TPM + Plantwide control rules

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Outline

Skogestad procedure for control structure design:

I. Top Down
   - Step S1: Define operational objective (cost) and constraints
   - Step S2: Identify degrees of freedom and optimize operation for disturbances
   - Step S3: Implementation of optimal operation
     - What to control? (primary CV's) (self-optimizing control)
   - Step S4: Where set the production rate? (Inventory control)

II. Bottom Up
   - Step S5: Regulatory control: What more to control (secondary CV's)?
   - Step S6: Supervisory control
   - Step S7: Real-time optimization

Step S4. Where set production rate?

- "Throughput manipulator" (TPM)
  = MV for controlling throughput (production rate, network flow)
- Where set the production rate = Where locate the TPM?
  - Traditionally: At the feed
  - For maximum production (with small backoff): at the bottleneck

- Very important **dynamic** decision that determines the structure of the rest of the inventory control system!
- May also have important economic implications
- Link between **Top-down** (economics) and **Bottom-up** (stabilization) parts
  - Inventory control is the most important part of stabilizing control

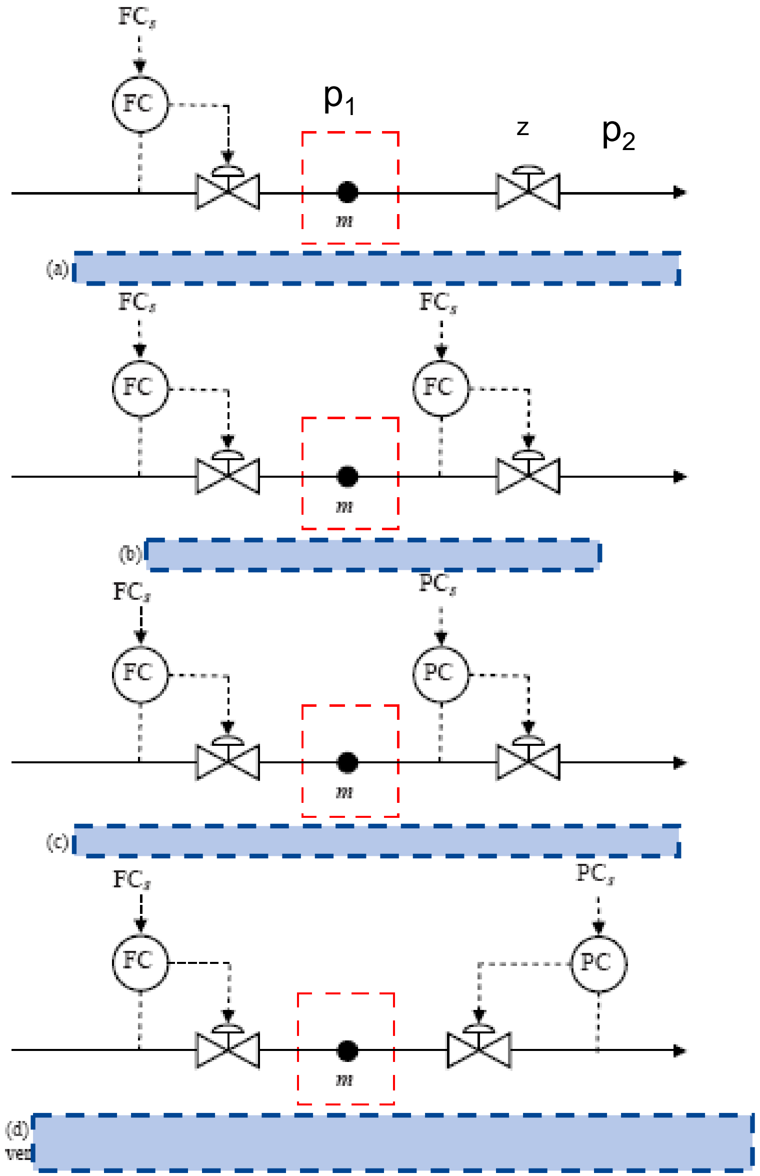
Consistency of inventory control

- **Consistency** (required property):

  An inventory control system is said to be **consistent** *if the steady-state mass balances (total, components and phases) are satisfied for any part of the process, including the individual units and the overall plant.*

QUIZ 1

Consistent?



Flow rate: $q = C_v f(z)\sqrt{\frac{p_1 - p_2}{\rho}}$ $[\mathrm{m}^3/\mathrm{s}]$

Controlling pressure at outlet is possible, but not consistent! (*m* is uncontrolled)

Generalization of consistency to components and phases

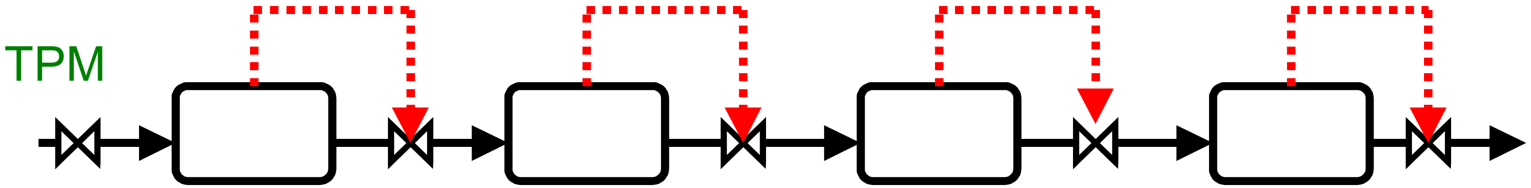
Rule . ***Consistency*** *requires that*

1. ***The total inventory (mass) of any part of the process must be regulated by its in- or outflows****, which implies that at least one flow in or out of any part of the process must depend on the inventory inside that part of the process.*
2. *For systems with several components, the* ***inventory of each component*** *of any part of the process must be regulated by its in- or outflows or by chemical reaction.*
   - *"All components must find a way out".*
   - *Example: May need to add extra separator, purge or reaction combined with recycle*
3. *For systems with several phases, the* ***inventory of each phase*** *of any part of the process must be regulated by its in- or outflows or by phase transition.*
   - *Example: Flash tank*

Proof: Mass balances

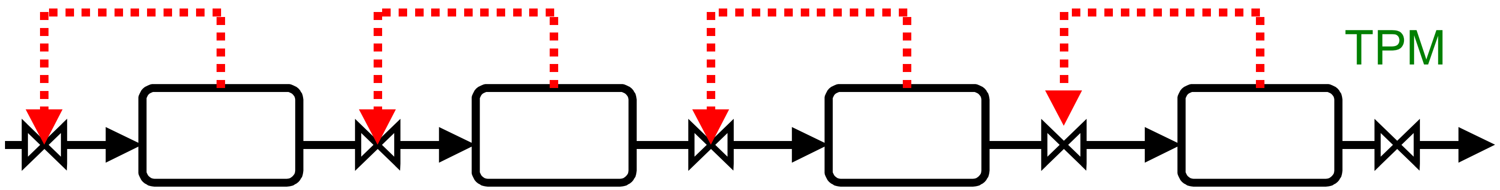
Inventory control of series of units

Production rate set at inlet:
Inventory control in direction of flow*



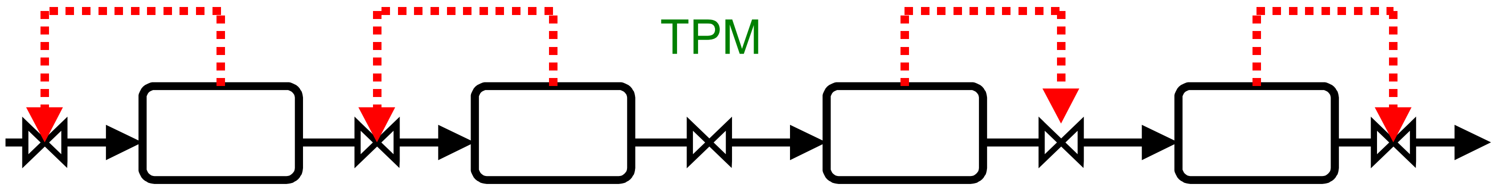
*Required to get “local-consistent” inventory control
Should also follow “par close” rule to avoid “long loops”

Production rate set at outlet:
Inventory control opposite flow*



* Required to get “local-consistent” inventory control

Production rate set inside process*



* Required to get “local-consistent” inventory control

General: “Need radiating inventory control around TPM” (Georgakis)

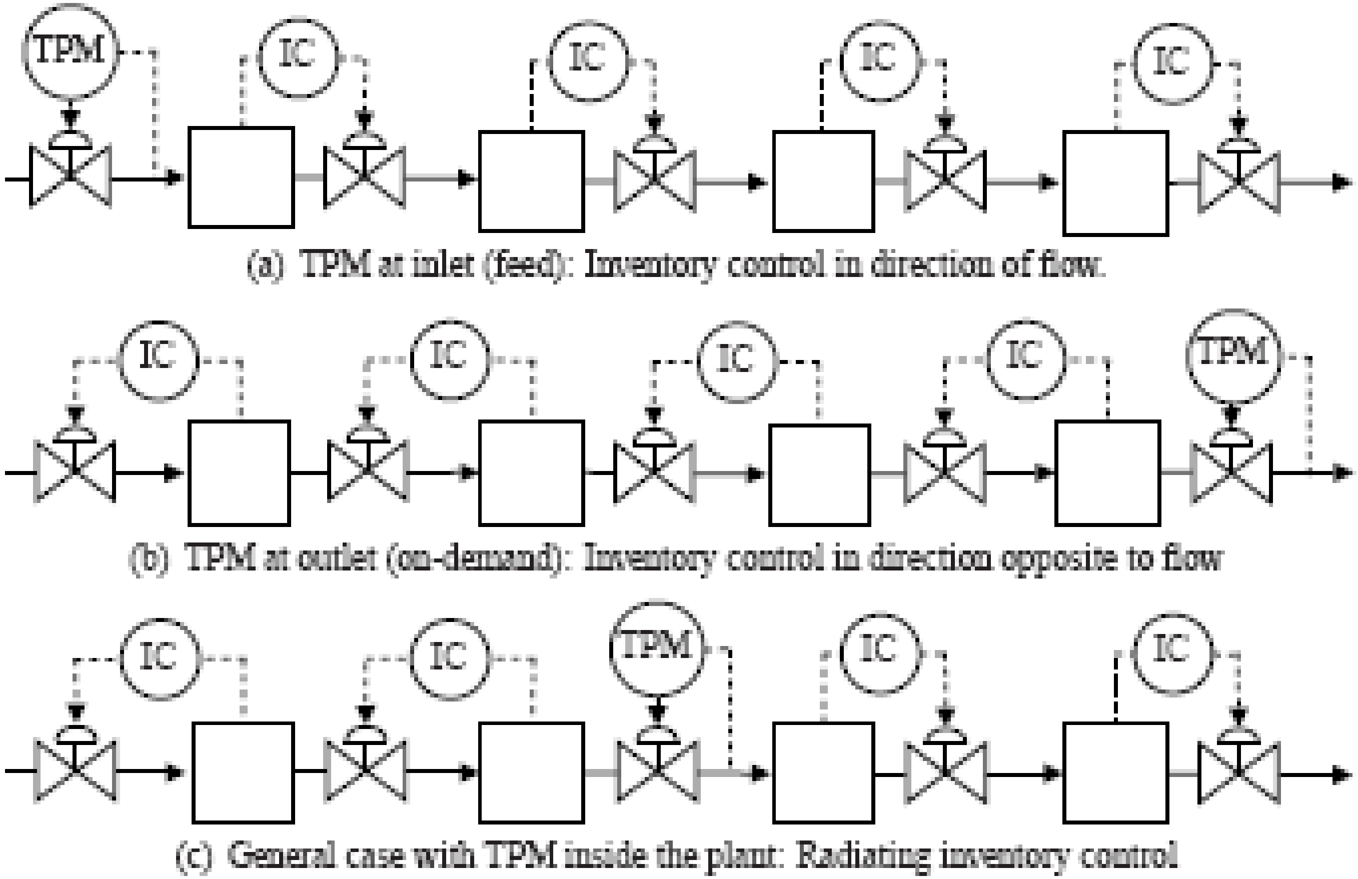


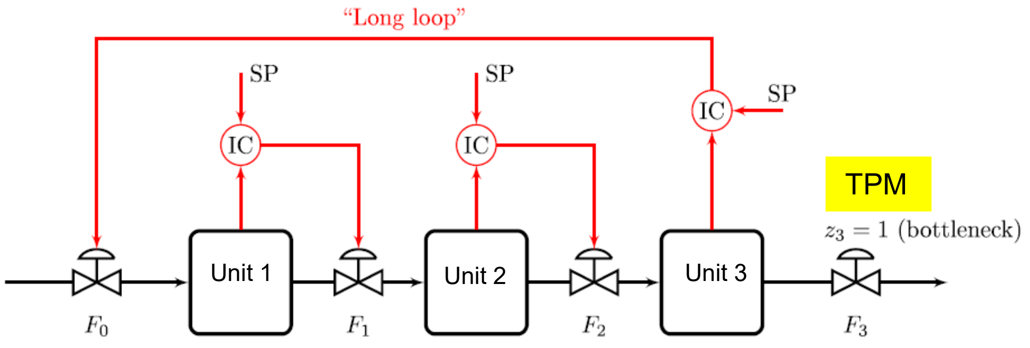
Figure 2.6: Self-consistency requires a radiating inventory control around a fixed flow (TPM)

«Long loop»

- Loop where initial response depends on other loops being closed to work
- Typical case: pair variables with process gain = 0 (open loop)

Radiation rule for inventory control

Breaking the radiation rule locally leads to undesirable «long loops»*:



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

Comment: Originally the TPM was at the feed (F_0) - but then the outflow (F_3) reached saturation (so this became the TPM) – and we let F_0 take over the inventory control in the last unit.
This may work OK if the inventory control in units 1 and 2 is very fast.

