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Emnemodul: Advanced Process Control

89.5/100

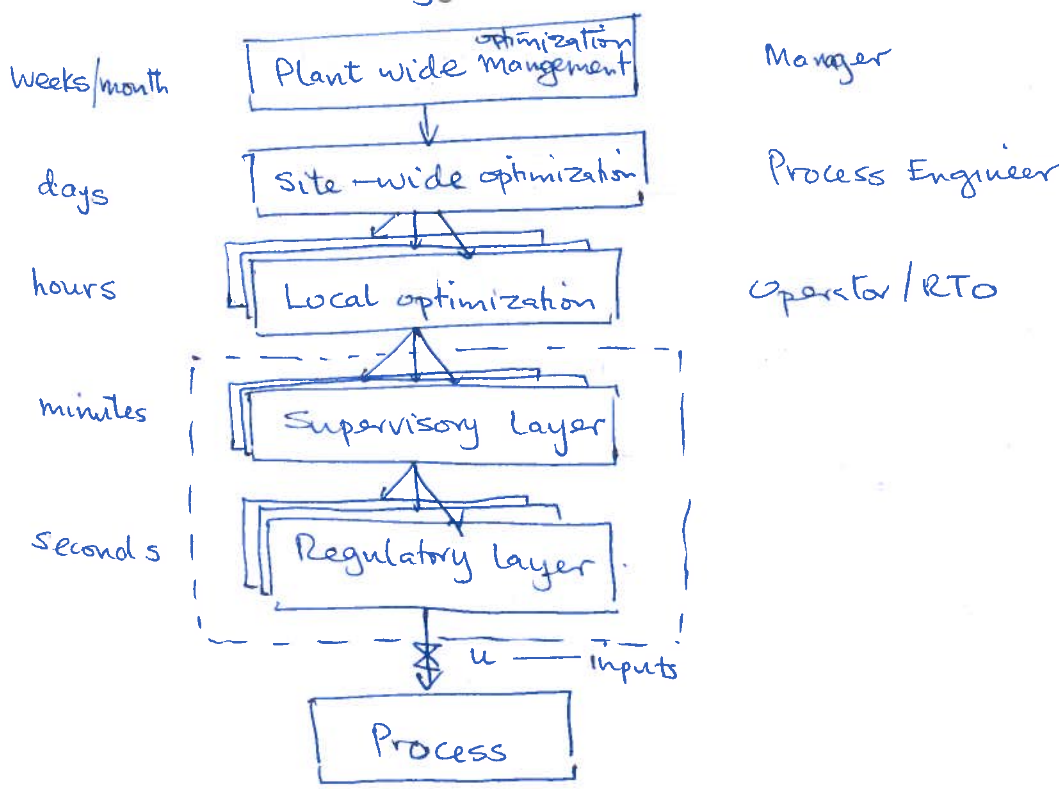
29. Nov. 2018. Time: 0915 – 1200.

Answer as carefully as possible, preferably using the available space. Answer all questions.
If possible, do not write on the backside of the exam.
You may answer in Norwegian; however, English is preferred.

Problem 1 – General Questions (15%)

a) State the hierarchical structure often used in process industries.
b) What are the tasks in each layer?
c) In which section of the control hierarchy would you position self-optimizing control?
d) Is it possible to combine SOC with model predictive control (MPC)? Reason your answer!
e) State the top-down and bottom-up approach to control structure design.
f) What does a typical cost function in a chemical process look like?
*g) What is meant by "squeeze and shift"? Explain the basic idea and its practical significance.

1(a) Process Plant hierarchy



1 (b) 1. Plant Manager (Plant-wide optimization) layer: At this layer planning for process production is done in a time scale of weeks or months / long-term planning for overal process

2. ~~Site~~-wide optimization layer: optimizing the plant operations in the process done on daily basis. Based on steady state models to improve profit and process stability.

3. Local optimization layer: units in a process are locally optimized at each hour by an operator on site or by a real-time optimizer to provide optimal setpoints for supervisory layer. It is a steady-state model based optimization.

4. Supervisory layer: faster than local optimization layer at time scale of minutes. It is slower than regulatory layer. Supervisory layer aims at dynamically optimizing the process ~~for~~ to meets process ~~de~~ constraints. Provides set points for regulatory layer and variables controlled here are called primary variables. Usually an MPC, LQR, or advanced control strategies are placed at this layer

5. Regulatory layer: Aims at stabilizing the process. It consists of simple control loops of P, I and D controllers. It is at a faster time scale of seconds. Provides the optimal inputs for the MVs.

(c) The self-optimizing control is part of Supervisory layer.

(d) No. It does not make sense implementing SOC ~~with~~ below MPC. Because SOC is aimed at controlling a primary control variable that will result to "acceptable losses" from true optimal value ~~and MPC re-op~~ thus avoiding reoptimization while MPC maintains process at optimal point by constantly reoptimizing the process at ~~each~~ a set interval. In brief, they are both aimed at driving the process to optimal operation by measuring primary control variables therefore, they cannot be used together.

1 (e) Top–Down :

Step S1 : Define the operational objective and its constraints.

Step S2 : Identify degrees of freedom and optimize for expected disturbances.

- DoFs = primary control variables
- optimal operating point ? look for active constraints remaining DoFs are self-optimizing .

