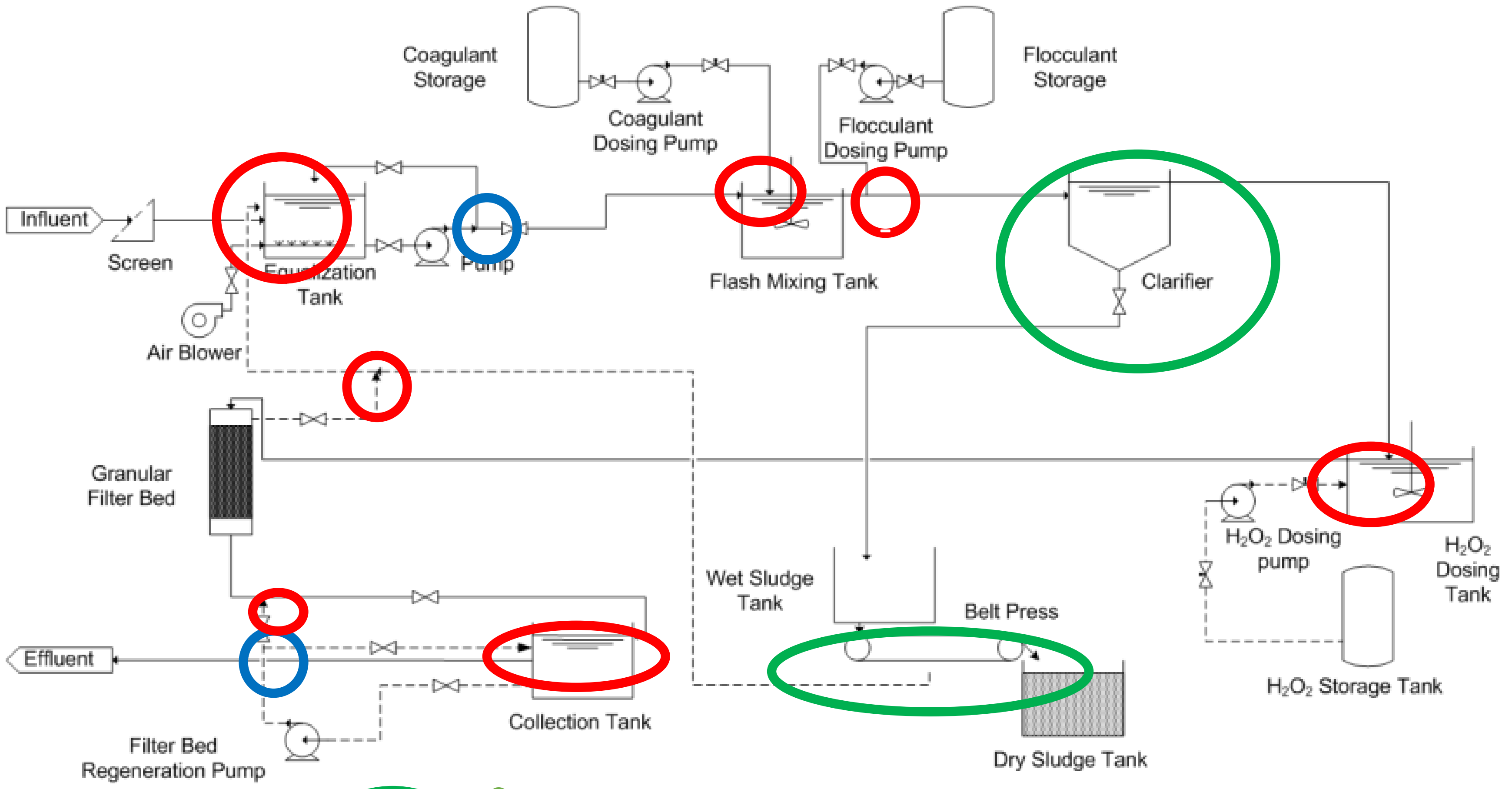


Fig. 37. Bidirectional inventory control scheme for maximizing throughput (dashed black lines) while attempting to satisfy minimum flow constraint on F_2 (red lines). H, L, M_L and M_H are inventory setpoints.

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.

Bidirectional control for plants with recycle



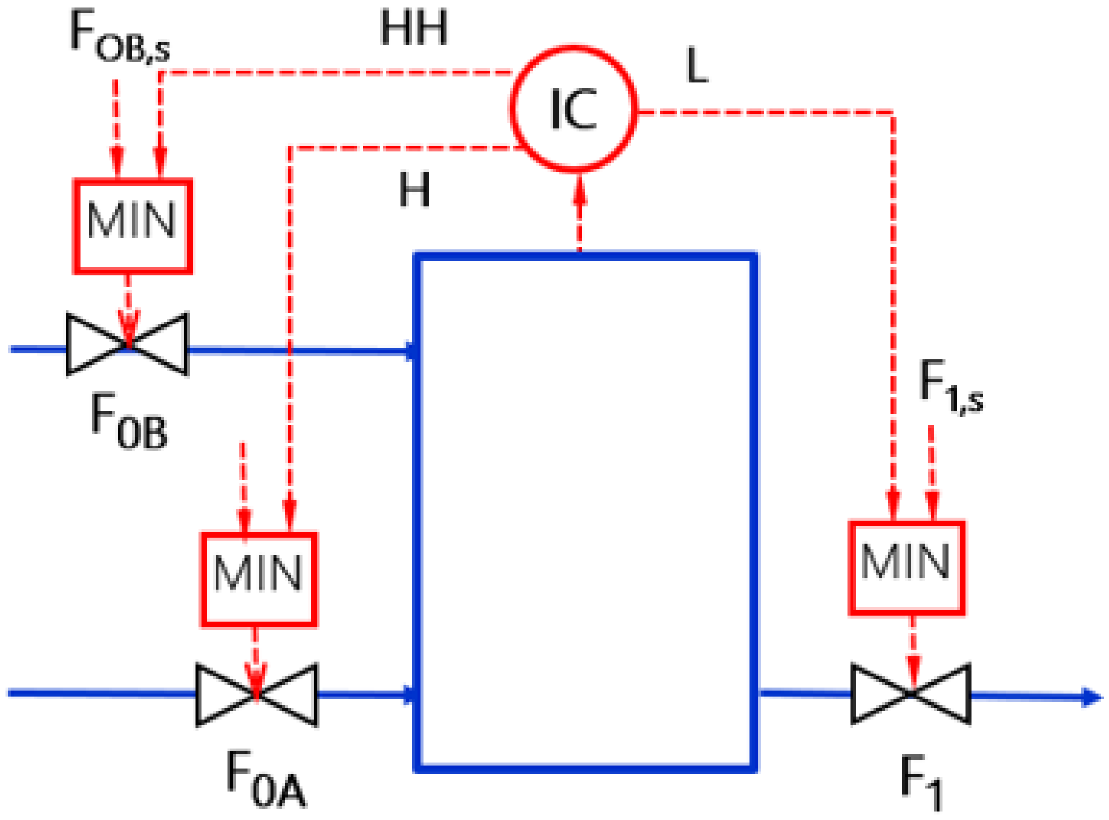
Mixing
(use ratio control for feeds)

Separator
(«fixed» split)