*«Long loop»: Does not follow the «pair close» rule,
so the (inital) functioning of a long loop depends on other loops being closed.

Avoiding long loops: Local-consistency rule

Rule. *Local-consistency* *requires that*

1. ***The total inventory (mass) of any part of the process must be locally regulated by its in- or outflows**, which implies that at least one flow in or out of any part of the process must depend on the inventory inside that part of the process.*
2. *For systems with several components, the **inventory of each component** of any part of the process must be locally regulated by its in- or outflows or by chemical reaction.*
3. *For systems with several phases, the **inventory of each phase** of any part of the process must be locally regulated by its in- or outflows or by phase transition.*

Proof: Mass balances

Note: Without the word "local" one gets the more general consistency rule

"Local" means follow "pair-close" rule and avoid "long loops"

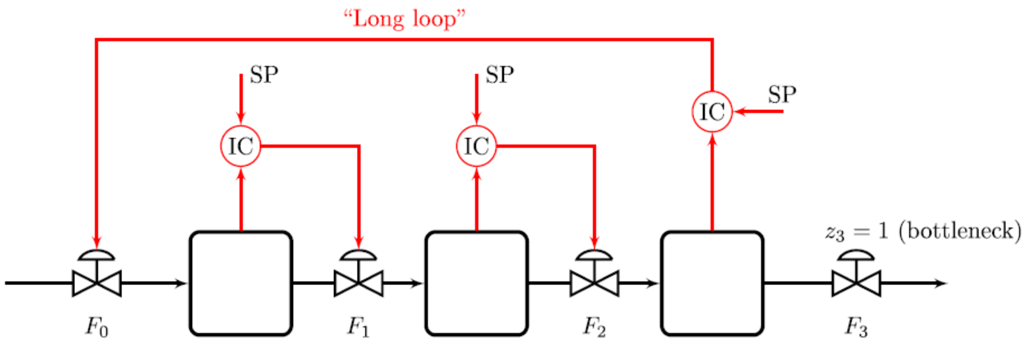
Summary: Rules for inventory control

Rule 1. Cannot control (set the flowrate) the same flow twice

Rule 2. Controlling inlet or outlet pressure indirectly sets the flow (indirectly makes it a TPM)

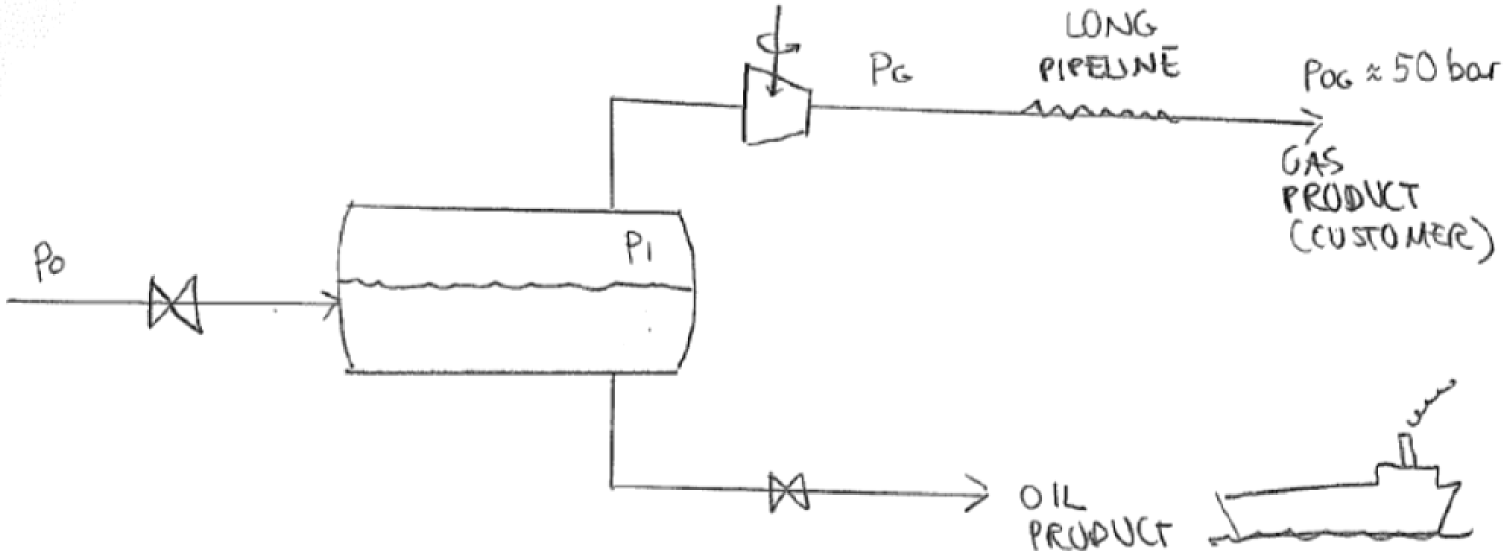
Rule 3. Follow the **radiation rule** whenever possible

Rule 4. No inventory loop can cross the TPM

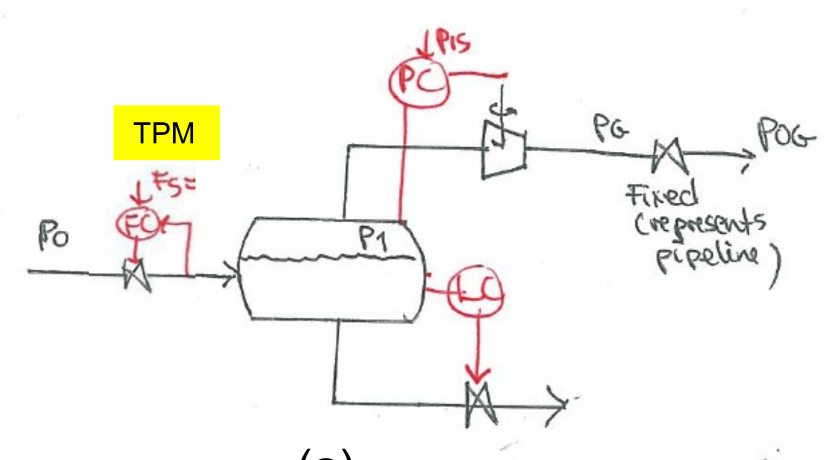


(d) Inventory control with undesired "long loop", not in accordance with the "radiation rule" (for given product flow, TPM= F_3)

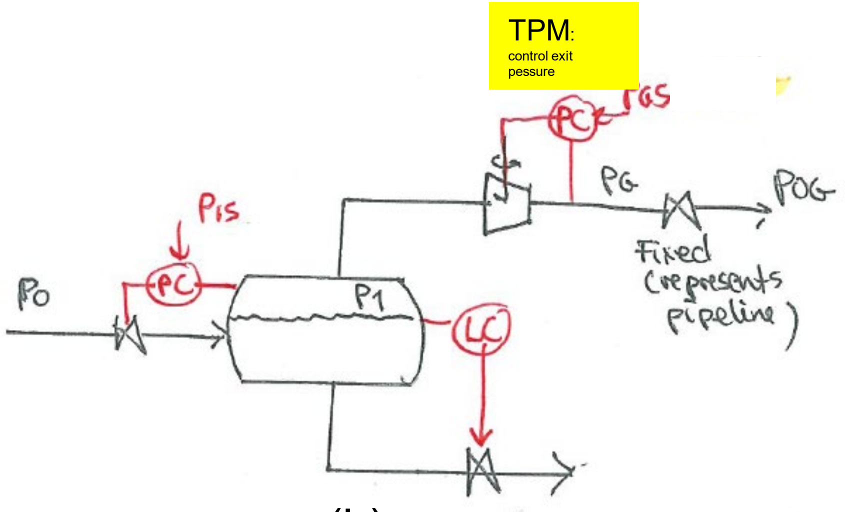
Example: Separator control (oil-gas separation offshore)



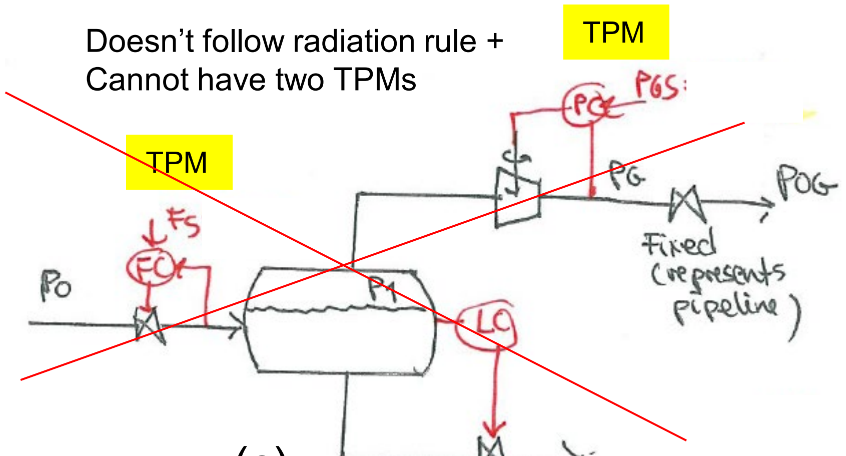
Quiz 2. Gas-liquid separator.
Where is TPM? Consistent (One is not)?



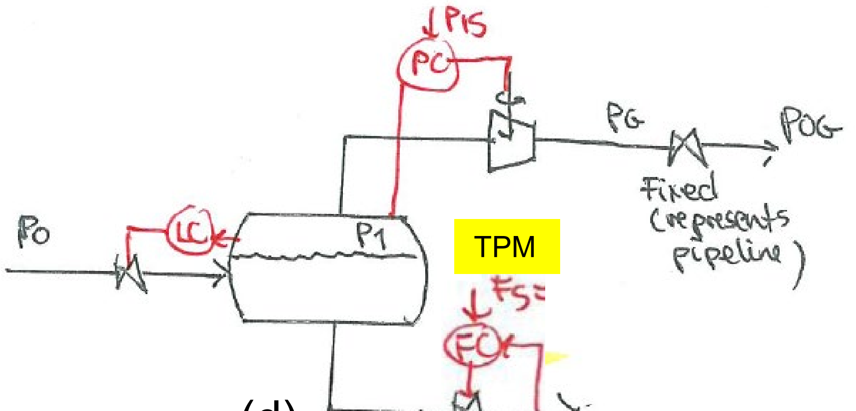
(a)



(b)



(c)

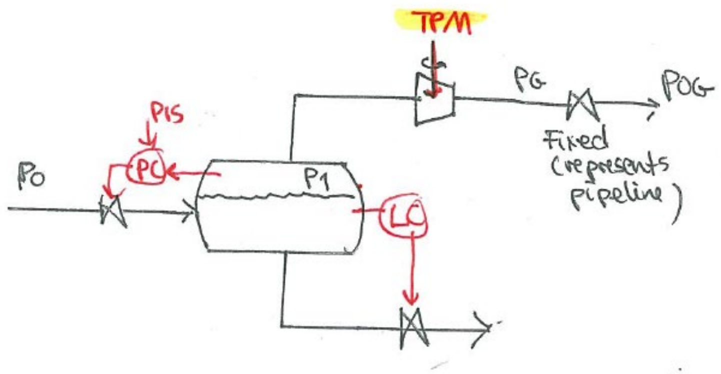


(d)