Step S3 : Implementation of optimal operation ?

- What do we control ? to bring process to optime
  - control active constraints
  - unconstrained degrees of freedom are (SOC)

Step S4 : Set the throughput manipulator .

- Where do we set TPM ? For inventory control.

Bottom – Up :

Step S5 : Regulatory control

- PID control to stabilize process . PID tuning .

Step S6 : Supervisory control

- Apply advanced control configuration depending on the nature of the process locally. Cascade control, MPC, decoupling .

Step S7 : Real-time optimization

Use of RTO if required .

(f) Typical cost function

$$J = -\text{Profit} = \text{Cost feed} - \text{cost energy} + \text{value products}$$

(g) Squeeze and shift means to reduce variance of measurements (squeeze), and "shift" is to move the set-point closer to the hard-constraint ~~due to~~ as a result of the reduced variation.

The significance of this is when you need to meet a hard constraint example direct delivery of product stream to a customer with required composition. ~~wil~~ It avoids the backoff which will eventually saves costs and ~~optimize~~ achieve economical operation .

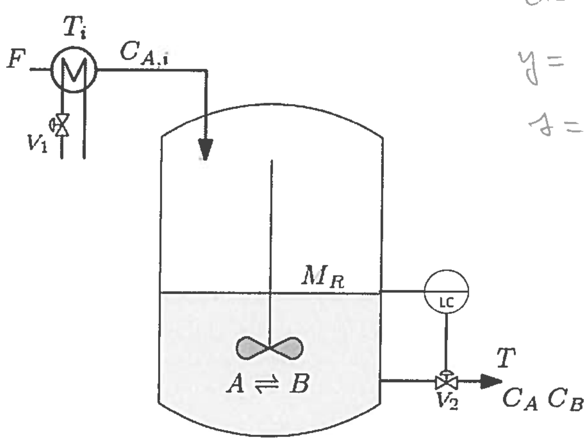
Problem 2 – Self–Optimizing Control (20%)

a) Consider the following exothermic CSTR process with a fixed given inflow rate F. The concentration of component A in the feed stream $C_{A,i}$ is an unknown disturbance. At nominal conditions $C_{A,i} = 1\ mol\ l^{-1}$.The reactor temperature T, concentration of components A and B in the outlet C_A and C_B are available measurements. M_R is the reactor hold-up and T_i is the inlet temperature. The objective is to maximize the profit given by, $J = 2.009\ C_B - (1.657 \times 10^{-3} T_i)^2$, that is, the revenue of the desired product concentration C_B minus a penalty for the heating usage.

$d = C_{A,i}$

$y = T, C_A, C_B$ M_R T_i

$J = 2.009\ C_B - (1.657 \times 10^{-3}\ T_i)^2$



For this process shown in the figure above, what would be the ideal self-optimizing variable that would give zero loss when controlled to a constant setpoint? Reason your answer.

b) The nullspace method is one method, which can be used to select a measurement combination $\mathbf{c} = \mathbf{Hy}$ as the self-optimizing variable. For this method, answer the following questions:

1. How many measurements are required for the nullspace method? Can the null space method be used for the system above?
2. The following table shows the *optimal values* for the nominal case ($C_{A,i} = 1\ mol\ l^{-1}$) and for a perturbed case, where the disturbance $C_{A,i} = 1.3\ mol\ l^{-1}$. Using this information, calculate the optimal sensitivity matrix **F**.

	$C_{A,i} = 1$	$C_{A,i} = 1.3$
$C_A\ [mol\ l^{-1}]$	0.498	0.644
$C_B\ [mol\ l^{-1}]$	0.502	0.656
$C_T\ [K]$	426.761	429.17

3. Calculate the measurement combination matrix **H** using the optimal sensitivity matrix **F**.
4. In which situation is the nullspace method optimal, *i.e.* it has zero loss? Derive an expression showing this optimality.

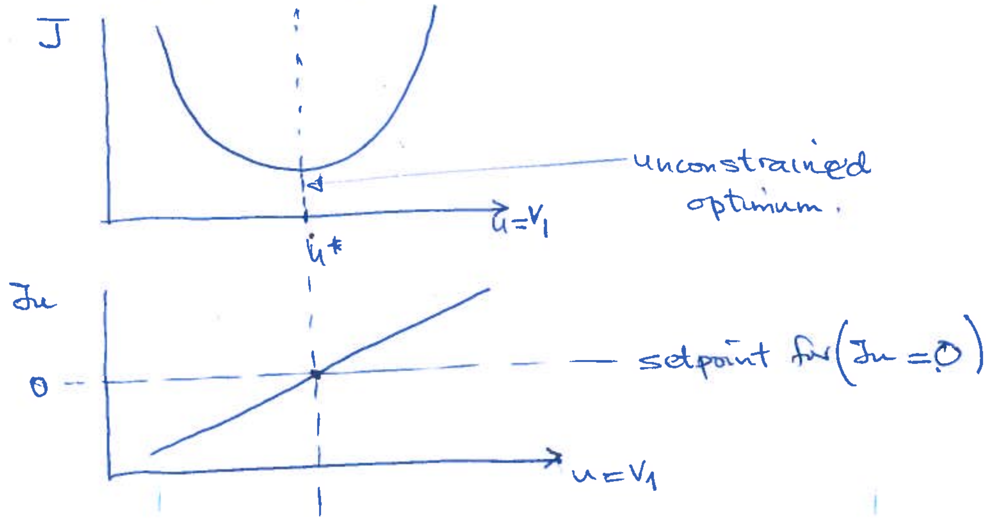
c) The exact local method is a generalization of the nullspace method and a second method to select the self-optimizing variable. What are the advantages of the exact local method compared to the nullspace method?

d) Is it a good idea to control a variable that reaches a maximum or minimum at the optimum? Why?