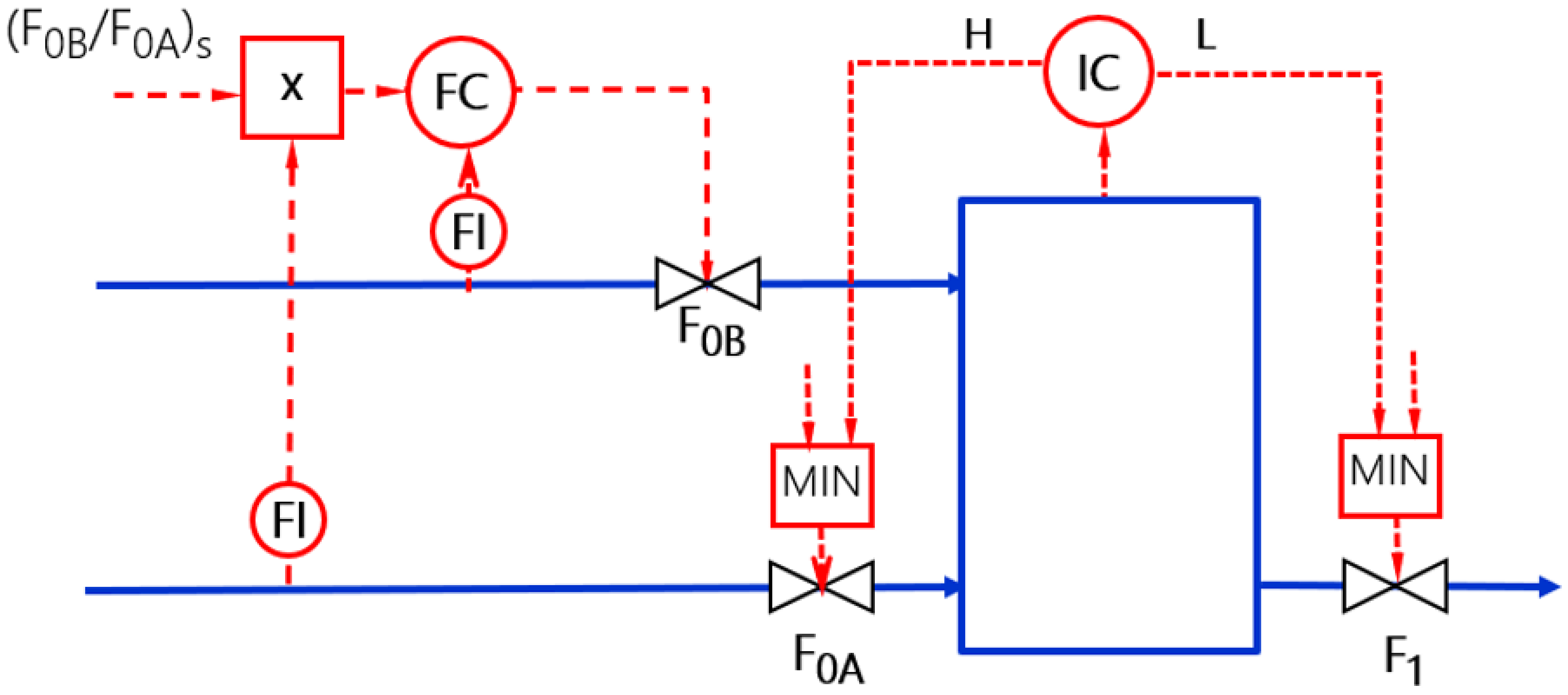
Adjustable split
(same composition)

Chemical Engineering Research Bulletin 14(2010) 65-68 / Rahman and Kabir

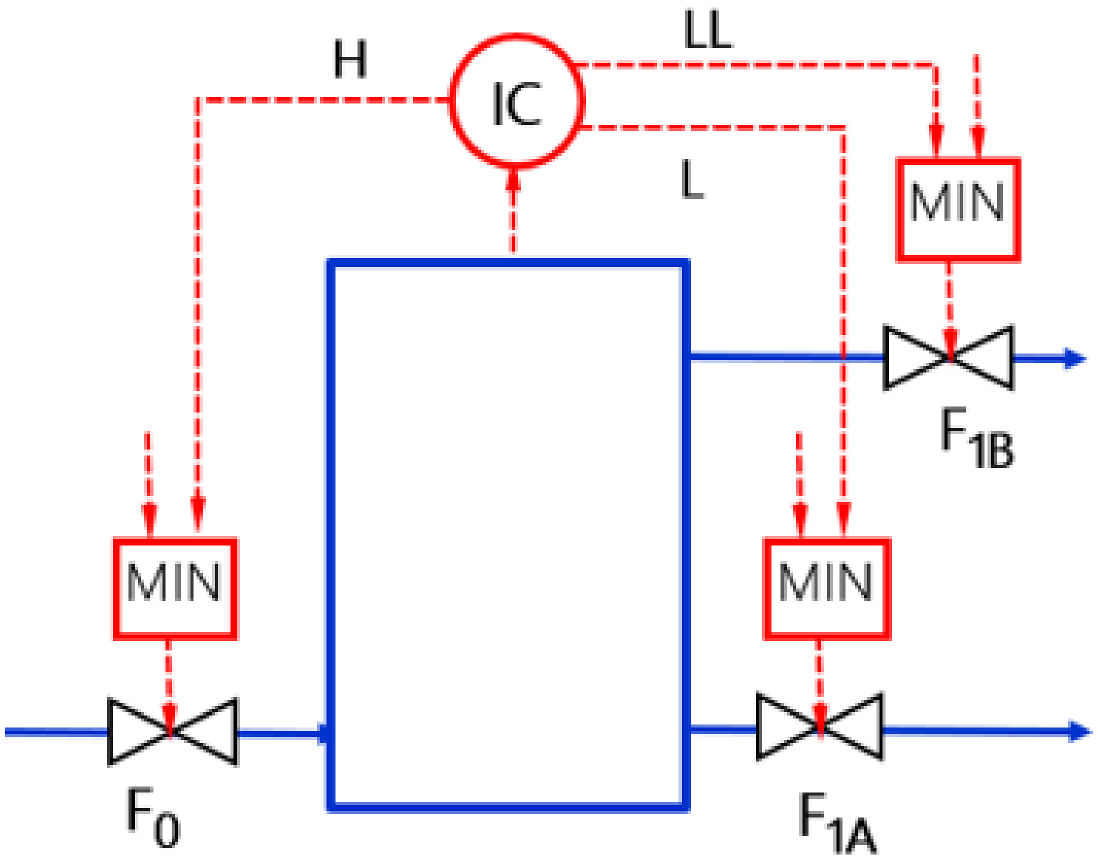
Mixing = Stream merging (junction)