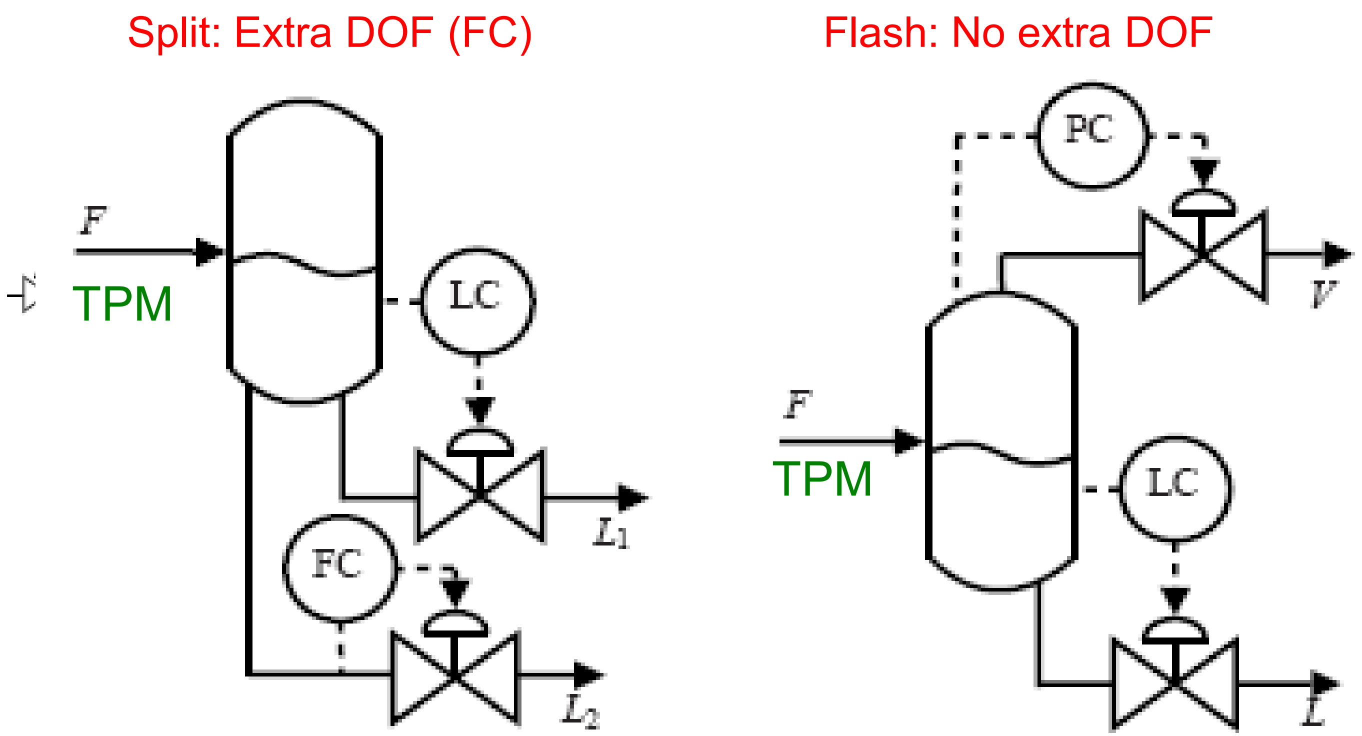
Case (a): Given feedrate. Could alternatively set p_0
Cases (b) and (c): Gas production limiting
Case (d): Liquid production limiting

Rule: Setting in-pressure p_0 sets inflow = TPM at inlet or inlet direction (no cases above)
Setting out-pressure p_G sets outflow = TPM at outlet or outlket direction (offdiagonal two cases)

More?



Flow split: May give extra DOF



(a) Adjustable split: Single-phase tank

(b) Fixed split: Two-phase tank

Figure 2.5: Adjustable split introduces a degree of freedom but a phase transition requires that all phases are on inventory control.

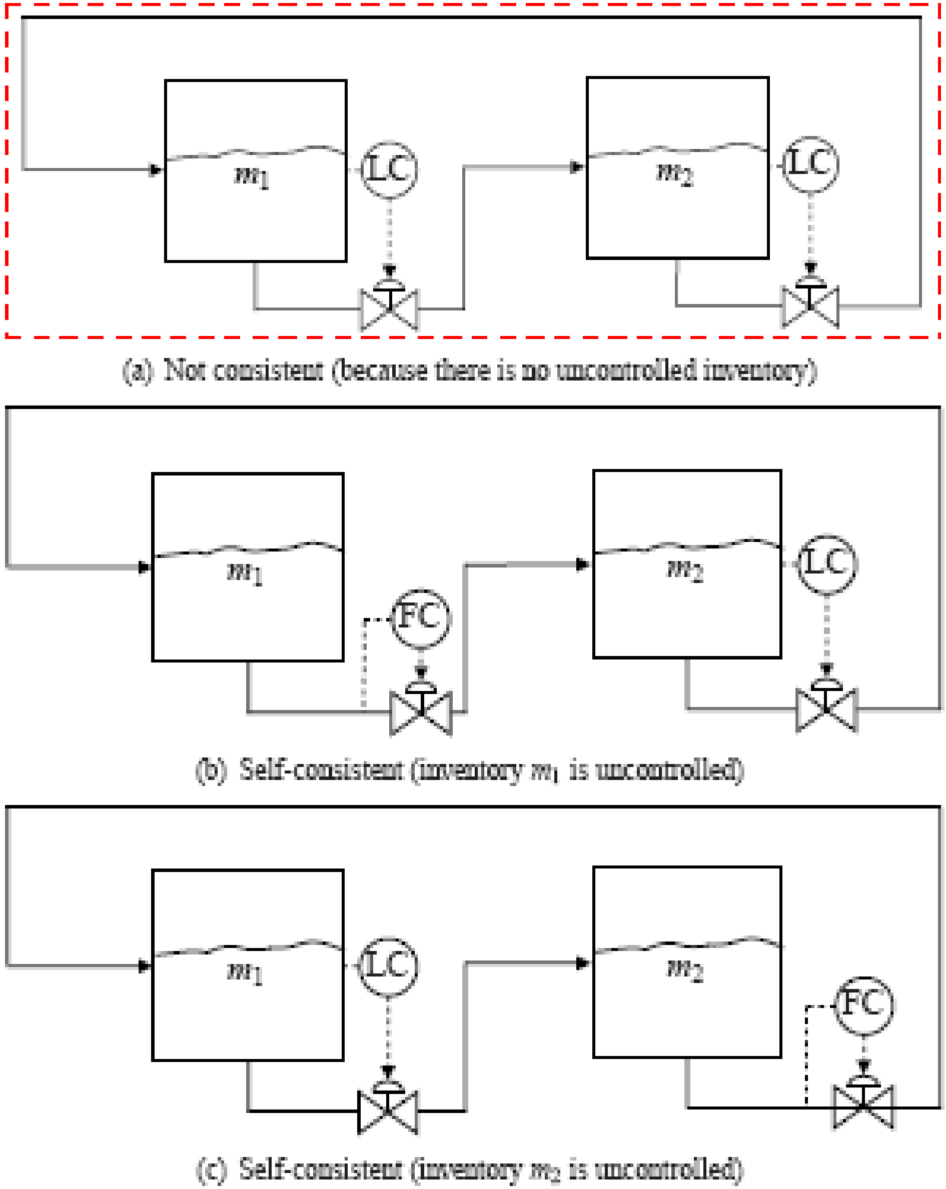


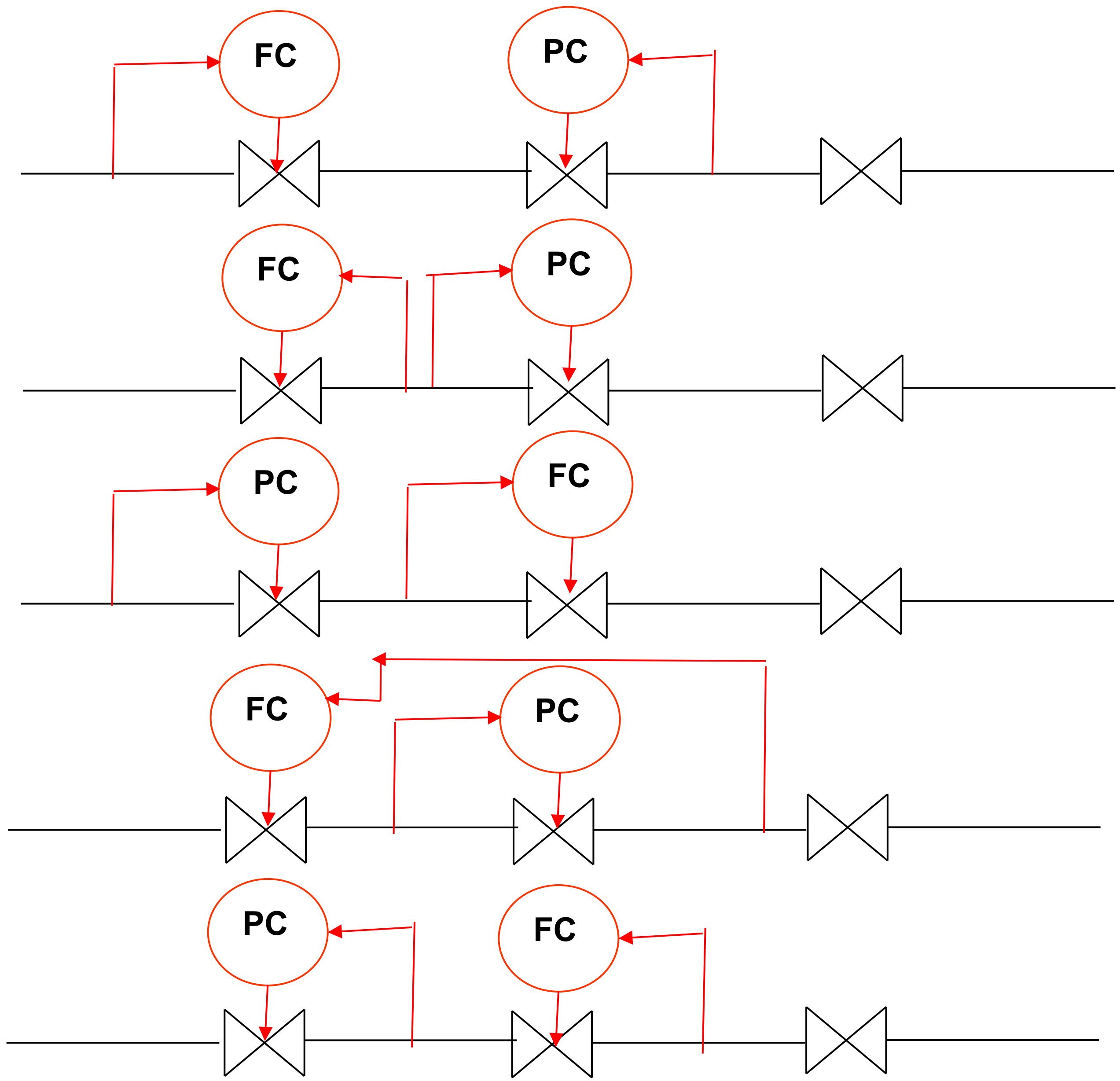
Figure 2.8: Inventory control for closed system.

Closed system: Must leave one inventory uncontrolled

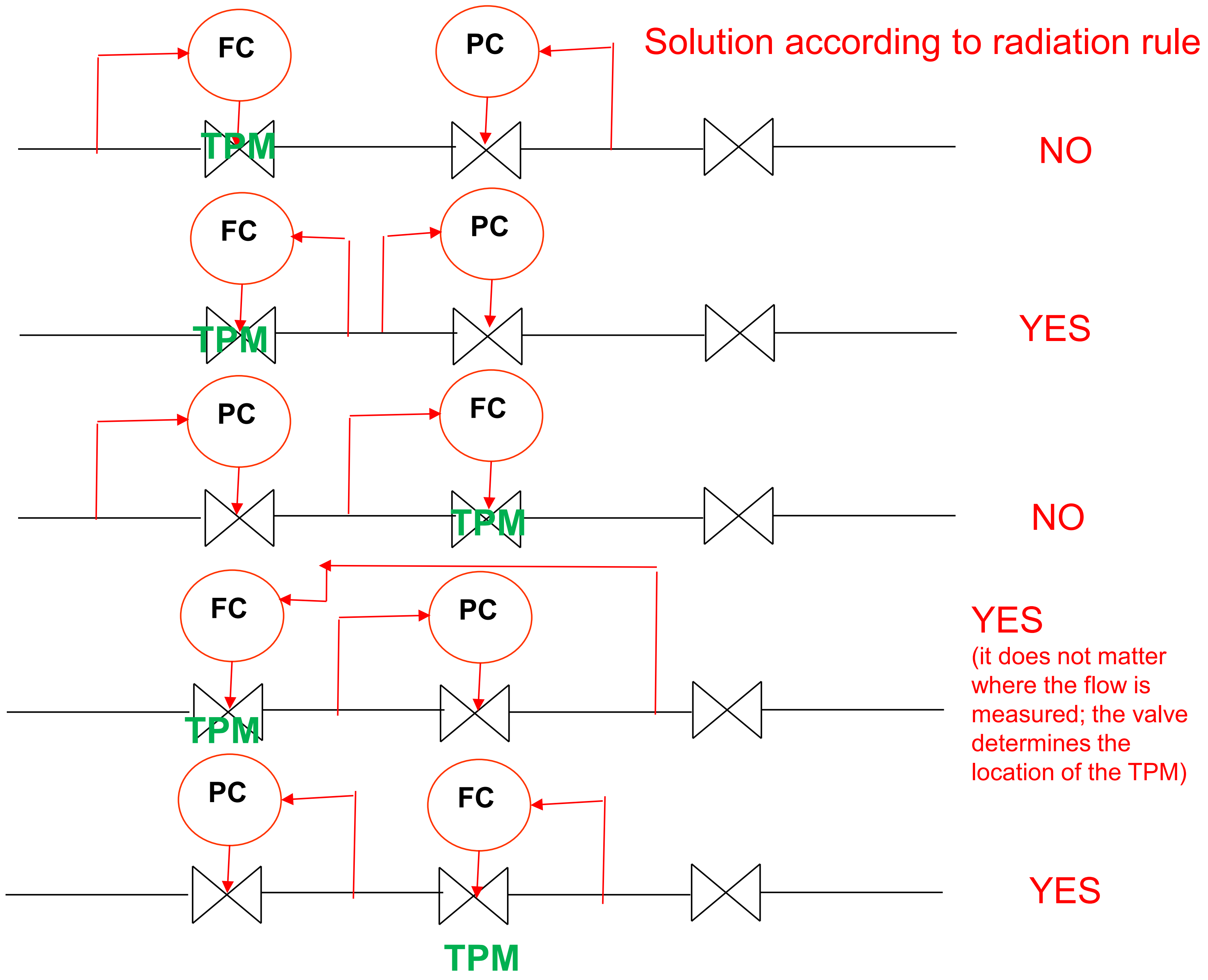
Control smallest inventory!
(b) is best if m_1 is large