2 (a) Ideal SOC variable is the gradient (J_u). In this case, $u = V_1$.

So if $\frac{\partial J}{\partial V_1}$ is evaluated in terms of the measurements C_A, C_B and T then it is the ideal self-optimizing variable.

Reason: This is because controlling $J_u = \partial J / \partial V_1 = 0$ will drive the process towards ~~the~~ optimality as specified by KKT conditions of optimality.



~~(b)~~

2 (b) 1. For null space method:

number of measurements $n_y \geq n_u + n_d$

where n_u = number of DOFs (MVs available) and n_d = number of disturbances.

In this case $n_d = 3$ (T_i, $C_{A,i}$ and F are disturbances) $n_u = 1$ (V_1)

$$n_y \geq 3+1 = 4.$$

For a unique solution of H $n_y = 4$.

2. $d^* = C_{Ai}^* = 1 \text{ mol L}^{-1}$ → nominal

$d = C_{Ai} = 1.3 \text{ mol L}^{-1}$ → perturbed.

$$\underline{F} = \frac{\delta y^{opt}}{\delta d} = \frac{\begin{bmatrix} 0.644 \\ 0.656 \\ 429.17 \end{bmatrix} - \begin{bmatrix} 0.498 \\ 0.502 \\ 426.761 \end{bmatrix}}{1.3 - 1} = \begin{bmatrix} 0.4867 & 0.5133 & 8.03 \end{bmatrix}^T$$

3. $$\underline{H}\,\underline{F} = \underline{0}$$

$\underline{H} = \text{null}(\underline{F})$ $\underline{H} = [h_1 \ h_2 \ h_3]$

$$0.4867 h_1 + 0.5133 h_2 + 8.03 h_3 = 0$$

set $h_1 = 1$

$$0.5133 h_2 + 8.03 h_3 = -0.4867$$

set $h_2 = 1$

$$0.5133 + 8.03 h_3 = -0.4867$$

$$h_3 = -0.1245$$

$$\underline{H} = [1 \ \ 1 \ \ -0.1245]$$

Note: $\underline{H}$ has many solutions since $n_y > n_u + n_d$.

2 (b) 4. Null space method is optimal when there is perfect control without controll error and measurement disturbances.

Intuitively, If $\underline{F} = \frac{\delta y^{opt}}{\delta d}$ i.e. ($\underline{n}^y = 0$)

$$\Delta y^{opt} = \underline{F} \Delta d$$

If $\Delta c = \underline{H} \Delta y$

then $\Delta c^{opt} = \underline{H} \Delta y^{opt}$

$$\Delta c^{opt} = \underline{H} \underline{F} \Delta d$$

If $\underline{H}\underline{F} = 0$ then $\Delta c^{opt} = 0$ hence optimal.

(c) Advantages of exact local method over nullspace method

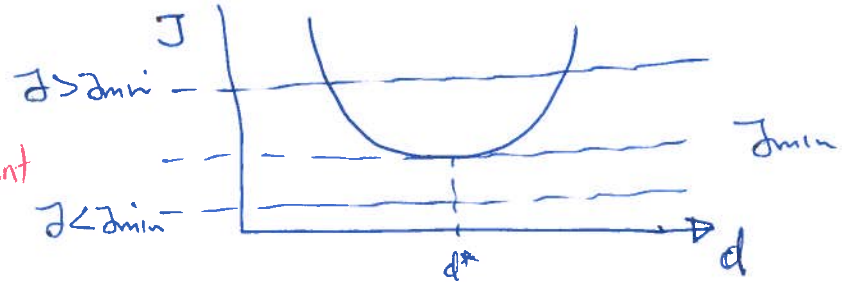
(i) It can be used even if ($n_y \leq n_u + n_d$) the measurements available are fewer than sum of degrees of freedom and disturbances.

(ii) It considers maximum gain rule in finding the best Combination.

(iii) The method includes measurement noise and is suited if noise is significant.

(d) It is not a good idea because when the variable $J > J_{min}$ then it will have two steady-state optimal points but if $J < J_{min}$, there will be no feasiblity.

2/2 ~~0/4~~



~~Variable that reaches constraint (at max/min) is a good CV to control.~~ → Active Constraint Control.