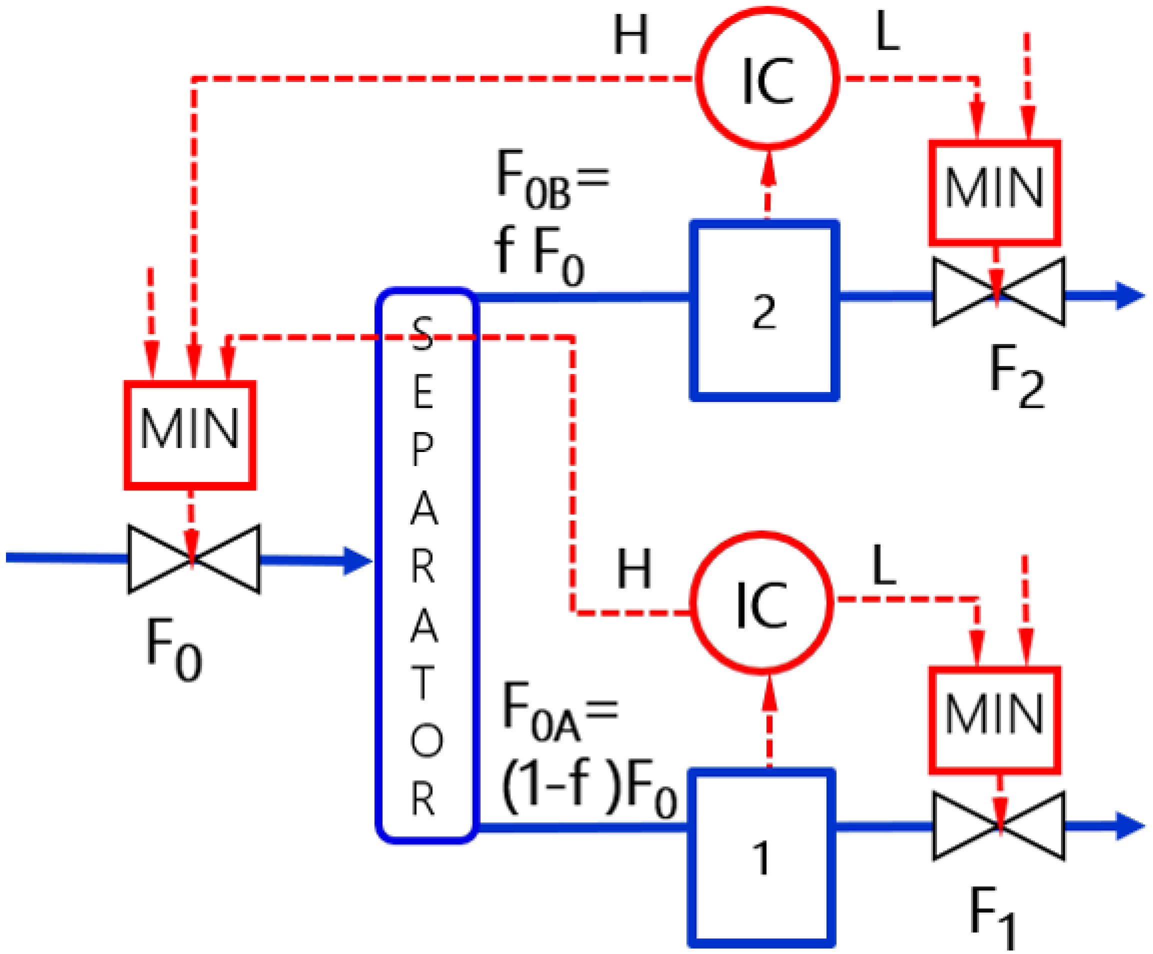
Stream merging (mixing) with ratio control



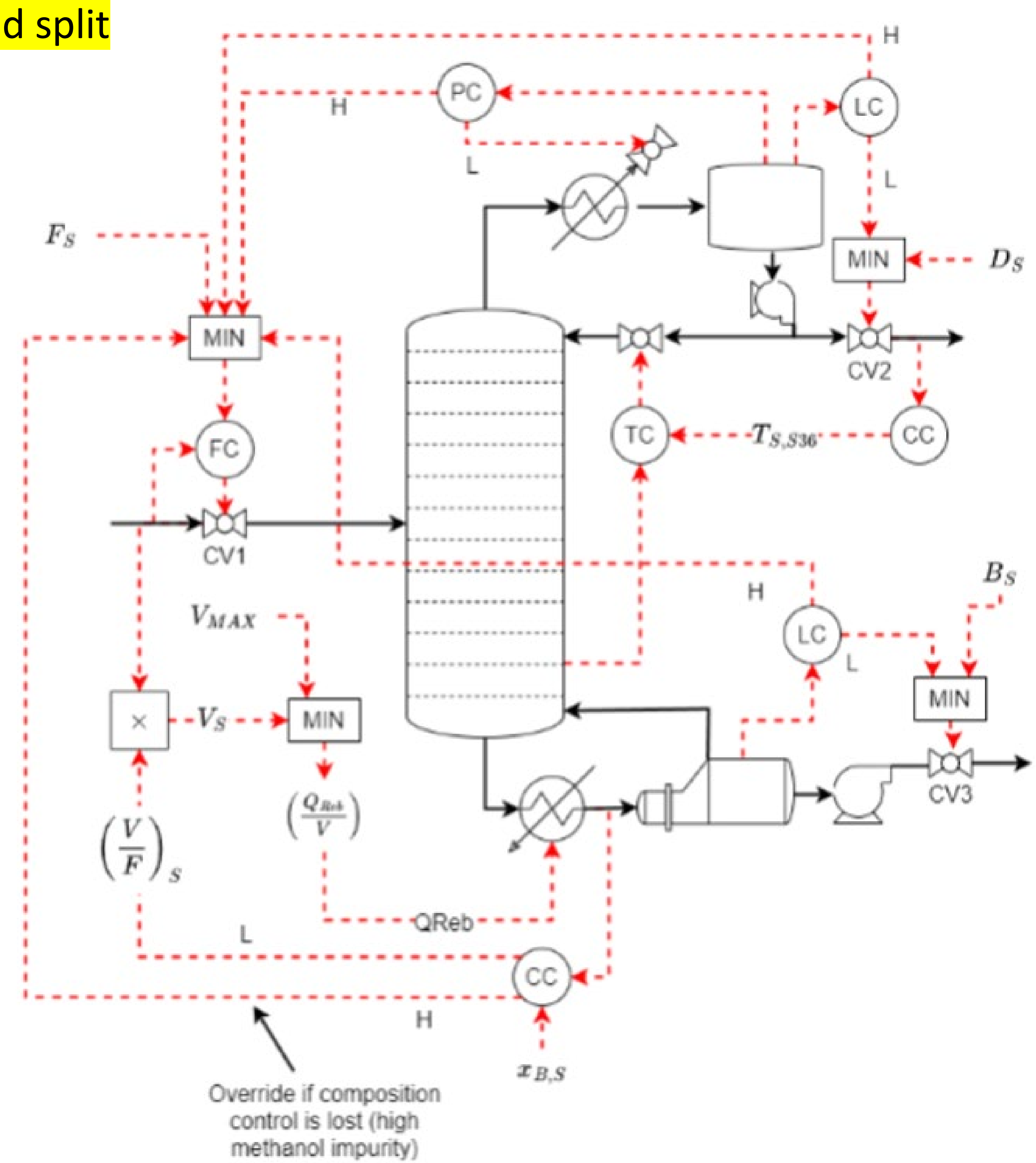
Adjustable split (same composition for F_{1A} and F_{1B})