Location of sensors

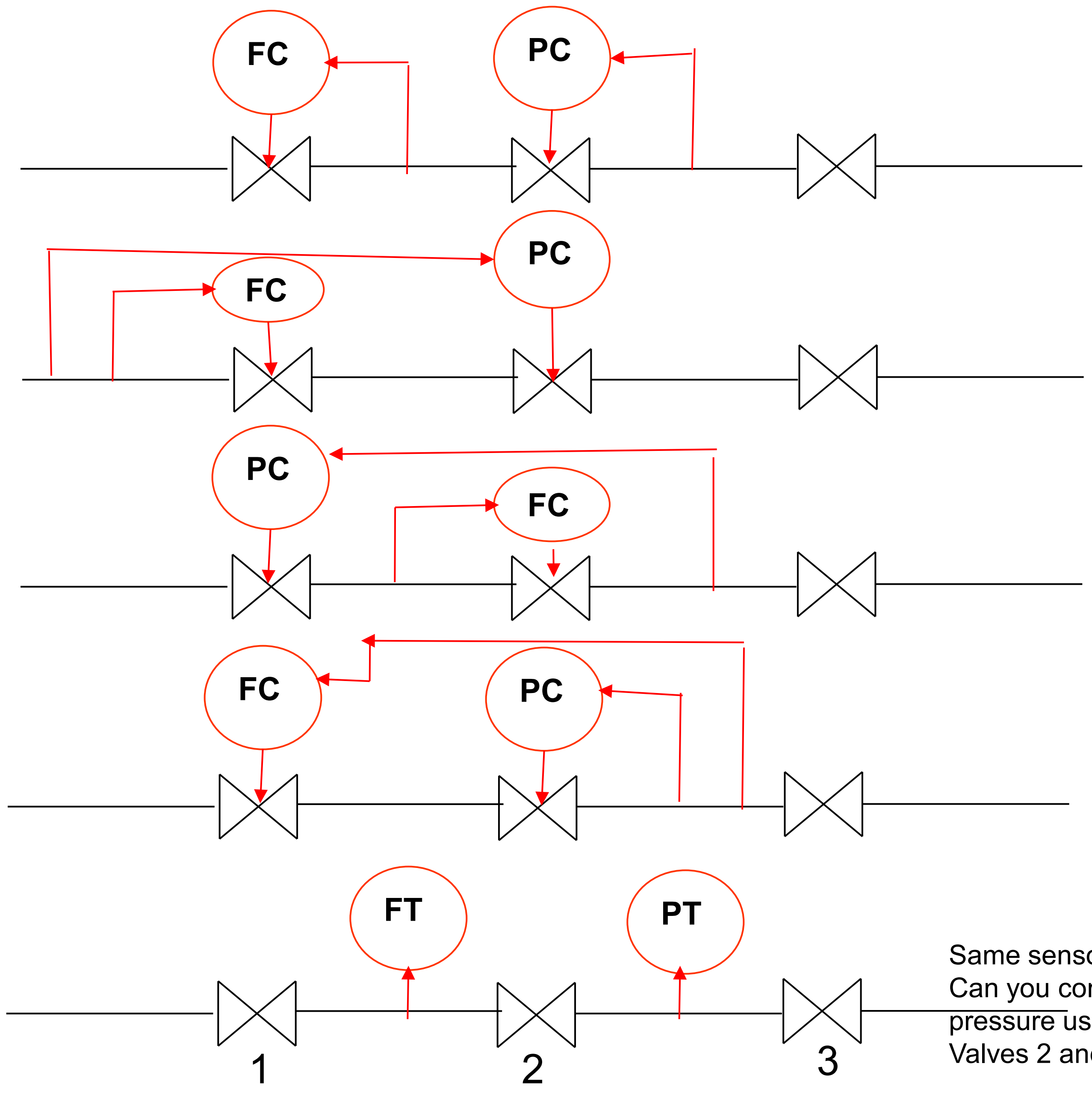
- Location **flow sensor** (before or after valve or pump): Does not matter from consistency point of view
  - Locate to get best flow measurement
    - Before pump: Beware of cavitation
    - After pump: Beware of noisy measurement
- Location of **pressure sensor** (before or after valve, pump or compressor): important from consistency point of view



For each of the five structures; Where is the TPM? Is it consistent?

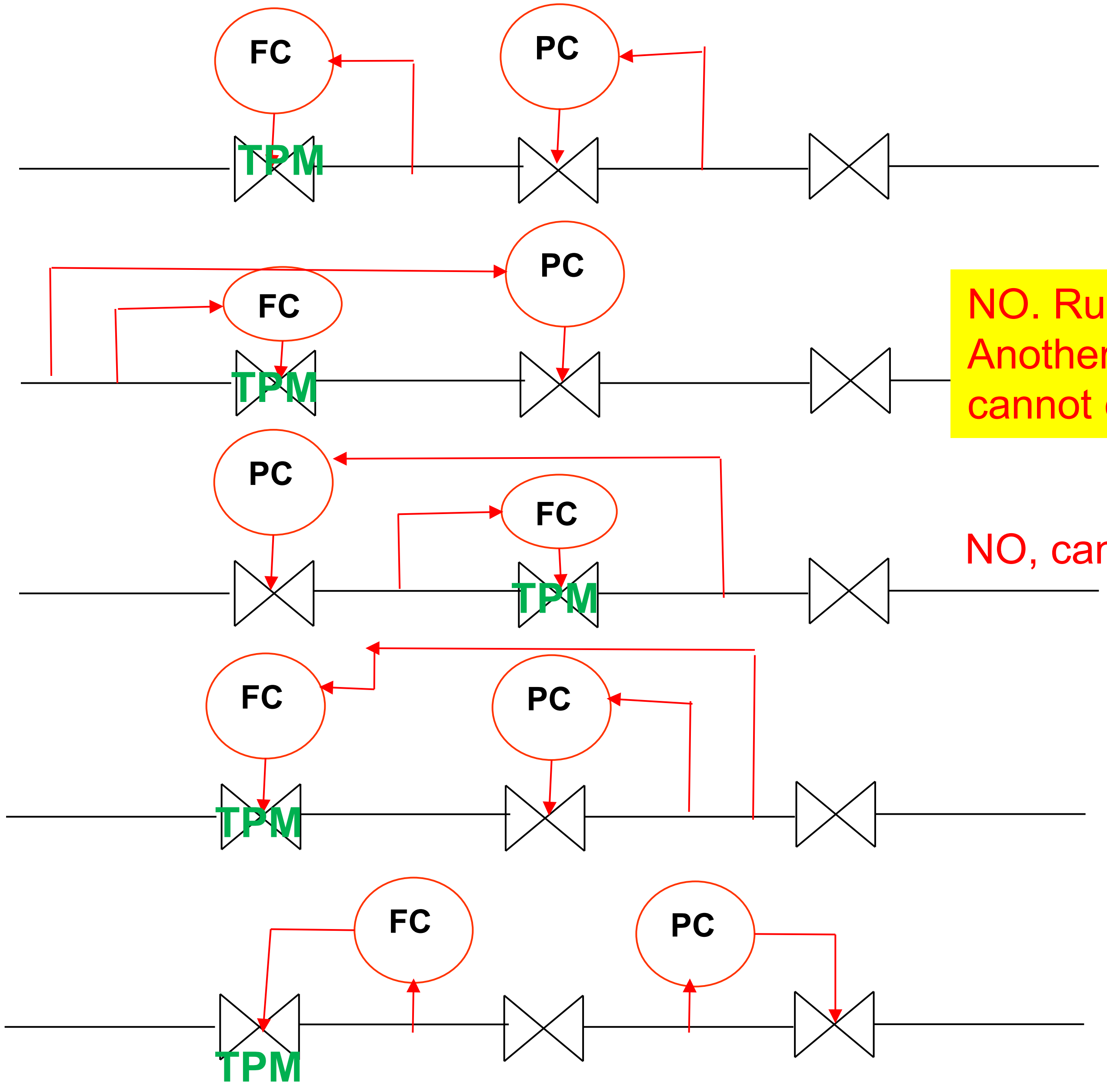


Note: Can never control pressure at ends (upstream first control valve or downstream last valve)!



Same sensor location as case 3. Can you control both flow and pressure using valves 1 and 2, Valves 2 and 3?

Some more. For each of the five structures; Where is the TPM? Is it feasible?



NO

NO. Rule:
Another loop
cannot cross the TPM,

NO, cannot cross TPM

NO

YES. More possibilities,
4 possible, 2 impossible

Some more. For each of the five structures; Where is the TPM? Is it feasible?

Fixed location of TPM. Where should we place it? (Dynamics!)

- TPM = MV used to control throughput
- Traditionally: TPM = Main feed valve (or pump/compressor)
  - Gives inventory control "in direction of flow"

Consider moving TPM if:

1. There is an important CV that could otherwise not be well controlled
   - Special case: Max. production is important: Locate TPM at process bottleneck* !
     - TPM can then be used to achieve tight bottleneck control (= achieve max. production)
     - Economics: Max. production is very favorable in "sellers marked"
2. If placing it at the feed may yield infeasible operation ("overfeeding")
   - If "snowballing" is a problem (accumulation in recycle loop), then consider placing TPM inside recycle loop

BUT: Avoid a variable that may (optimally) saturate as TPM (unless it is at bottleneck)

- Reason: To keep controlling CV=throughput, we would need to reconfigure (move TPM)**

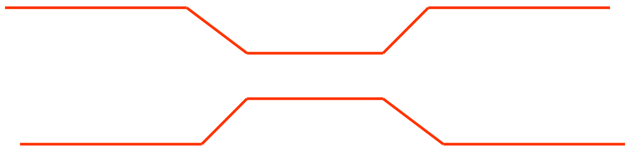
*Bottleneck: Last constraint to become active as we increase throughput -> TPM must be used for bottleneck control

**Input saturation pairing rule (to reduce need for reassigning loops): "Pair MV that may (optimally) saturate with CV that may be given up"

Often optimal: Locate TPM at bottleneck!

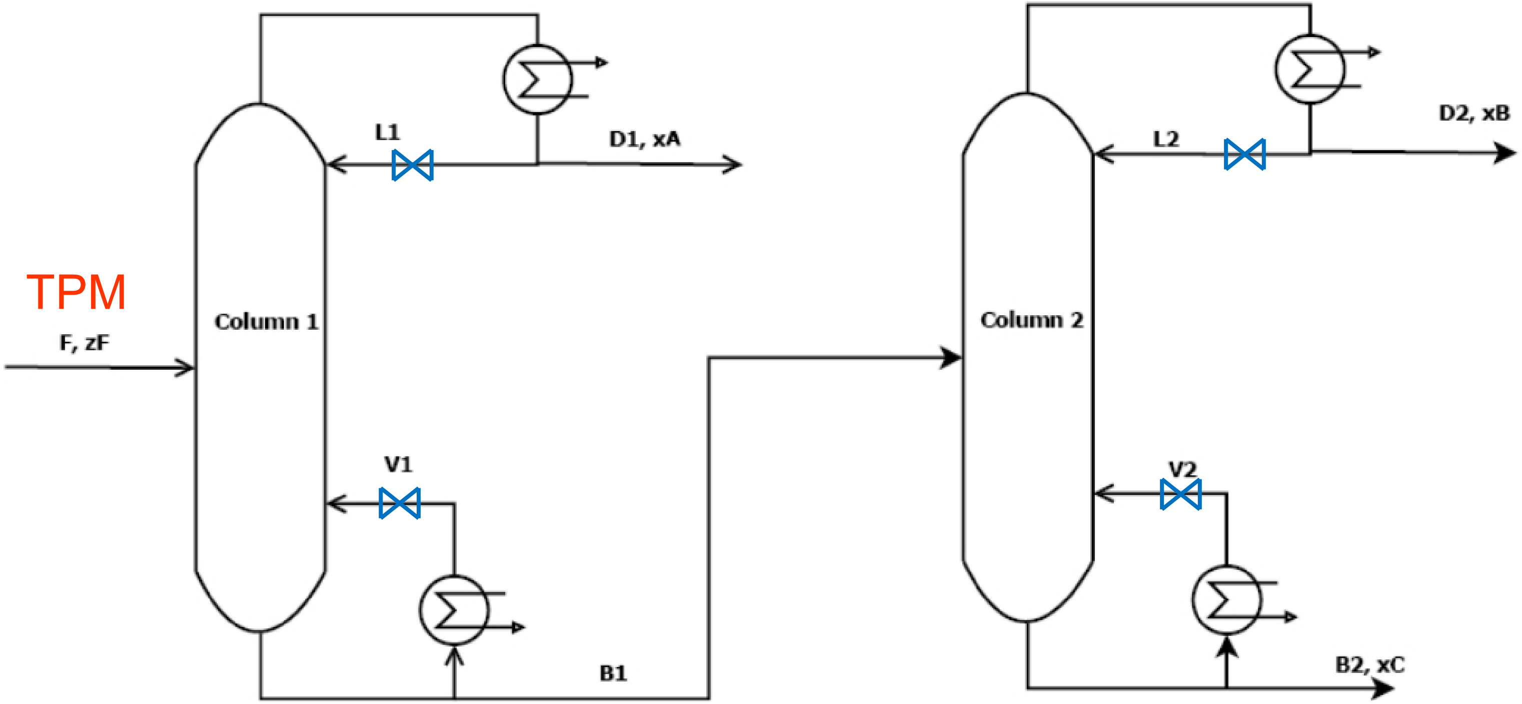
- "A bottleneck is a unit where we reach a constraint which makes further increase in throughput infeasible"
- If feed is cheap and available: locate TPM at bottleneck (dynamic reasons)

- If the flow for some time is not at its maximum through the bottleneck, then this loss can never be recovered.