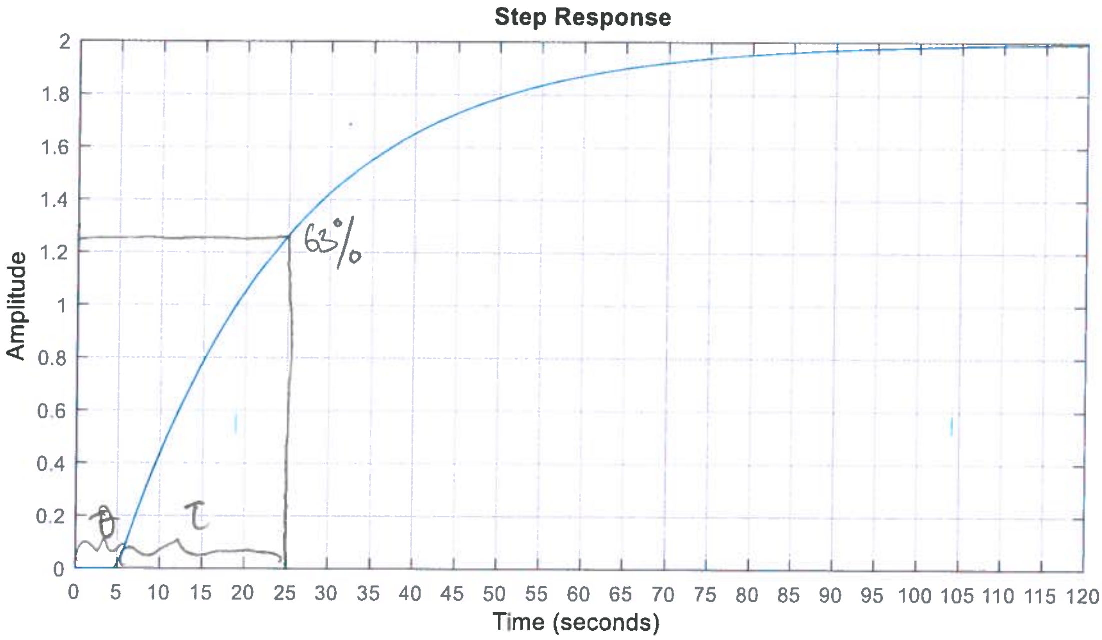
Problem 3 - PID controller tuning (15%)

a) Consider a process given by the following process model

$$G(s) = \frac{2(1-s)e^{-s}}{(12s+1)(3s+1)(0.2s+1)(0.05s+1)}$$

1. Using the half rule, give a first order plus time delay (FOPTD) approximation of the process.
2. Design a PI controller for this process using the SIMC rules with a desired closed loop time constant $\tau_c = 10$.
3. For the same process model $G(s)$, give a second order plus time delay (SOPTD) approximation of the process using the half-rule.
4. Design a PID controller for this process using the SIMC rules with a desired closed loop time constant $\tau_c = 10$.

b) Consider a different process, for which the step response to a unit step change in the input at time t = 0 is given as shown below.



Assuming perfect measurements (i.e. no additional measurement delay), design a PI controller when $\tau_c = \theta$.

3(a) 1. effective delay $\approx 1 + \frac{3}{2} + 0.2 + 0.05 + 1 = 3.75$

inverse response term. considered delay

$\tau \approx 12 + \frac{3}{2} = 13.5$

FOPTD approximation $G(s) \approx \dfrac{2e^{-3.75s}}{13.5s+1}$

2. PI controller. $\tau_c = 10$

$$K_c = \frac{\tau}{K(\tau_c + \theta)} = \frac{13.5}{2(10+3.75)} = 0.491$$

$$\tau_I = \min(\tau, 4(\tau_c+\theta)) = \tau = 13.5$$

3. SOPTD approximation

effective delay $= 1 + \frac{0.2}{2} + 0.05 + 1 = 2.15$

$\tau_1 = 12$

$\tau_2 = 3 + \frac{0.2}{2} = 3.1$

$$G(s) \approx \frac{2e^{-2.15s}}{(12s+1)(3.1s+1)}$$

4. PID controller $\tau_c = 10$

P: $$K_c = \frac{\tau_1}{K(\tau_c+\theta)} = \frac{12}{2(10+2.15)} = 0.4938$$

I: $$\tau_I = \min(\tau_1, 4(\tau_c+\theta)) = \tau_1 = 12$$

D: $$\tau_D = \tau_2 = 3.1s$$

3 (b) From the graph $\theta \approx 5s$

$$\Delta y_\infty = 2$$

$$\Delta u = 1$$

$$k = \frac{\Delta y}{\Delta u} = \frac{2}{1} = 2.$$

$$y(\tau) = 0.63\Delta y + y(0)$$

$$y(\tau) = 1.26$$

$$\tau + \theta \approx 25$$

$$\tau = 25 - 5 = 20s$$

$$G(s) = \frac{2e^{-5s}}{20s+1}$$

PI controller:

$$K_c = \frac{\tau}{K(\tau_c+\theta)} = \frac{20}{2(2\times5)} = 1$$

$$\tau_I = \min(\tau, 4(\tau_c+\theta)) = \min(20, 40) = 20$$

Problem 4 – Consistency (15%)

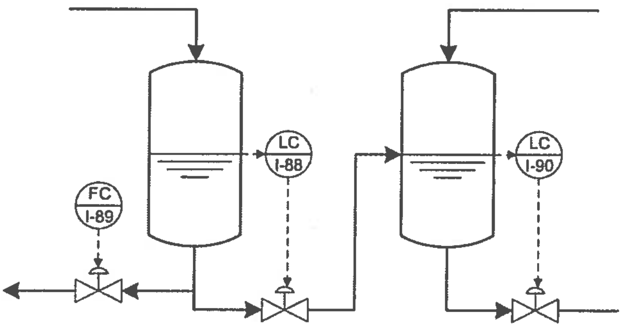
Consistency is a required property for a process in chemical industry. It should be fulfilled in all processes.

a) What do you mean by a consistent control structure? When is a consistent control structure locally consistent?