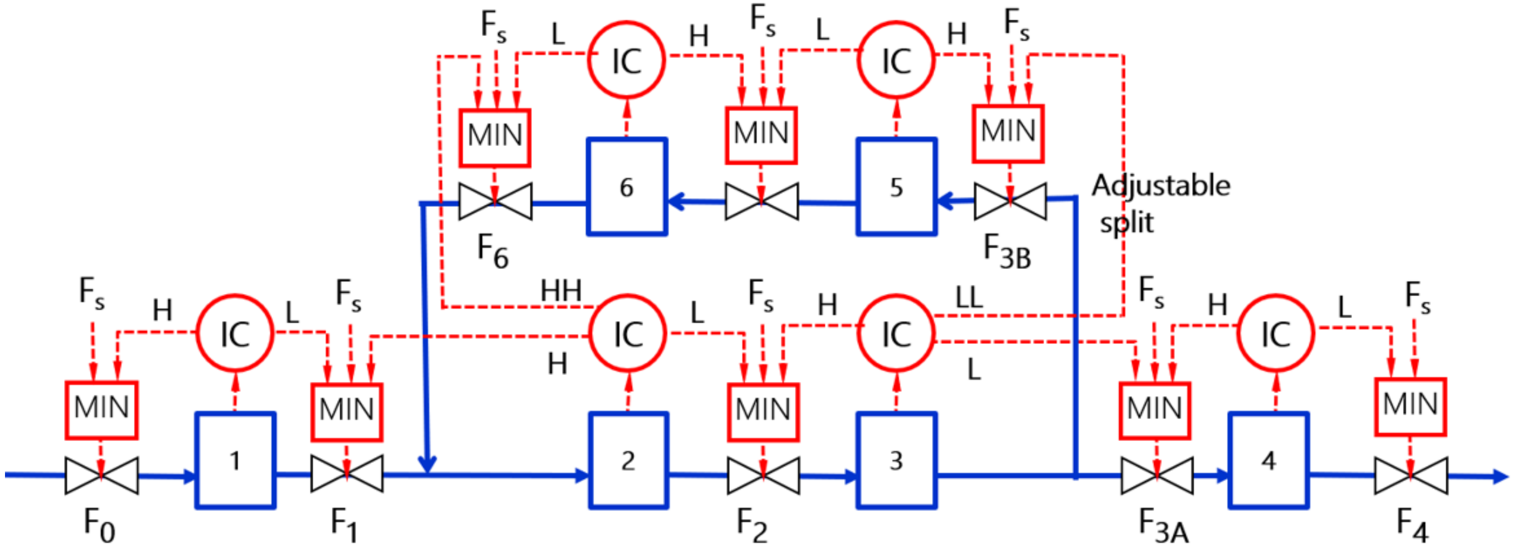
Fixed split fraction (separator, different compositions



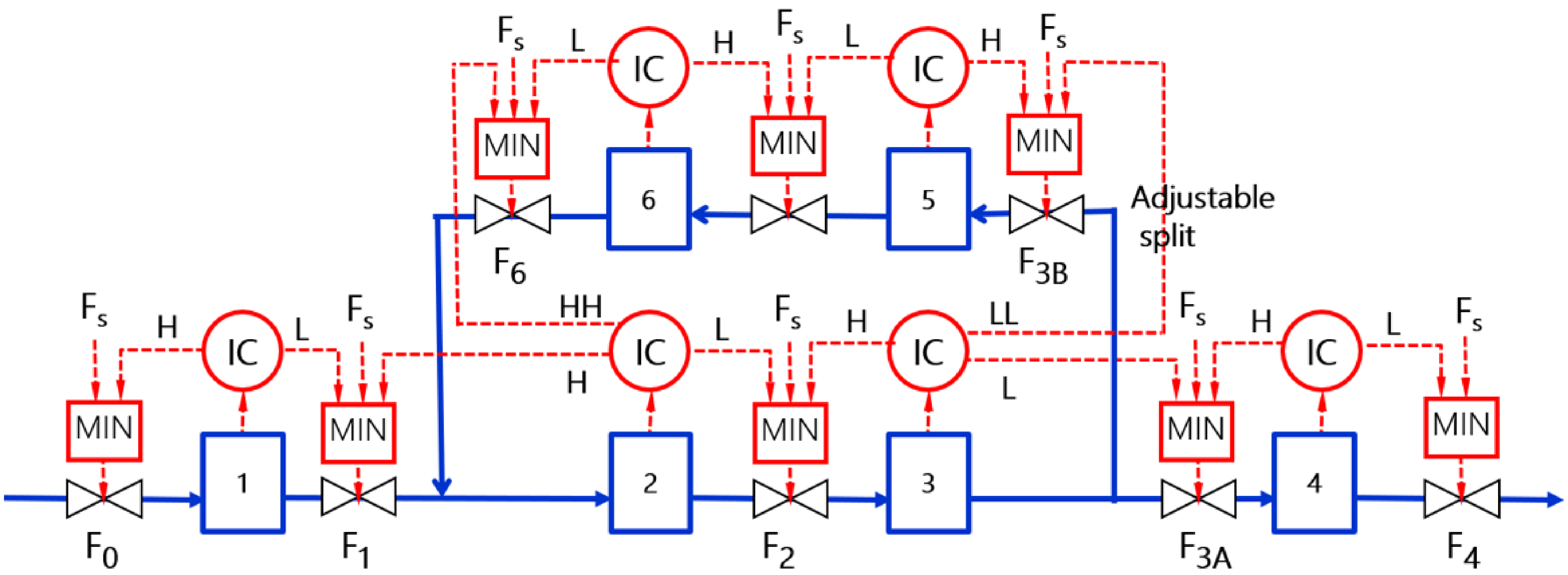
- For: Distillation, cyclone, filter, crystallizer, phase separator, Bidirectional inventory control with fixed split Distillation



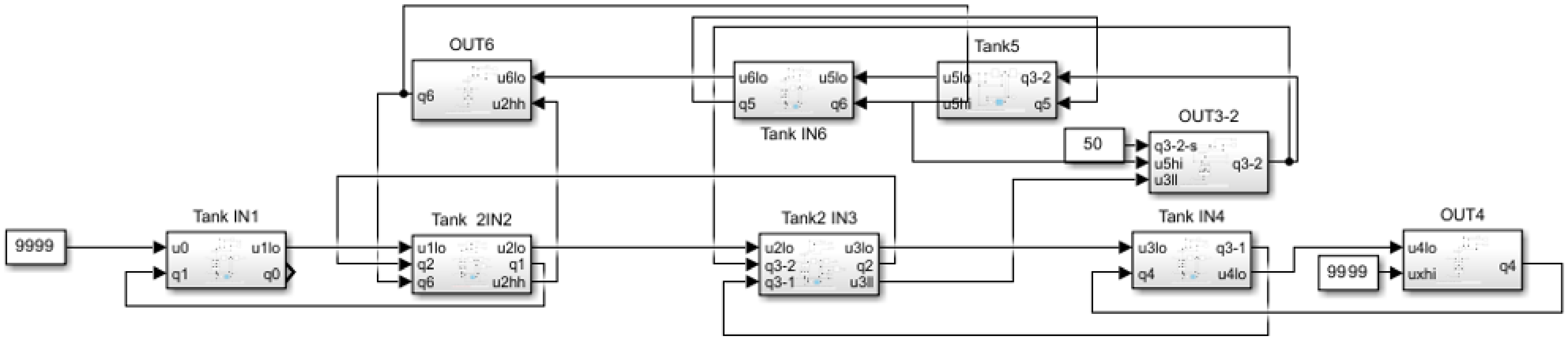
Recycle systems



Recycle example with adjustable split (set at split point)