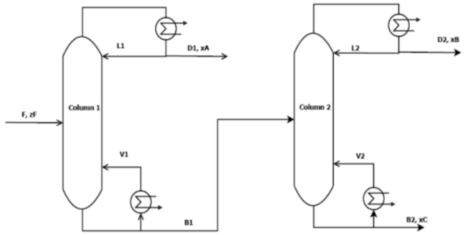


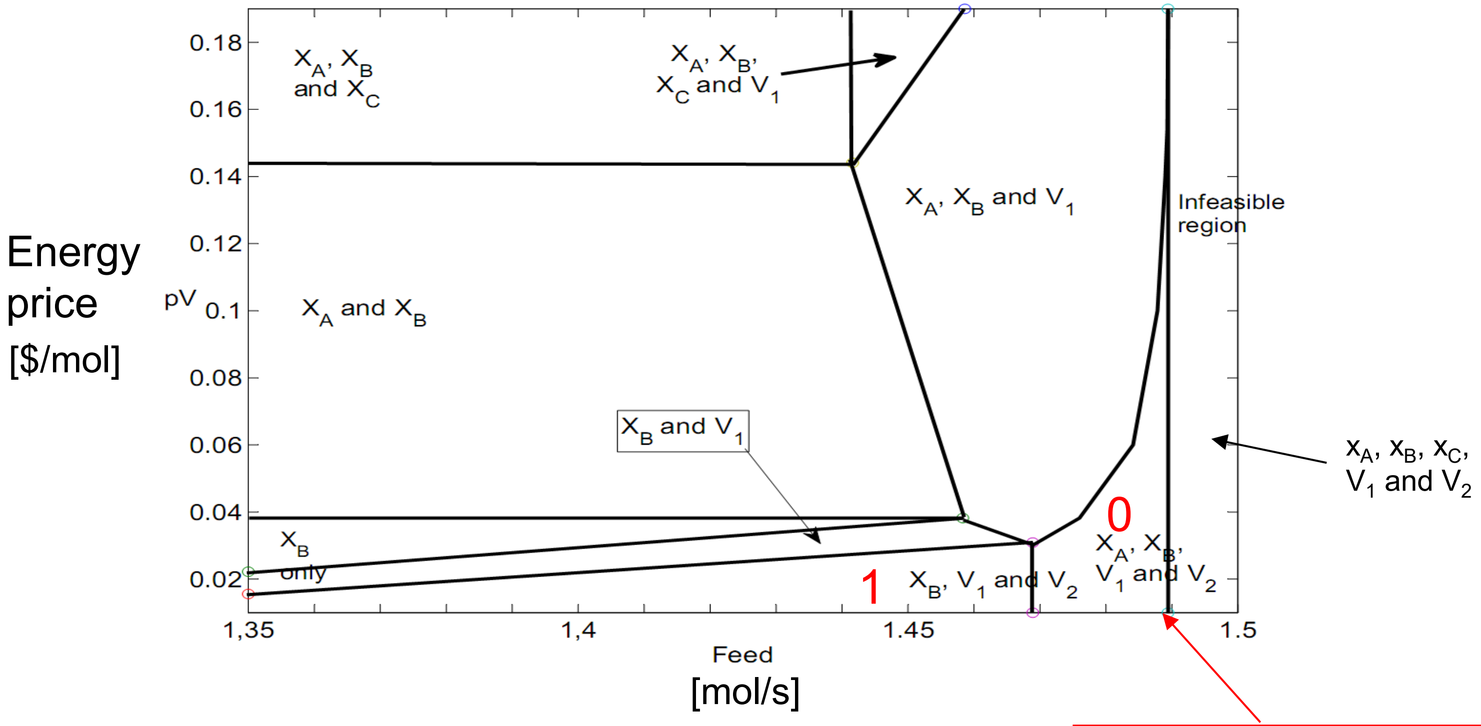
Example: two distillation columns in series

4 steady-state DOFs



Active constraint regions for distillation columns in series



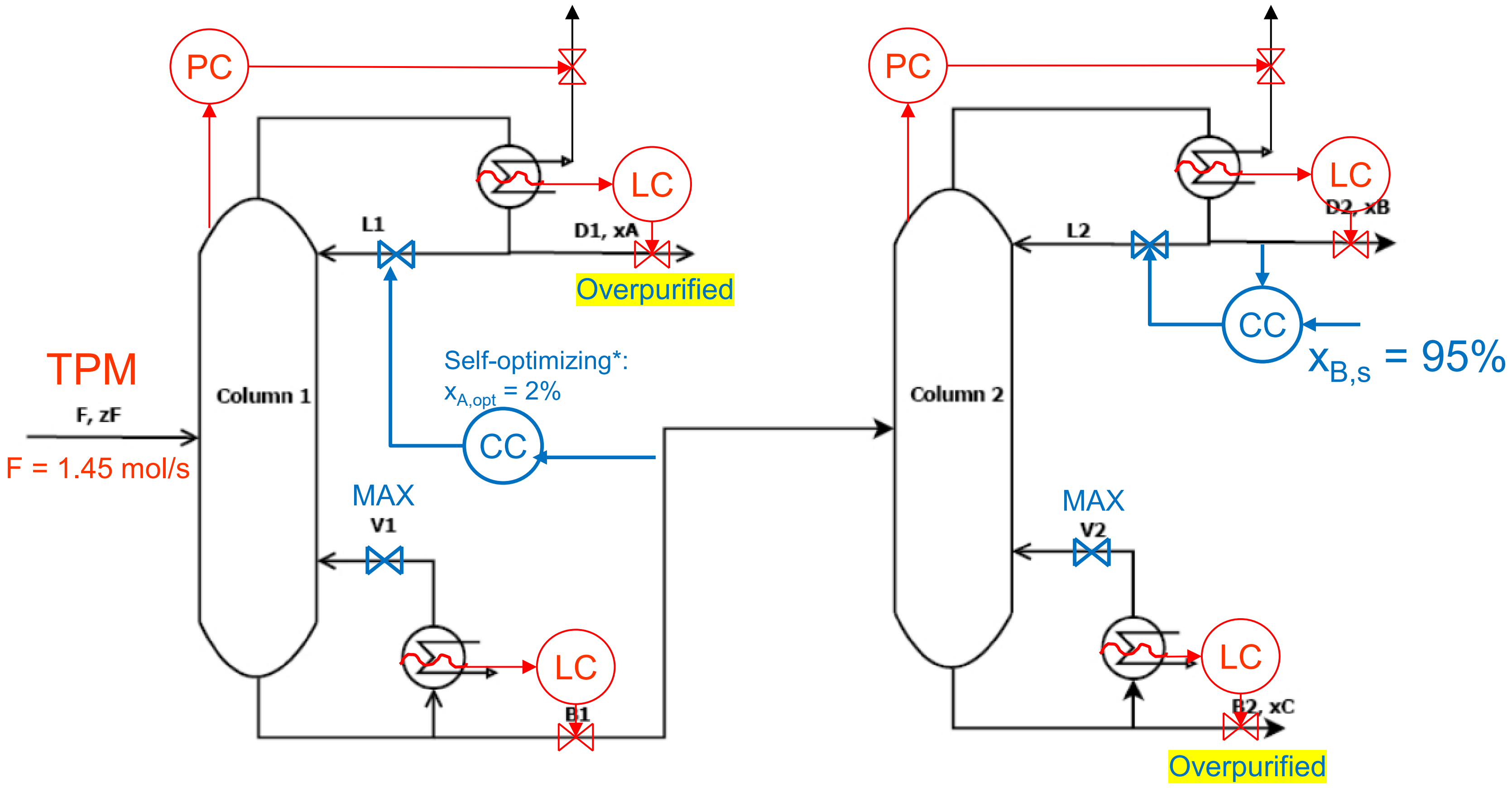


operate at BOTTLENECK

Higher F infeasible because all 5 constraints reached

Control of distillation columns

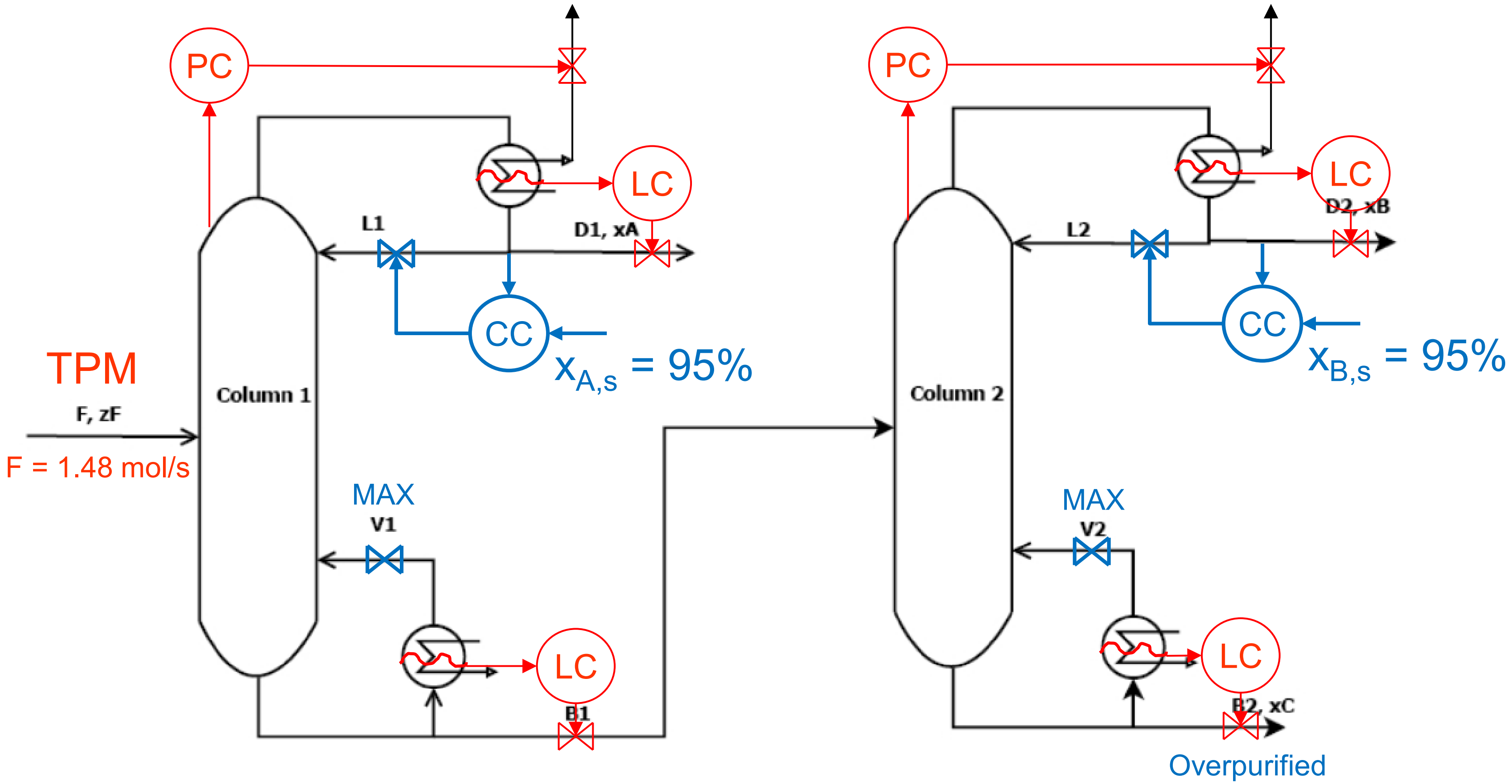
Energy price = 0.02$/mol (low)