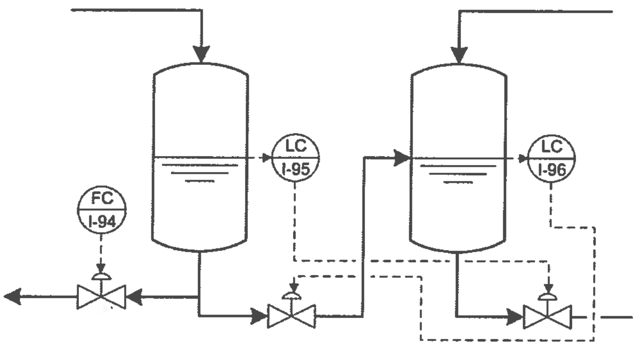
b) What is the “so-called” throughput manipulator (TPM)? Where should one place the throughput manipulator in a system with recycle?

c) Are the following control structure locally and globally consistent and what is (are) the TPM(s)? Justify your answers for global consistency.

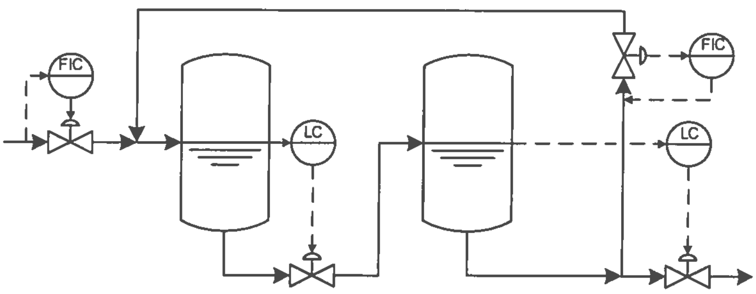
i)



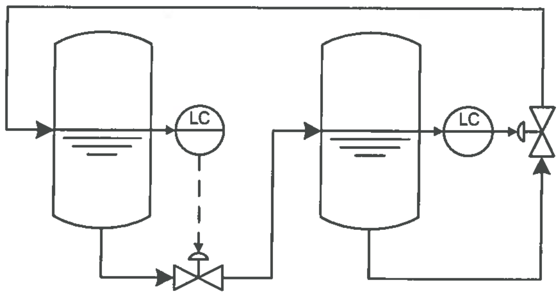
ii)



iii)



iv)



4 (a) A consistent control structure is one whose ~~control~~ inventory control satisfies steady state mass balances (total, composition or phase) ~~at~~ across the whole process and ~~unit~~ individual units.

Local consistency is when the inventory of a unit has its control directly linked to either one of its inflows or outflows or any variable within the unit.

(b) TPM is a degree of freedom that affects the network flow and is not directly or indirectly linked to control of any inventory in the process network. TPM is the manipulated variable that sets the production rate.

For a system with recycle, one should place the TPM in one of the recycle streams.

(c) (i) Yes and yes (both globally and locally consistent). The TPM is the FC and the inlets to the columns. The inventories levels are controlled by LCs that obey radiating rule to satisfy global consistency and local consistency.

(ii) Not locally consistent but globally consistent.

4 (c) (iii) Globally consistent but not locally consistent.

TPM ~~are~~ at both FICs and since one of the ~~fixes~~ set the flow for recycle loop then it causes no inconsistency.

(iv) Not globally or locally consistent. It is not good to control all the tanks in a loop because it will cause cycling.

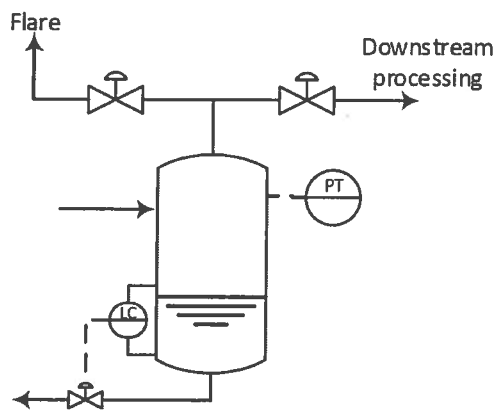
The best way is to leave one tank uncontrolled (usually the largest).

Problem 5 – Advanced Control Structures (15%)

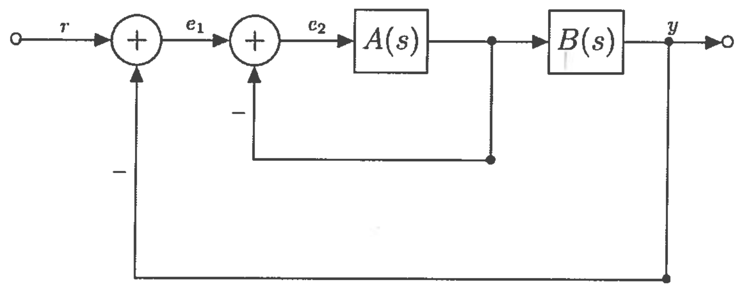
In the process shown below, we need tight control of the pressure in the vessel. Under normal operation, the gas is routed to further downstream processing. For safety reasons, the vessel is also equipped with a flare line, where under extreme conditions, excess gas must be flared to the atmosphere if the pressure in the vessel becomes too high.

a) Since you have 2 MVs and 1CV, propose a control structure to achieve this objective and explain how the proposed control structure achieves this.