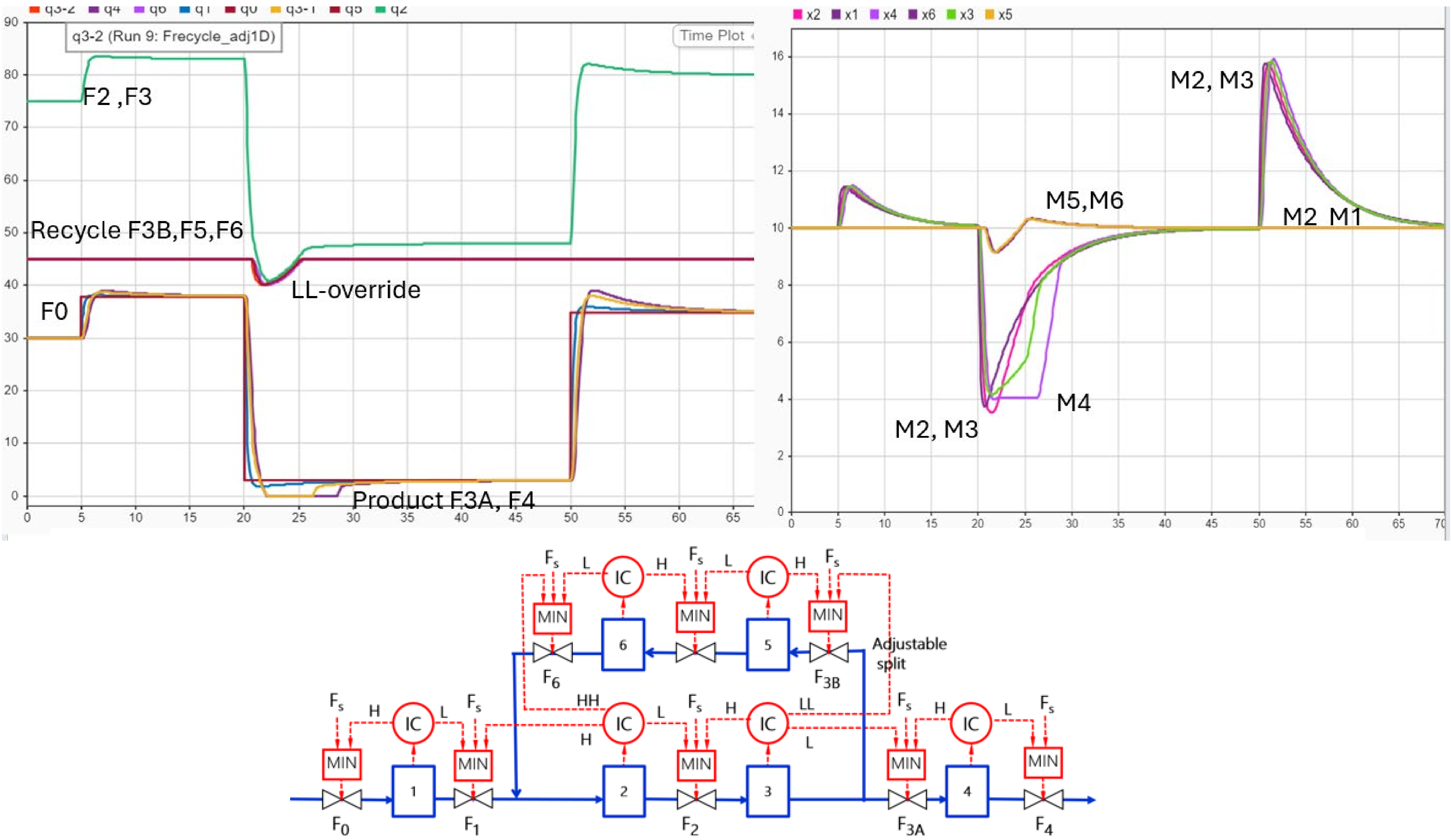


Simulations with Matlab

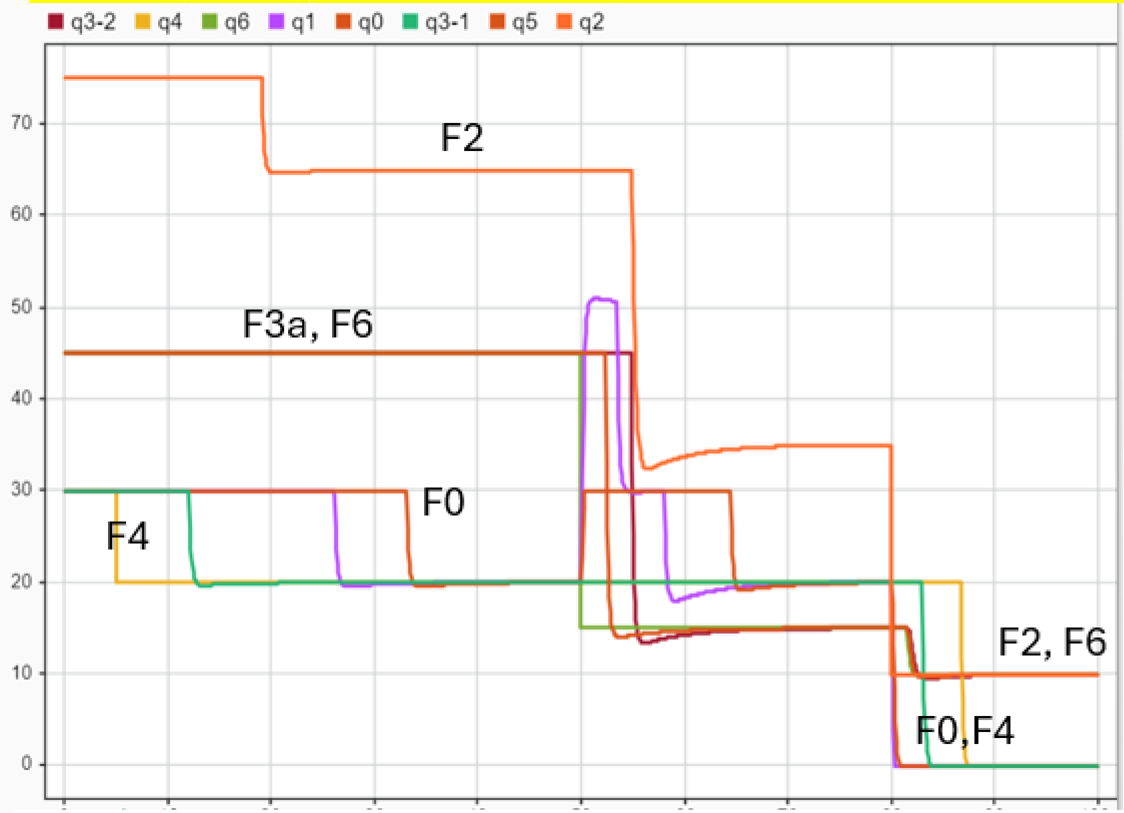


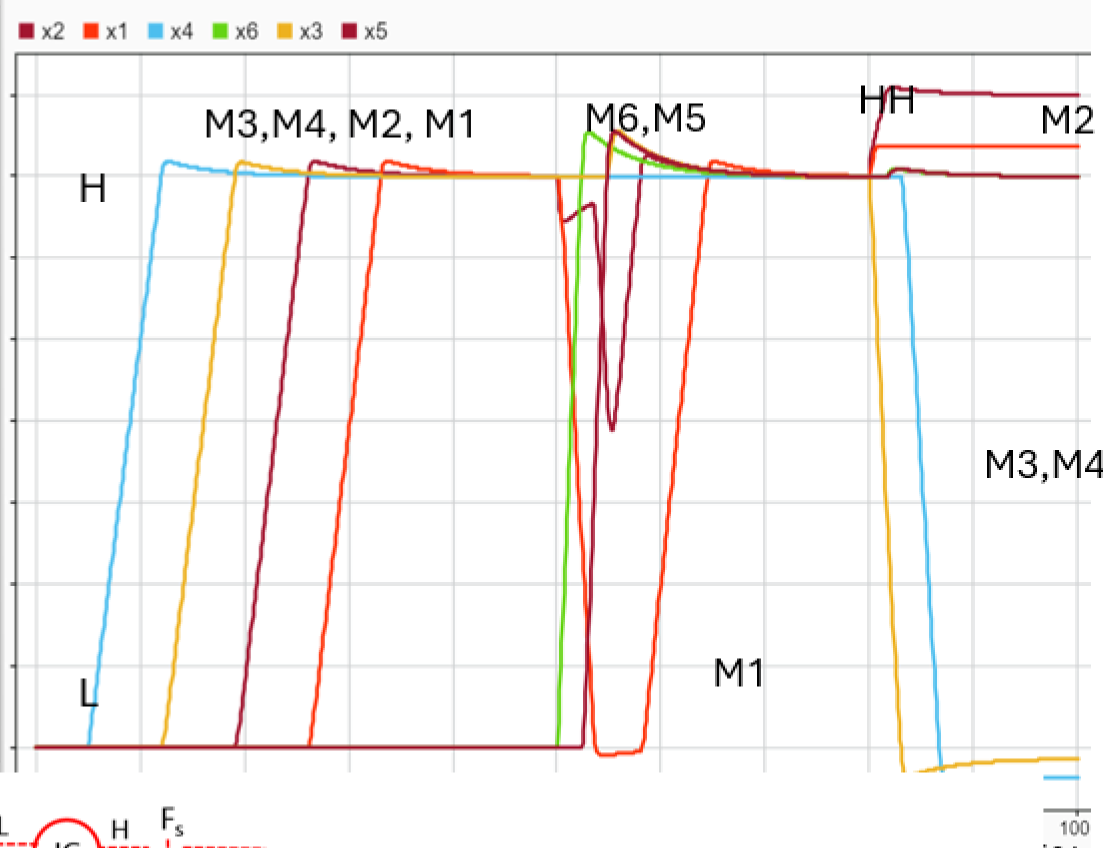
All 14 level controllers (6 H, 6 L, 1 HH, 1 LL):