Overpurified: To avoid loss of valuable product B

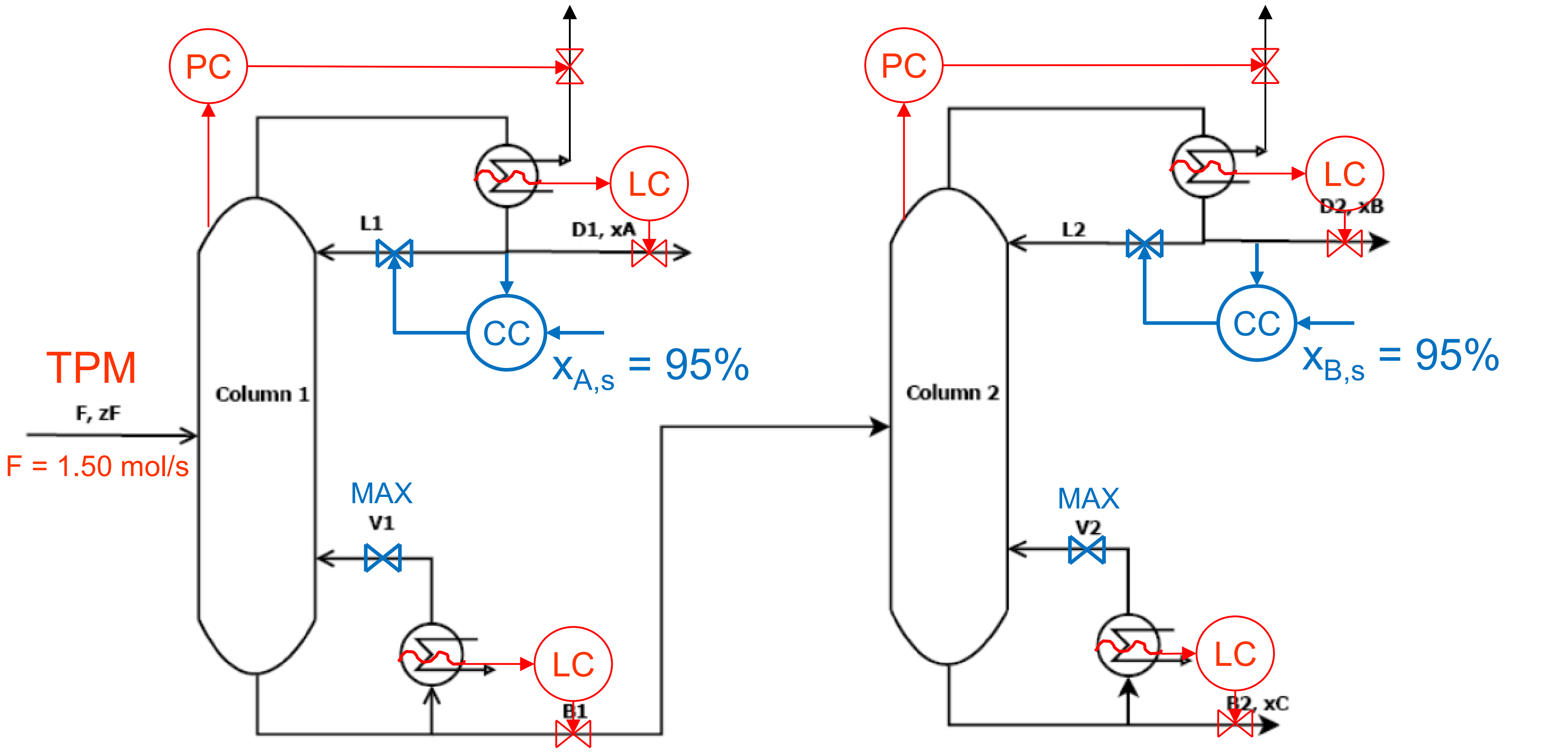
*Setpoint for $X_{A,B1}$ may be set by XA-controller on D2.

Increase feedrate: reach x_A constraint



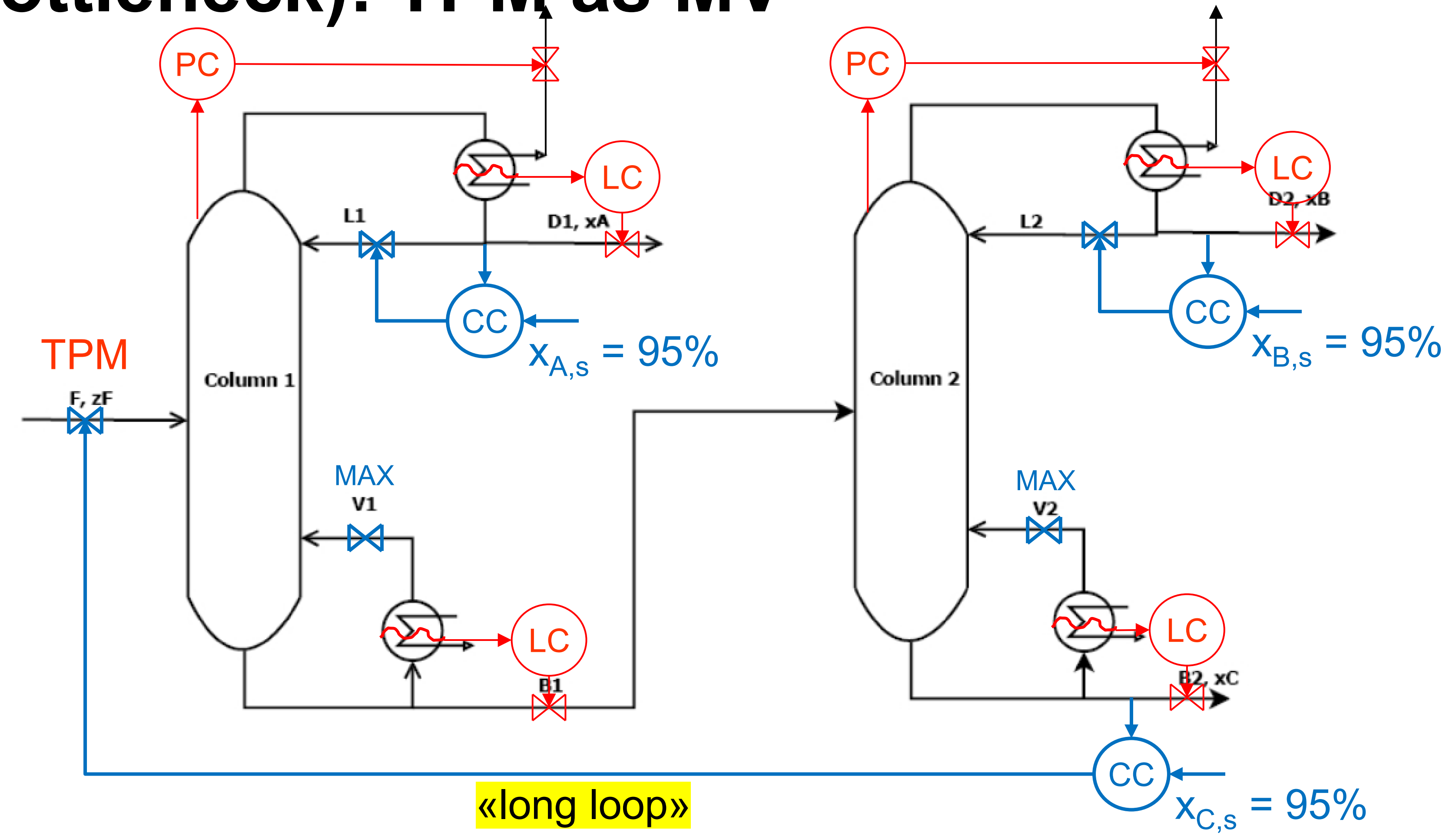
- Implement: Use Max-selector for L1.
- BUT: Moves A over to B2 and gives problems of keeping xB=95% in B2, so will eventually reach constraint on xC (and have reached bottleneck).
- How would you control this?

Increase feedrate further: reach also x_C constraint → infeasible



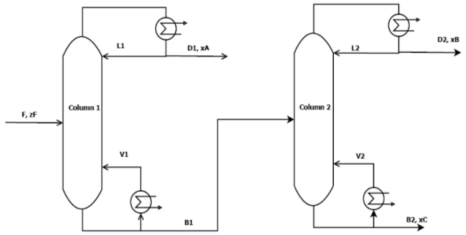
x_C UNCONTROLLED!
Will drop below 95% C if F gets too large

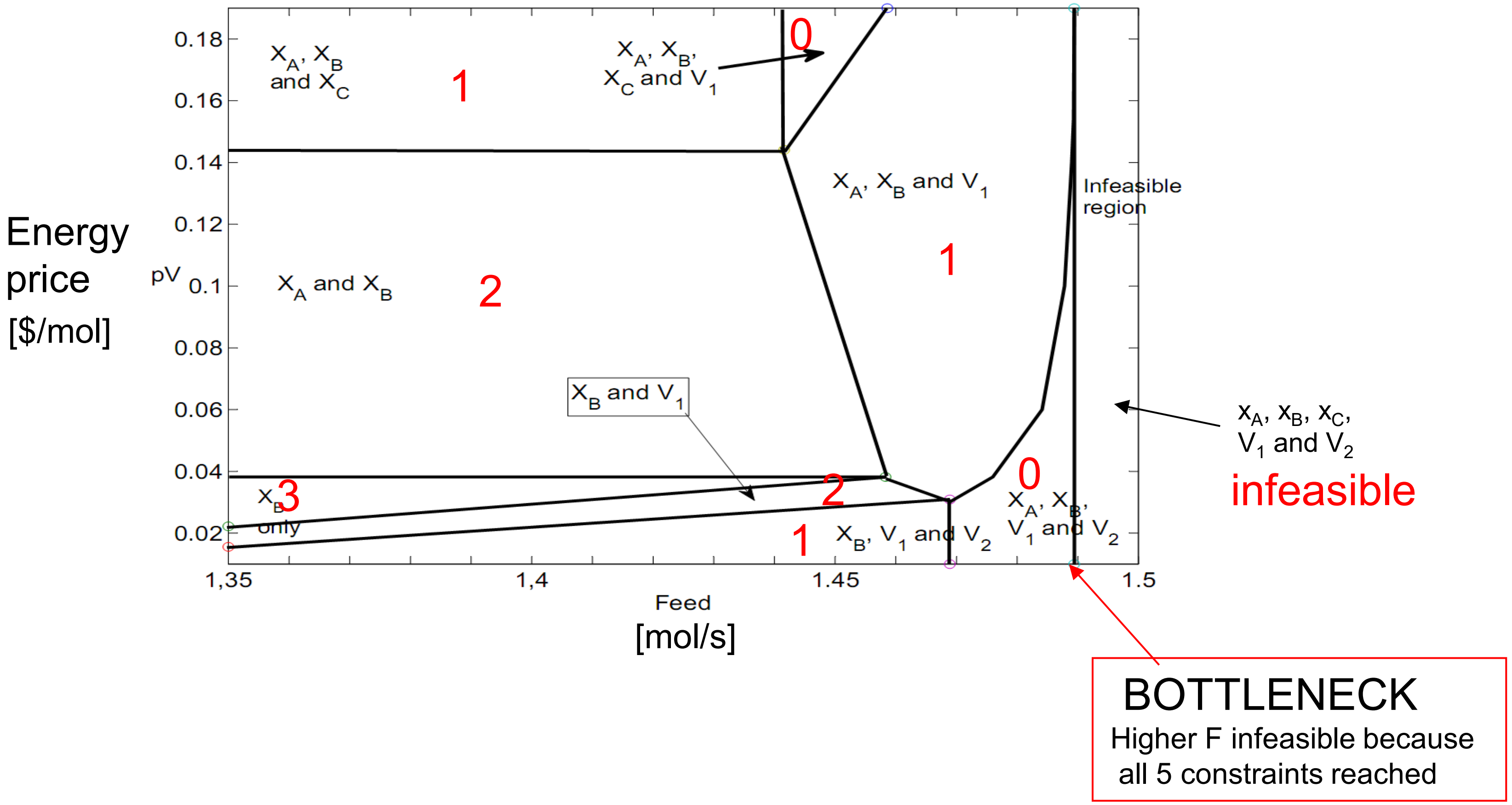
Feasible operation with 5 constraints (process bottleneck): TPM as MV



Implement: Use Min-selector for F
Possible solution to avoid long loop: Move TPM to B1 =feed column 2 (and change LC in column 1)
Comment: This may be undesirable because there is a delay from feed to bottom level (typically 1 min)