Note: Assume that the level in the tank is tightly controlled using the level controller as shown in the figure below. The pressure measurement is also shown in the figure below.



b) The figure below shows the block diagram of a cascade controller:



1. Derive the transfer function $T(s)$ from $r \to y$.

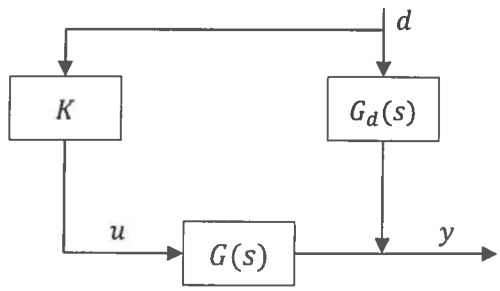
2. Name two properties that represent a good candidate CV for the inner loop in a cascade controller.

c) The relative gain array (RGA) is a tool one can use to decide on controller pairing in multi-variable systems. Additionally, it gives you information about the influence on coupling. Consider the following steady state RGA for a process with 3 inputs u_1, u_2 and u_3 and 3 controlled variables y_1, y_2 and y_3:

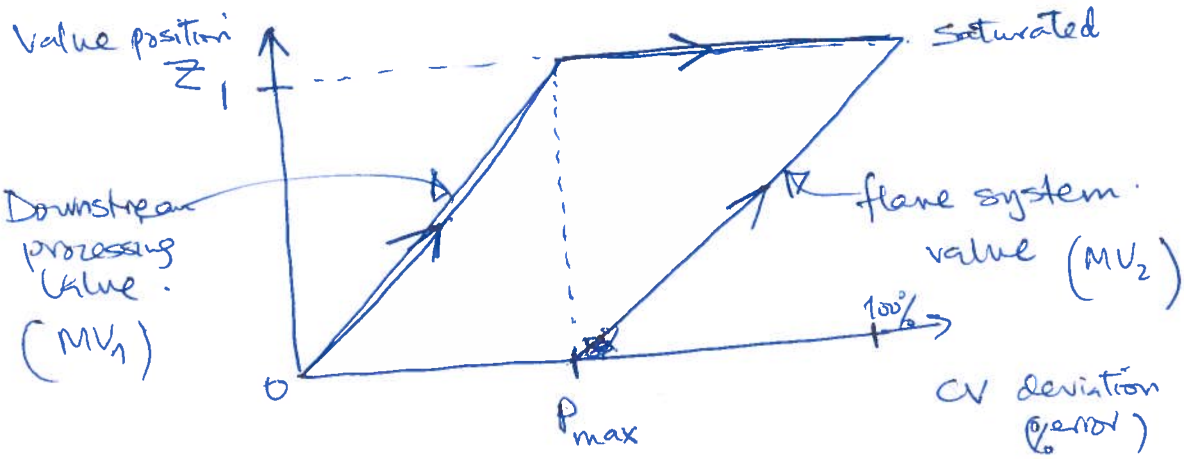
$$RGA = \begin{bmatrix} -0.08 & 1.18 & -0.10 \\ -0.33 & -0.46 & 1.79 \\ 1.41 & 0.28 & -0.69 \end{bmatrix}$$

How would you pair the inputs with the controlled variables?

d) Consider a process $G(s)$, which is subjected to disturbance given by $G_d(s)$, as shown in the block diagram below. If you want to design a feedforward controller to reject effect of distrubances for this system, what would be the ideal feedforward controller K (assuming that it is realizable)?



5 (a) In this case there are 2MVs and CV (maintain safe pressure).
In the case of overpressure, the flare valve must open to release the excess gas. The best control structure is split-range control.
In the case of normal pressure (low) the pressure is controlled by the downstream value. If pressure increases and reaches critical value with the downstream valve saturated the controller starts to opens the flare system valve in order to control the pressure.



(b) 1. T(s) from r → y ?

Inner loop. $T_{inner} = \frac{direct}{1 + loop} = \frac{A}{1+A}$

outer loop $T_{outer} = \frac{T_{inner} B}{1 + T_{inner} B}$

~~T(s) =~~ $T_{inner} B = \frac{AB}{1+A}$

$$T(s) = \frac{\frac{AB}{1+A}}{1 + \frac{AB}{1+A}} = \frac{AB}{1+A+AB} = \frac{AB}{1+A(1+B)}$$

~~2.~~ $\therefore T(s) = \frac{A(s)\,B(s)}{1 + A(s)\,(B(s)+1)}$

2. Inner CV properties:

- It should have faster dynamics. which ~~are~~ may have some non-linearities, small time delay.
- It is subjected to the main disturbances of the process.