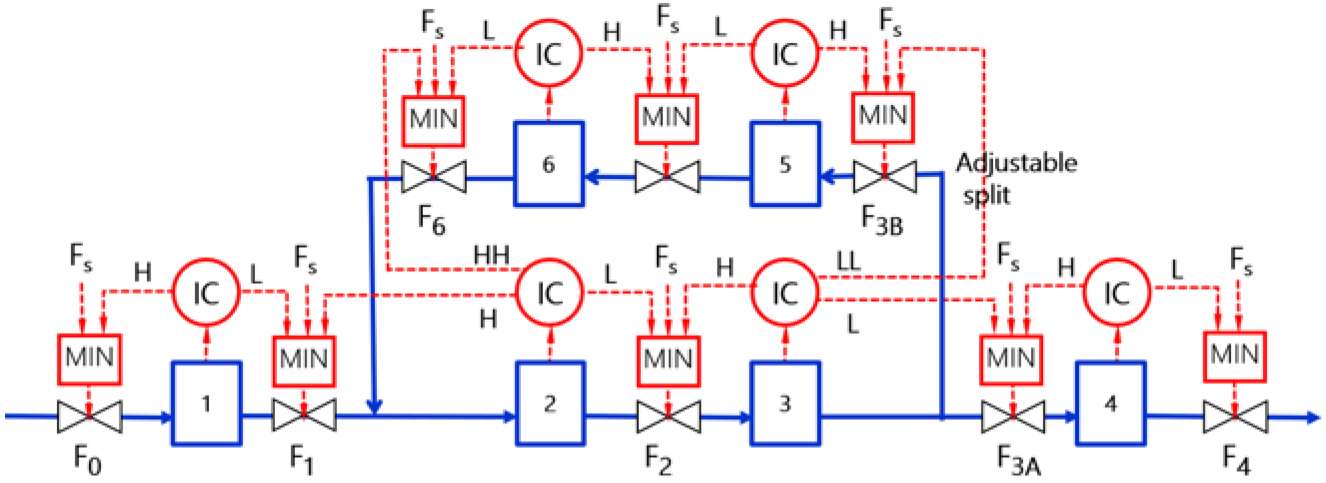
- Kc = 5 %/%
- Integral time, $\tau_I = 5\ \tau$ (where $\tau = M_{max}/F_{max}$ = redidence time =10 min)
  - No oscillations for single tank since Kc $\tau_I = 25\ \tau > 4\ \tau$
- Tracking time for anti-windup, $\tau_T = \tau_I$



q3-2
q4
q6
q1
q0
q3-1
q5
q2
q3-2 (Run 9: Frecycle_adj1D)
Time Plot
F2 ,F3
Recycle F3B,F5,F6
LL-override
F0
Product F3A, F4
x2
x1
x4
x6
x3
x5
M2, M3
M5,M6
M2 M1
M4
M2, M3
Fs
L
IC
H
MIN
6
5
Adjustable split
F6
F3B
HH
LL
1
2
3
4
F0
F1
F2
F3A
F4