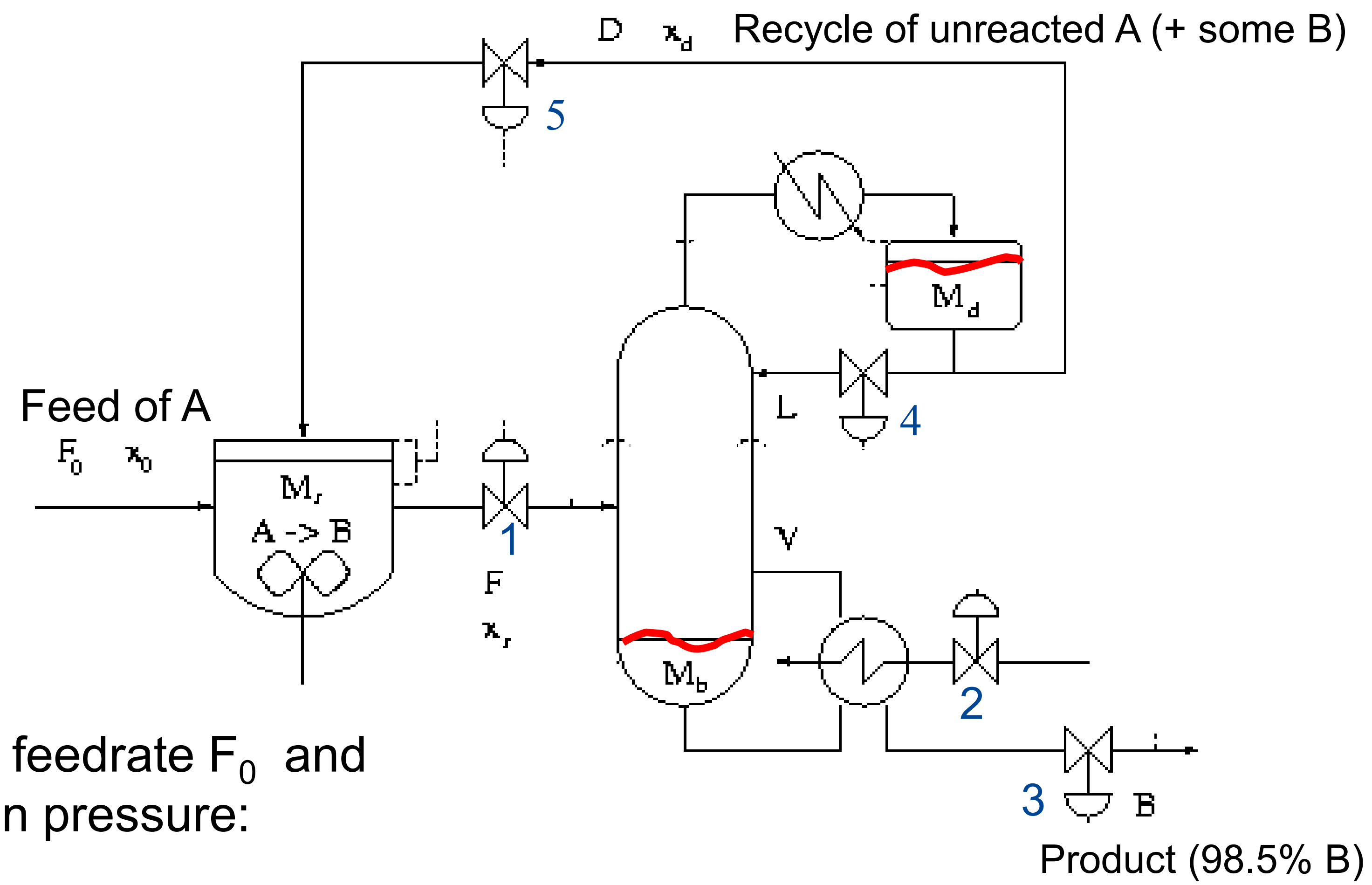
Can we find a simple structure (with selectors) that works in all regions?
No, need variable setpoints for all unconstrained DOFs (and they depend on prices)





Red: Number of unconstrained degrees of freedoim (need self-optimizing CVs for these)

EXAMPLE: Recycle plant (Luyben, Yu, etc.)



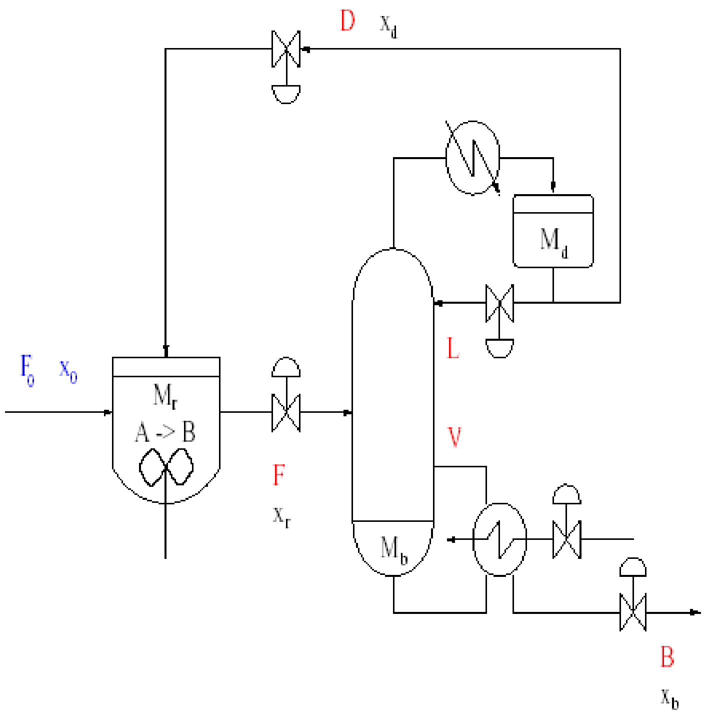
Given feedrate F_0 and column pressure:

Dynamic DOFs: $N_m = 5$

Column levels: $N_{0y} = 2$

Steady-state DOFs: $N_0 = 5 - 2 = 3$

Recycle plant: Optimal operation



Manipulated variables:

$$u^T = [V \; L \; B \; D \; F]$$

Steady-state degrees of freedom: 3

Minimize costs

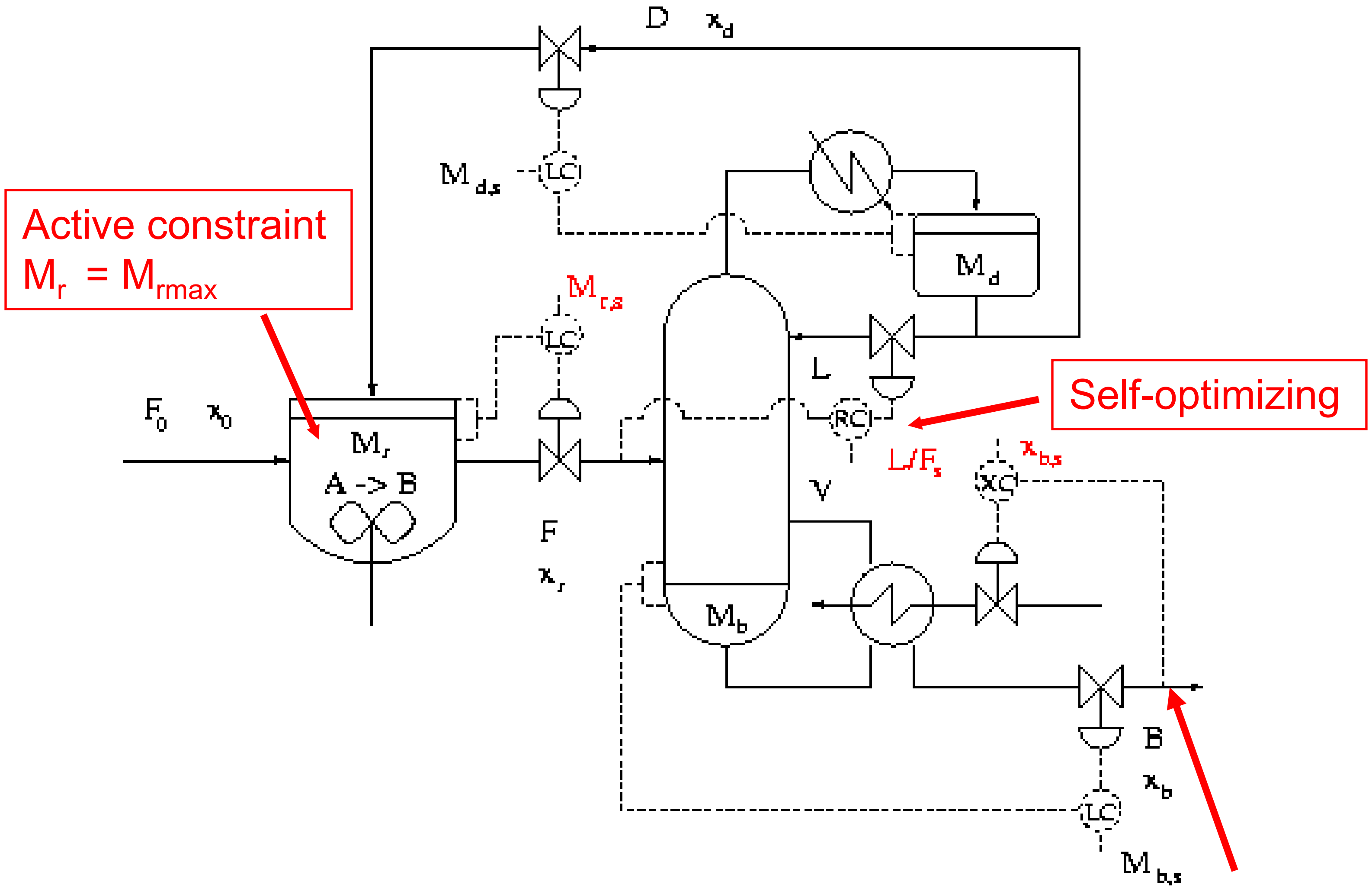
$$J = V$$

Constraints:

x_b	$\leq 0.015\ \frac{\text{molA}}{\text{mol}}$	active
M_r	≤ 2800 kmol	active
V	≤ 5000 kmol/h	
$Flows$	≥ 0 kmol/h	

1 remaining unconstrained degree of freedom

Control of recycle plant



L/F constant: Easier than "two-point" control

Assumption: Minimize energy (V)

Active constraint
$x_B = x_{Bmin}$

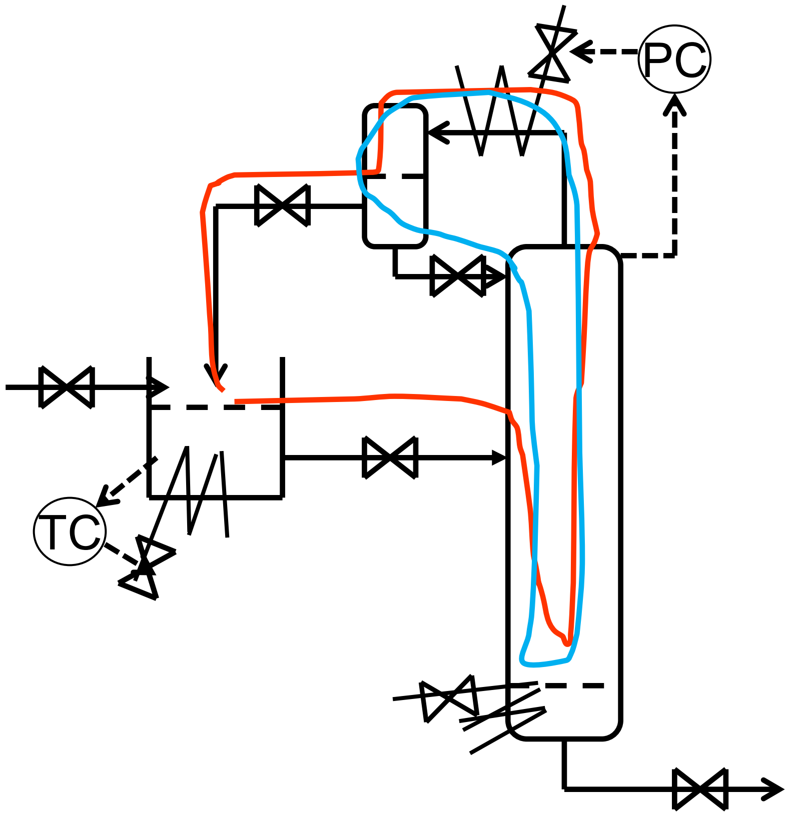
Modified Luyben's law to avoid snowballing

- Luyben law no. 1 ("Plantwide process control", 1998, pp. 57): ***"A stream somewhere in all recycle loops must be flow controlled"***

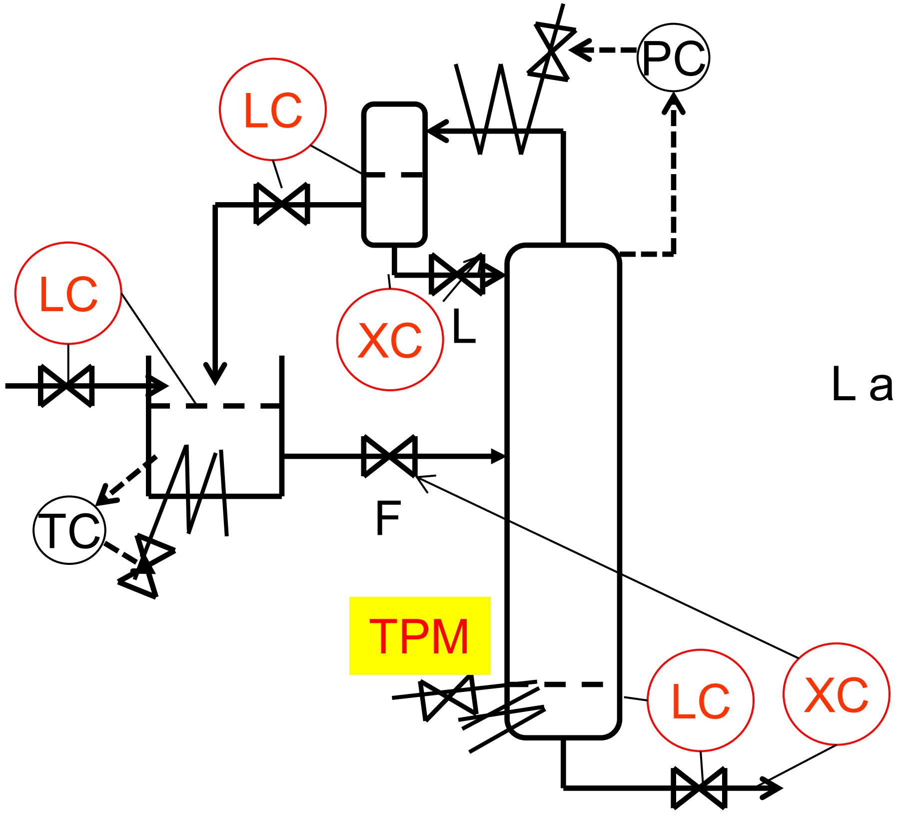
- Luyben rule is OK dynamically (short time scale)…
- BUT economically (steady-state): Recycle should increase with throughput
- Modified Luyben's law 1 (by Sigurd): ***"Consider moving the TPM inside the recycle loop"***

NOTE: There are actually two recycles!