(c) $RGA = \begin{bmatrix} -0.08 & 1.18 & -0.10 \\ -0.33 & -0.46 & 1.79 \\ 1.41 & 0.28 & -0.69 \end{bmatrix}$

Rule 1: pair +ve values avoid -ve RGA elements

Rule 2: pick RGA elements closer to 1 as possible

Best pairing:

$y_1 \leftrightarrow u_2$

$y_2 \leftrightarrow u_3$

$y_3 \leftrightarrow u_1$

5 (d)

$$y = G_d(s)\,d + G(s)\,u$$

$$u = Kd$$

$$y = G(s)Kd + G_d(s)d.$$

$$y = \left(G(s)K + G_d(s)\right)d.$$

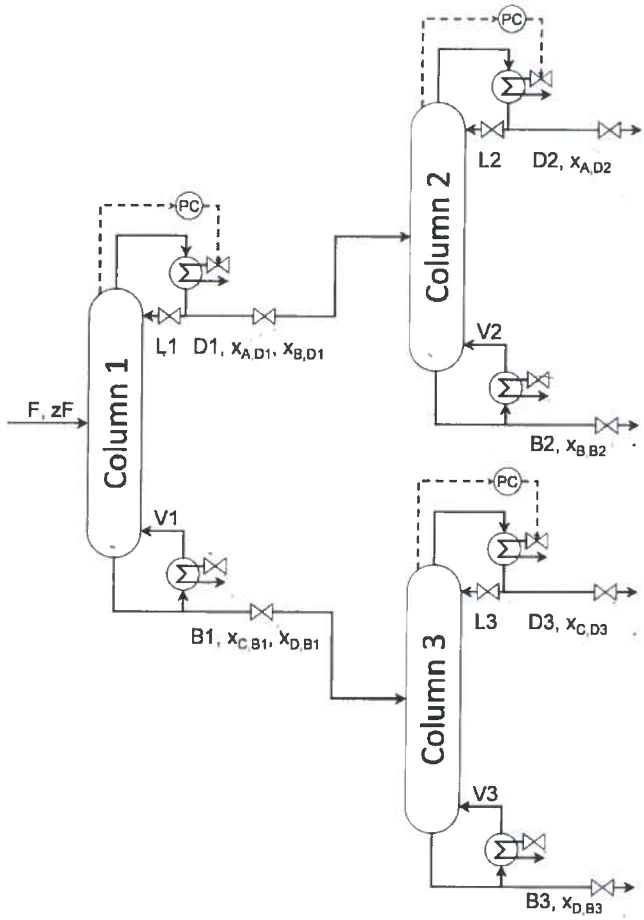
For ideal controller, assume perfect control ie. $y=0$

$$0 = G(s)K + G_d(s)$$

$$\therefore \quad K_{ideal} = -\frac{G_d(s)}{G(s)}$$

Problem 7 (20%)

Consider a sequence of distillation columns for separating four components A, B, C, and D as shown in the figure below.



The feed and its composition are given by the upstream processes and our outside of the analysis. The pressure in each column is controlled by the coolant flowrate (see figure).

Purity constraints are imposed on the four products (distillate and bottom streams of column 2 and 3) as follows

$x_{A,D2} \geq 97\%$ A

$x_{B,B2} \geq 97\%$ B

$x_{C,D3} \geq 97\%$ C

$x_{D,B3} \geq 97\%$ D

To prevent flooding of the columns, the maximum vapor flowrate in the stripping sections of the columns are given by:

$V_1 \leq 5$ mol/s

$V_2 \leq 3$ mol/s

$V_3 \leq 4$ mol/s

Your task is the minimization of the costs given by

$$J = -\text{Profit} = p_F F + p_V(V_1 + V_2 + V_3) - p_{D2}D_2 - p_{B2}B_2 - p_{D3}D_3 - p_{B3}B_3$$

In which the product prices are given by

$p_A = 5\,\$/\text{mol}$

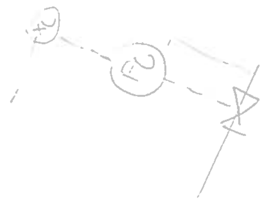
$p_B = 1\,\$/\text{mol}$

$p_C = 5\,\$/\text{mol}$

$p_D = 1\,\$/\text{mol}$

As the process is operated in Iceland with cheap industrial energy prices, the energy price is compared to the general industrial energy price very low and given by:

$p_V = 0.0001\,\$/\text{mol}$



Based on the above information, answer the following questions.

a) How many dynamic and steady-state degree of freedoms does this system have?
b) Based on your experience and engineering know-how, which constraints will be active for the above mentioned system. Justify your answer.
c) Propose a control structure for this case and draw it in the figure on the next page. Explain your choice.
d) If you have degree of freedoms remaining that are not used for controlling active constraints or stabilizing inventories, then propose possibilities on how to use them.
e) What happens, if the feed rate is increased? Would you suggest moving the TPM?
f) The product streams of A and C shall be sold directly to the customers, hence the product composition constraints are hard constraint and may not be violated at any point. Can the idea of squeeze and shift be used? How would you apply it?