q3-2 q4 q6 q1 q0 q3-1 q5 q2
70
60
50
40
30
20
10
0
F2
F3a, F6
F0
F4
F2, F6
F0,F4

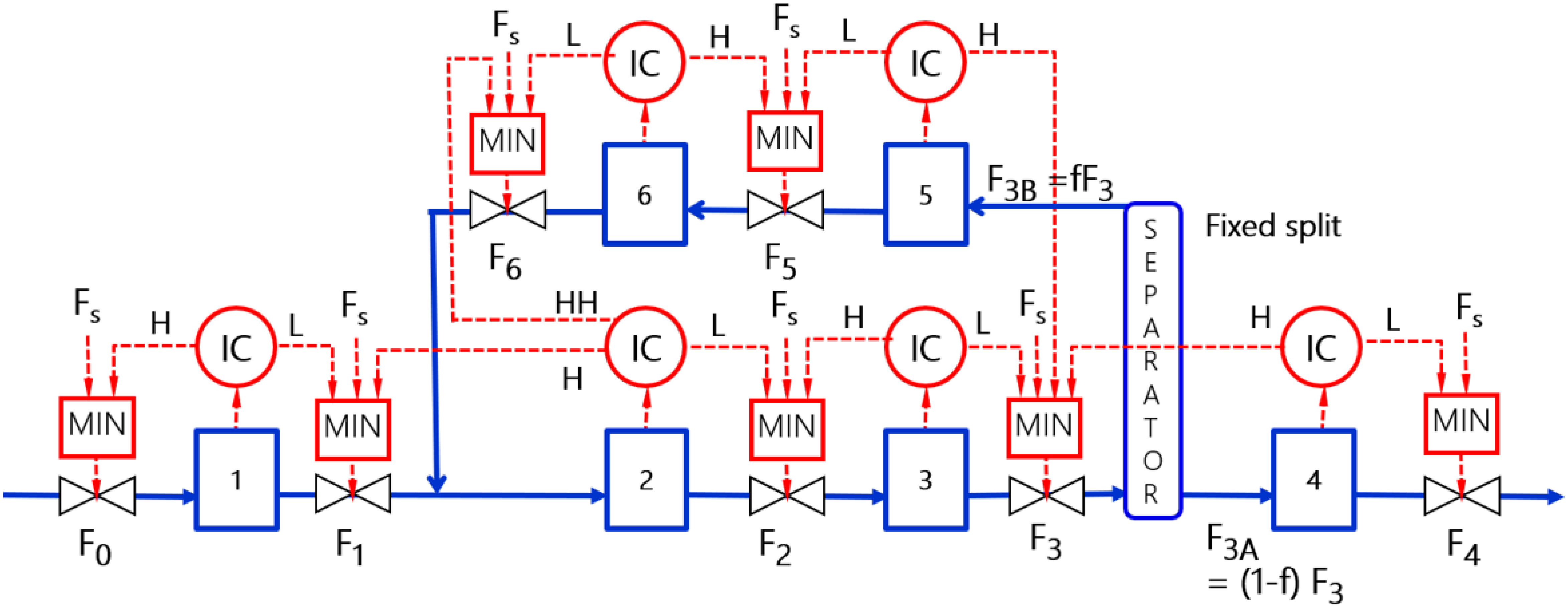

x2 x1 x4 x6 x3 x5
M3,M4, M2, M1
M6,M5
HH
M2
H
L
M3,M4
M1
100

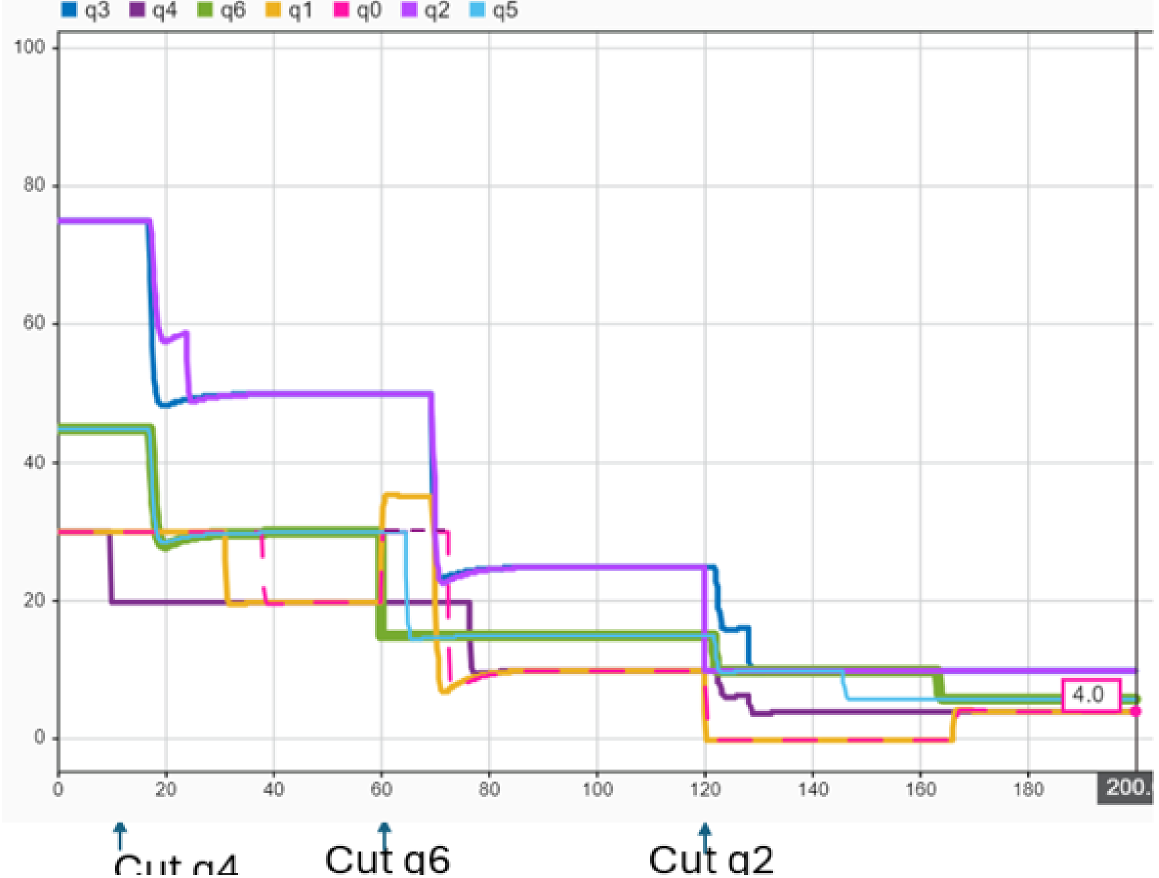

IC
MIN
Fs
H
L
HH
LL
1
2
3
4
5
6
Adjustable split
F_0
F_1
F_2
F_{3A}
F_{3B}
F_4
F_6

%

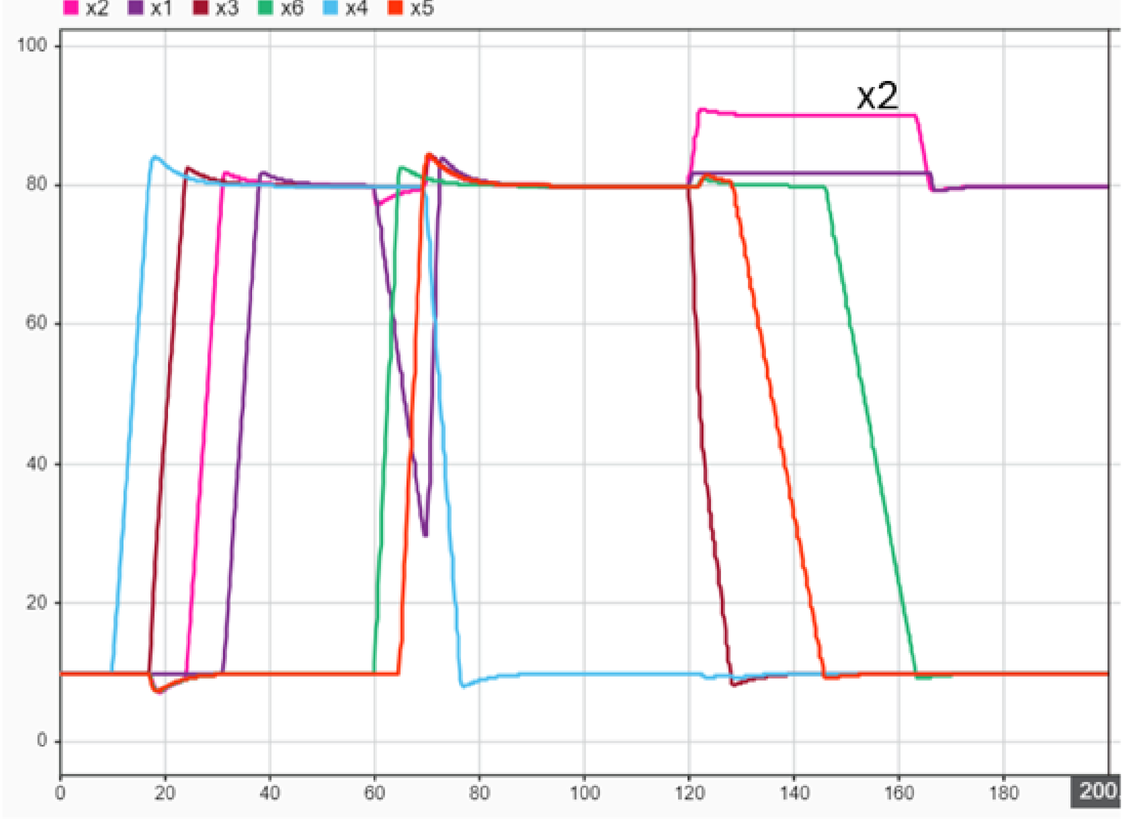
ne

Recycle example with separator (fixed split)