- One through the reactor (D or F)
- One through the column (L)
- One flow inside both recycle loops: V
- Alternative: TPM=V if we want to break both recycle loops!



Changing TPM to V

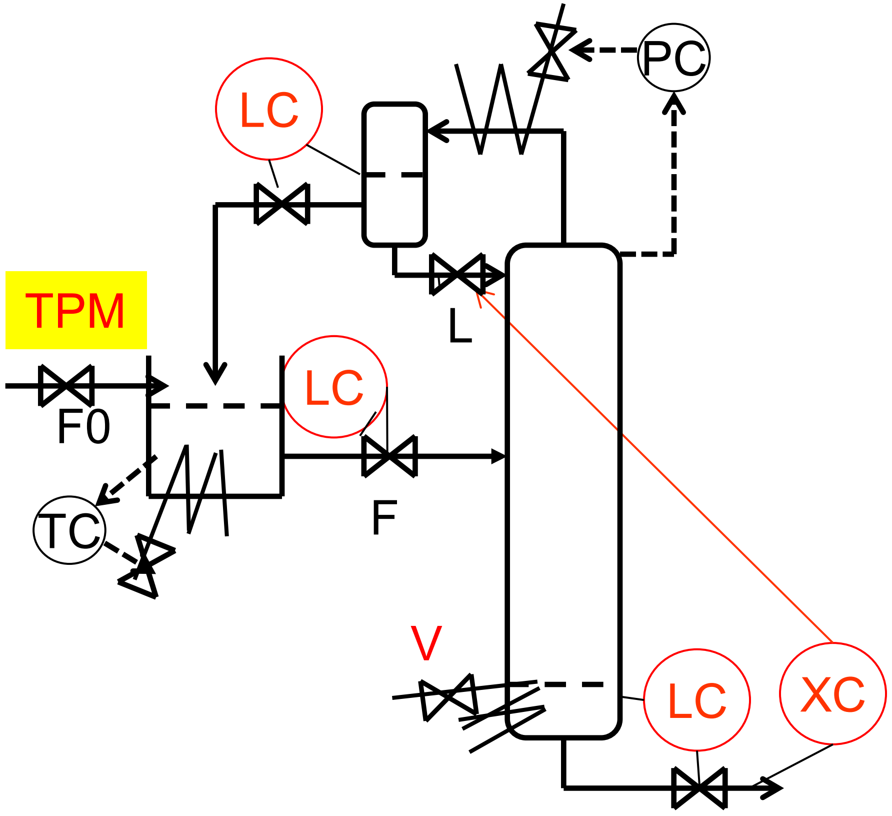


Simulations (to be done) confirm
This is the best!

L and F for composition control: OK!

What about keeping V constant?

(in addition to having another TPM)



NO! Never control cost J=V

Conclusion TPM (production rate manipulator)

- Think carefully about where to place it!
- Difficult to undo after design

Summary: Rules for plantwide control

- Here we present a set of simple rules for economic plantwide control to facilitate a close-to-optimal control structure design in cases where the optimization of the plant model is not possible.

- The rules may be conflicting in some cases and in such cases, human reasoning is strongly advised.

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Simple Rules for Economic Plantwide Control

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https://folk.ntnu.no/skoge/publications/2015/minasidis-pse-escape-2015/

Rule 1: Control the active constraints.

- In general, process optimization is required to determine the active constraints, but in many cases these can be identified based on a good process knowledge and engineering insight. Here is one useful rule:
- Rule 1A: The purity constraint of the valuable product is always active and should be controlled.
- This follows, because we want to maximize the amount of valuable product and avoid product "give away" (Jacobsen and Skogestad, 2011). Thus, we should always control the purity of the valuable product at its specification.
- For "cheap" products we may want to overpurify (purity constraint may not be active) because this may reduce the loss of a more valuable component.
- In other cases, we must rely on our process knowledge and engineering insight. For reactors with simple kinetics, we usually find that, the reaction and conversion rates are maximized by operating at maximum temperature and maximum volume (liquid phase reactor). For gas phase reactor, high pressure may increase the reaction rate, but this must be balanced against the compression costs.

Rule 2: Control "self-optimizing" variables (for remaining unconstrained DOFs).

- This choice is usually not obvious, as there may be several alternatives, so this rule is in itself not very helpful. The ideal self-optimizing variable, at least, if it can be measured accurately, is the gradient of the cost function. Ju, which should be zero for any disturbance. Unfortunately, it is rarely possible to measure this variable directly and the "self-optimizing" variable may be viewed as an estimate of the gradient Ju

The two main properties of a good "self-optimizing" (CV1=c=Hy) variable are:

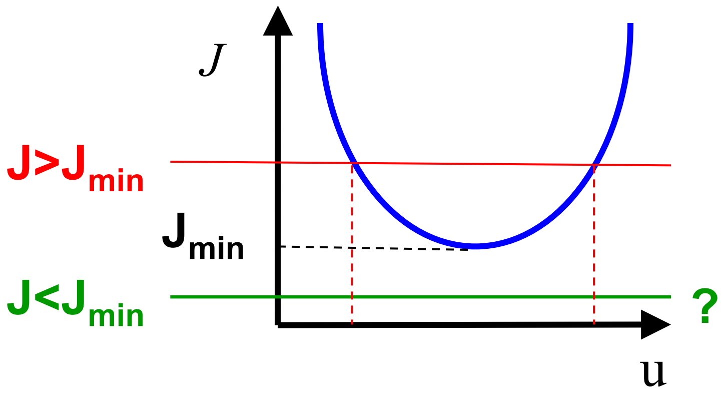
1. Its optimal value is insensitive to disturbances (such that the optimal sensitivity $dc_{opt}/dd = F^c = HF$ = is small)
2. It is sensitive to the plant inputs (so the process gain $dc/du = G = HG^y$ is large).

The following rule shows how to combine the two desired properties:

- Rule 2A: Select the set CV1=c such that the ratio $G^{-1}F^c$ is minimized.
- This rule is often called the "maximum scaled gain rule".

Rule 3: Never try to control the cost J

(or any other variable that reaches a maximum or minimum at the optimum)



- First, the cost function J has no sensitivity to the plant inputs at the optimal point and so G = 0 which violates Rule 2A.
- Second, if we specify J lower than its optimal value, then clearly, the operation will be infeasible
- Also, specifying J higher than its optimal value is problematic, as we have multiplicity of solutions. As mentioned above, rather controlling the cost J, we should control its gradient, Ju.

Rule 4: Locate the TPM close to the process bottleneck

- The justification for this rule is to take advantage of the large economic benefits of maximizing production in times when product prices are high relative to feed and energy costs (Mode 2). To maximize the production rate, one needs to achieve tight control of the active constraints, in particular, of the bottleneck, which is defined as the last constraint to become active when increasing the throughput rate (Jagtap et al., 2013).

Rule 5: Locate the TPM inside the recycle loop, if there is one.

- The point is to avoid “overfeeding” the recycle loop which may easily occur if we operate close to the throughput where “snowballing” in the recycle loop occurs. This is a restatement of Luyben’s rule “Fix a Flow in Every Recycle Loop” (Luyben et al., 1997). From this perspective, snowballing can be thought of as the dynamic consequence of operating close to a bottleneck which is within a recycle system.
- In many cases, the process bottleneck is located inside the recycle loop and Rules 4 and 5 give the same result.

Rule 6: Arrange the inventory control loops around the TPM location according to the radiation rule.

- The radiation rule (Price et al., 1994), says that the inventory loops upstream of the TPM location must be arranged opposite of flow direction. For flow downstream of TPM location it must be arranged in the same direction. This ensures “local consistency”, i.e., all inventories are controlled by their local in or outflows.

Rule 7: Select “sensitive/drifting” variables as controlled variables CV 2 for regulatory control

- This will generally include inventories (levels and pressures), plus certain other drifting (integrating) variables, for example, a reactor temperature or a sensitive temperature in a distillation column. This ensures “stable operation, as seen from an operator’s point of view.
- Some component inventories may also need to be controlled, especially for recycle systems. For example, according to “Down’s drill” one must make sure that all component inventories are “self-regulated” by flows out of the system or by removal by reactions, otherwise their composition may need to be controlled (Luyben, 1999).

Rule 8: Economically important active constraints should be selected as controlled variables CV 2 in the regulatory layer

- Economic variables CV 1 are generally controlled in the supervisory layer. Moving them to the faster regulatory layer may ensure tighter control with a smaller backoff. The backoff is the difference between the actual average value (setpoint) and the optimal value (constraint).

Rule 9: "Pair-close" rule: The pairings should be selected such that, effective delays and loop interactions are minimal.

Rule 10: : Avoid using MVs that may optimally saturate (at steady state) to control CVs in CV 2.

- The reason is that we want to avoid re-configuring the regulatory control layer. To follow this rule, one needs to consider also other regions of operation than the nominal, for example, operating at maximum capacity (Mode 2) where we usually have more active constraints.

Rule 11: MVs that may optimally saturate (at steady state) should be paired with the subset of CV 1 that may be given up.

- This is the **«input saturation pairing rule»**
- This rule applies for cases when we use decentralized control in the supervisory layer and we want to avoid reconfiguration of loops. The rule follows because when a MV optimally saturates, then, there will be one less degree of freedom, so there will be a CV1 which may be given up without any economic loss. The rule should be considered together with rule 10.

- Time scale separation:
  - *A rule of thumb is to have a time scale separation between layers (cascade loops) in the range 4 (minimum) to 10 (preferable).*
- Two most important pairing rules (Rules 9 and 10 from Minasidis et al. 2015)
  - **"Pair close" pairing rule**: The MV should have a large, fast, and direct effect on the CV. In particular, we want a small effective delay (small θ), and we also want a large steady-state gain (large k) and a fast dynamic response (small τ).
    - Recommendation: Avoid pairing on a zero element. Breaking this rule leads to a "long loop", that is, a control loop that only works when other loops are closed
    - Recommendation: Avoid pairing on negative steady-state elements. Breaking this rule will lead to instability in certain cases, for example, if a MV saturates or a loop is put in manual.
  - **"Input saturation" pairing rule**: A MV that may saturate should only be paired with a CV that we can "give up" (stop controlling) when the MV saturates.
- PID tuning rules.
  - τ_c = desired closed-loop time constant [s, min]
  - *PID Rule:* $\tau_c \geq \theta$ = effective time delay for process.
  - Measurement filter F. *Rule:* $\tau_F \leq \tau_c/2$ (preferably much smaller);

https://folk.ntnu.no/skoge/publications/2015/minasidis-pse-...

Some important comments:

1. Each plant generally has one TPM and its location is very important from a dynamic point of view (from a steady-state point of view its location does not matter).
Usually the TPM-varable is NOT included as a degree of freedom when we choose CVs (because we want to be able to adjust the throughput, so it is in essence already a CV).

2. In this paper, we talk about moving the TPM away from the feed, partly for economic reasons (bottleneck control) and partly for dynami reasons (avoid snowballing). However, there may be good reasons to keep the TPM at the feed (not really discussed in this paper).

3. In terms of selecting CVs, this paper requires one to define a cost function (J), which should be minimized
with respect to the degrees of freedom, subject to satisfying the constraints.
Next, we need to find a CV for each degree of freedom (MV).
These will be the active constraints + "self-optimizing" unconstrained variables for the remaining.

4. The "bottle neck" for the process is reached when we reach a constraint where no further increase in the throughput is feasible (feasible means that we satisfy the all constraints).
The TPM is then used to control this bottleneck constraint (in order to have optimal operation, which at this
point is at maximize throyghput).

Note: If this bottleneck constraint is the cost function (which is quite common), then we still need to keep controlling an
unconstrained variable (to keep the bottleneck variable minimized), so it is NOT necessarily true that all MVs are
are to control active constraints at the bottleneck.

Some less important comments

1. If we fix the compressor speed, then the compressor work is no longer a degree of freedom (instead it becomes something we want to minimize).

Plantwide control. Other references

- The following paper summarizes the procedure:
  - S. Skogestad, "Control structure design for complete chemical plants", *Computers and Chemical Engineering*, **28** (1-2), 219-234 (2004).
- There are many approaches to plantwide control as discussed in the following review paper:
  - T. Larsson and S. Skogestad, "Plantwide control: A review and a new design procedure" *Modeling, Identification and Control*, **21**, 209-240 (2000).
- The following paper updates the procedure:
  - S. Skogestad, "Economic plantwide control", Book chapter in V. Kariwala and V.P. Rangaiah (Eds), "*Plant-Wide Control: Recent Developments and Applications*", Wiley (2012).
- More information:

http://www.nt.ntnu.no/users/skoge/plantwide

All papers available at: http://www.nt.ntnu.no/users/skoge/