6. (a) $N_{dynamic} = N_{valves} = 12$

$N_{0_{ss}}$ (non-steady state effect DoFs) $= 3$

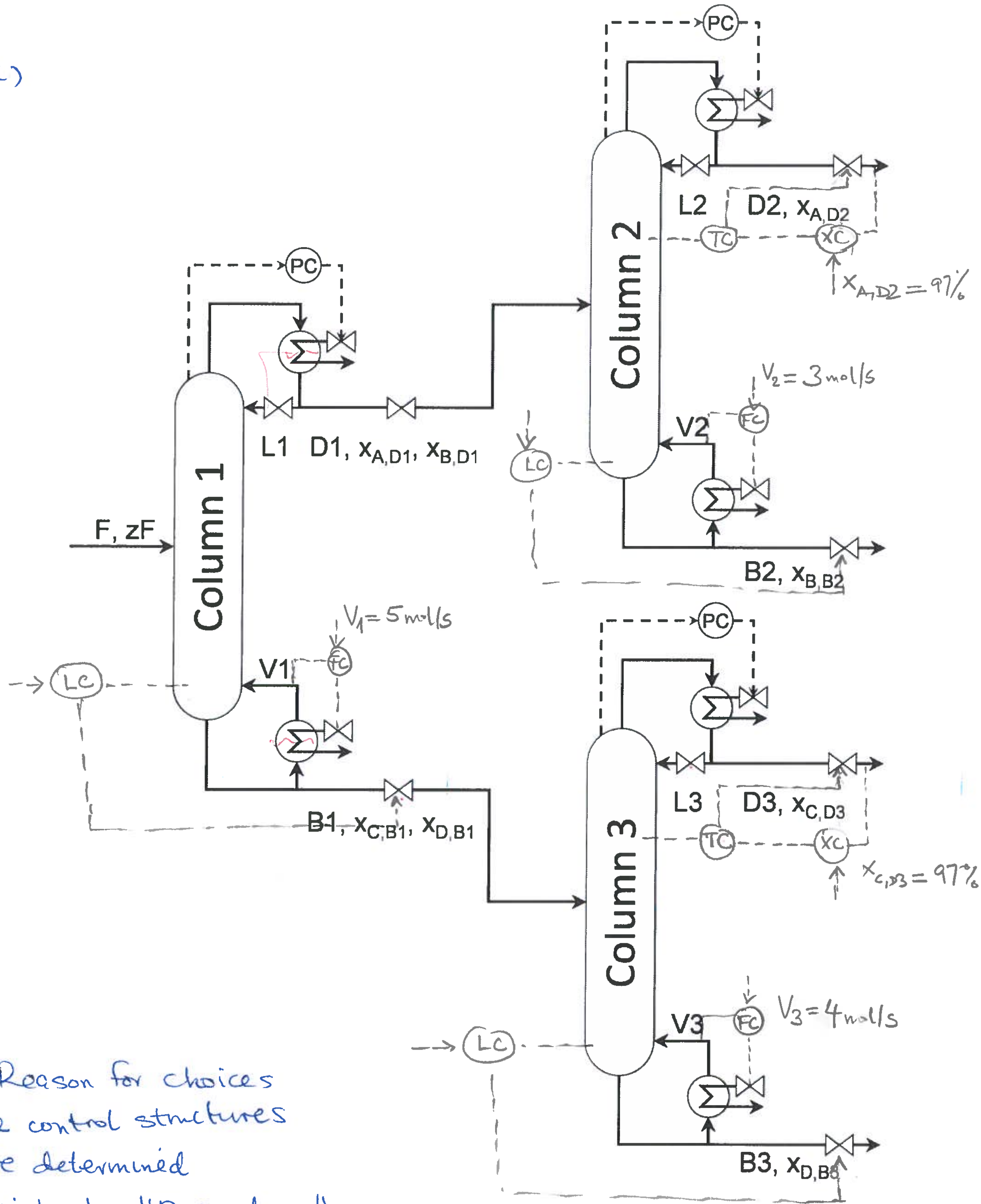
$N_{ss} = N_{valves} - N_{0_{ss}} = 12 - 3 = 8$ steady-state DoFs

(b) Based on Sigurd's rules for primary variable selection.

1. Avoid product give-away of valuable product.
   Most valuable products A, C
   So these constraints for purity will be active.
   $X_{A,D2} \geq 97\%$
   $X_{C,D3} \geq 97\%$
   The other cheap products may be overpurified.
2. For low energy costs the heating is set to maximum.
   $V_1 \leq 5 mol/s$
   $V_2 \leq 3 mol/s$
   $V_3 \leq 4 mol/s$ are both active.

because at very low price the energy costs are insignificant to our objective value.

(e)



c) Reason for choices
The control structures are determined simply by "Pair-close" rule.
Also in distillation, the concentration control is usually in cascade with the temperature controll

6 (d) The extra degrees of freedom remaining can be used for ~~as s~~ control of self-optimizing variables.

~~However usual~~

A combination of measurements can be controlled by the remaining DoFs.

~~(e) If feed~~

(e) If feed rate is increased the composition of the cheaper products also become active, If feed rate goes to maximum then $x_{B,B2}$ and $x_{D,B3}$ become active. These are bottlenecks of the process.

I would suggest moving the TPM to the vapour flow streams V3 and V2.

(f) Yes squeeze and shift can be used because it is a hard constraint.

The dynamic violations of the x_A and x_C is due to the non-linearities present in the process. To break these linearities and reduce variation (squeeze) use cascade control with inner loop taking temperature measurements in one of the trays in the column and the master controller is the composition controller. Then by doing that the set point could be shifted (back-off) reduced ~~to~~ close to the hard-constraint