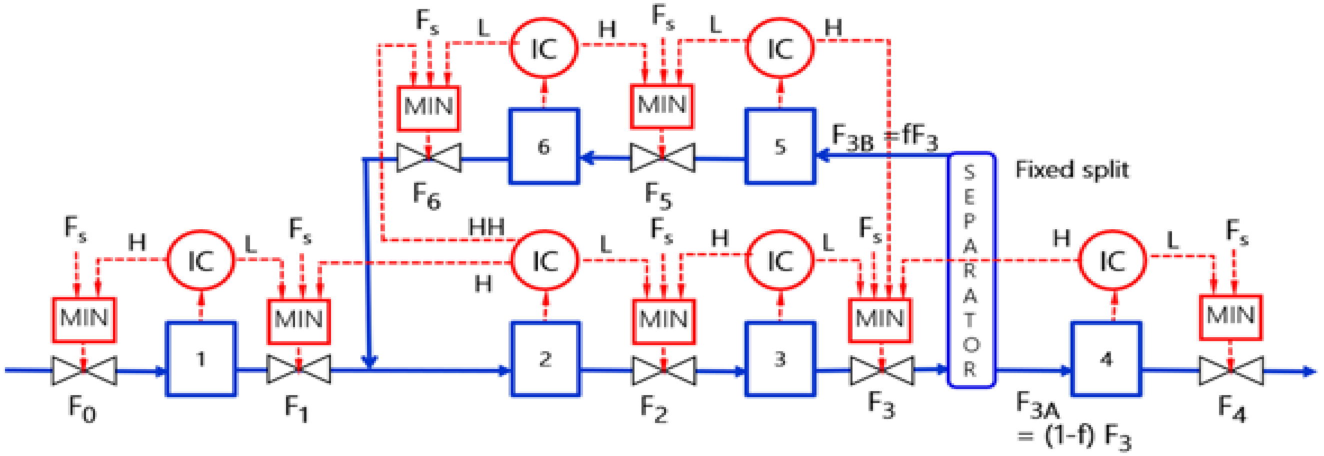




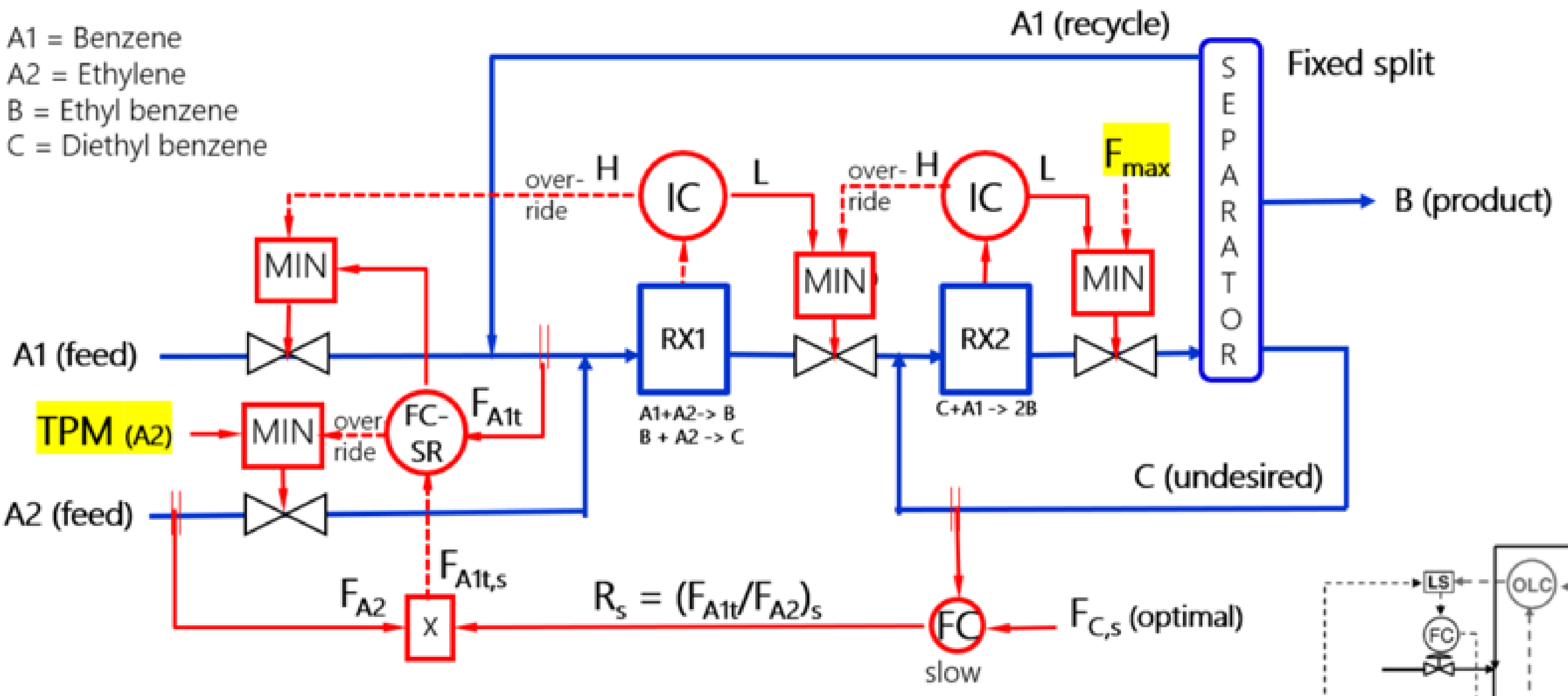
q3 q4 q6 q1 q0 q2 q5
100
80
60
40
20
0
0 20 40 60 80 100 120 140 160 180 200.
4.0
Cut q4
Cut q6
Cut q2



x2 x1 x3 x6 x4 x5
x2
100
80
60
40
20
0
0 20 40 60 80 100 120 140 160 180 200.



IC
MIN
F_s
L
H
HH
1
2
3
4
5
6
F_0
F_1
F_2
F_3
F_4
F_5
F_6
F_{3B} =fF_3
Fixed split
SEPARATOR
F_{3A} = (1-f) F_3



nax is reduced when we reach Vmax in separator (column), and then overrides are activated.
itio control is maintained as a fast loop under all conditions. The FC for C=DEB is slow.

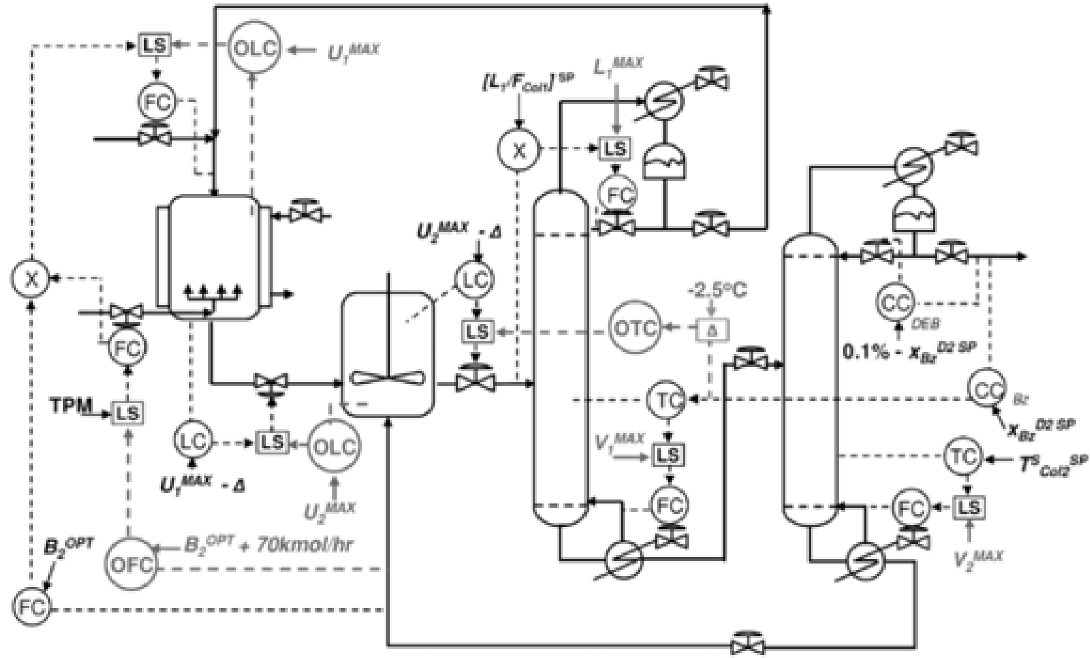


Figure 7. CS2 with overrides for handling equipment capacity constraints.

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

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

Academic process control community fish pond

Optimal centralized
Solution (**EMPC**)

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