 Name: DINESH KRISHNAMOORTHY

# Emnemodul: Advanced Process Control

100/100

**30. Nov. 2016. Time: 0915 – 1200.**
**Answer as carefully as possible, preferably using the available space.**
You may answer in Norwegian; however, English is preferred.

20/20

## Problem 1 (20%)

a) What are the steps of Skogestad's plantwide control procedure? Classify them into two different categories.

b) Define self-optimizing control. Is it an alternative or complement to Real-time optimization (RTO)? Is it an alternative or complement to Model Predictive control?

c) The optimal sensitivity matrix $F$ is defined as $F = \frac{dy_{opt}}{dd}$. It can be additionally derived from the linear model for $y$ and the cost function as $F = -G^y J_{uu}^{-1} J_{ud} + G_d^y$. Show how to derive this expression.

d) How is the selection matrix $H$ obtained in the nullspace method?

e) In which case does the nullspace method give zero loss? Derive an expression which shows this.

f) An alternative of the nullspace method is the exact local method. How is selection matrix $H$ obtained in this case?

g) What are the advantages of the exact local method compared to the nullspace method?

h) Explain why using the gradient of the cost function, $J_u$, as self-optimizing control variable is a good idea.

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

2.5

Top down

S1: Identify the objectives and the constraints

S2: Identify the DOF and the active constraints as a function of disturbances (i.e. optimize for disturbances)

S3: Identify the primary CVs. What to control? - preferably self optimizing control

S4: Where to set the production rate? Locate the TPM

Bottom up

S5: Regulatory control. What more to control? Identify secondary CVs and pairing

S6: Supervisory Control

S7: Real time optimization

(e) $J_u(u,d) = J_u(\cancel{u^{opt}}^{0}, d) + J_{uu}\Delta u + J_{ud}\cancel{\Delta d}^{0}$ ?, No

since we ~~only~~ consider change in $u$.

$\Delta c = H\Delta y = HG^y \Delta u + H\cancel{G_d^y \Delta d}^{0}$

$\Delta u = (HG^y)^{-1} H \Delta y$

substituting this in the taylor series expansion

$J_u(u,d) = J_{uu}(HG^y)^{-1} H \Delta y$ | $\Delta y = \cancel{\text{H}} F\Delta d$

$\rightarrow J_{uu}(HG^y)^{-1} HF\Delta d$

since $HF = 0$

then $J_u = 0$

$\therefore$ when $u$ changes, with $HF = 0$, we get $J_u = 0$

(b) Self optimizing control is when we can achieve acceptable loss with constant setpoint for the CVs without needing to ~~se~~ reoptimize when disturbances occur.

~~2.5~~ 1.5

It is generally an alternative to RTO if the right variables are selected. It can be an alternative to model predictive control also since the idea is to move the optimization into the fast control layer and therefore eliminate online optimization. ok.

(However if the plant model mismatch is large, then an optimization layer on top could provide the setpoints for the optimal measurement combinati)

More a complement. ← Hm... This is your research area...?

(c) First order Taylor series expansion around $J_u(u^{opt}, d)$ gives

$$J_u(u,d) = J_u(u^{opt}, d) + J_{uu}\Delta u + J_{ud}\Delta d$$

where $J_u = \frac{\partial J}{\partial u}$, $J_{uu} = \frac{\partial^2 J}{\partial u^2}$, $J_{ud} = \frac{\partial J}{\partial u \partial d}$

2.5

$J_u(u^{opt}, d) = 0$ since it is the optimum, and want $J_u(u,d)$ also to be zero.

$$\therefore \quad 0 = 0 + J_{uu}\Delta u + J_{ud}\Delta d$$

$$\Rightarrow \quad \Delta u = -J_{uu}^{-1} J_{ud}\Delta d$$

Substituting $\Delta u$ in the plant model

We assume a linear plant model

$$\Delta y = G^y \Delta u + G_d^y \Delta d$$

$$\Delta y = \underbrace{\left[-G^y J_{uu}^{-1} J_{ud} + G_d^y\right]}_{F}\Delta d$$

this is of the form $y = F\Delta d$

$$\therefore \quad F = -G^y J_{uu}^{-1} J_{ud} + G_d^y$$

Q.E.D

(d) $HF = 0$. This is because we want $\Delta c^{opt}(d) = 0$

2.5

$$\Rightarrow H\Delta y^{opt}(d) = 0 \Rightarrow HF\Delta d = 0$$

$$\therefore \quad HF = 0$$

(e) no measurement noise and $n_y \geq n_u + n_d$ ✓

$\Delta c^{opt}(d) = 0$ (we want this)

2.5 $H \Delta y^{opt} = HF\Delta d = 0 \Rightarrow HF = 0$ (see derivation proof on previous page)

~~size of $z = n_y = n_d$~~

f) $H = \arg\min_H \frac{1}{2} \| J_{uu}^{1/2} (HG^y)^{-1} H [F W_d \; W_n] \|_2$

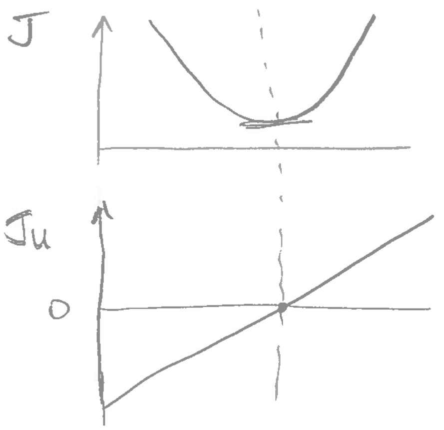
2.5 sol$^n$ to which is given by : $H = G^{yT} (\tilde{F}\tilde{F}^T)^{-1}$

where $\tilde{F} = [F W_d \; W_n]$

(g) Exact local method works in the presence of measurement noise/implementation error. ✓

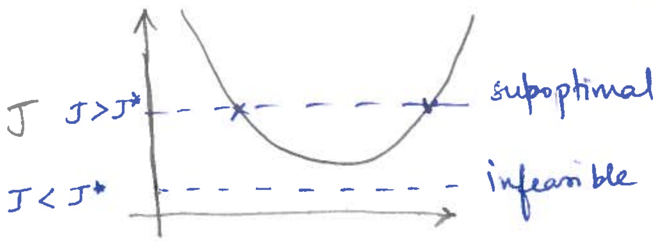
1.5 + can have any $n_y$.

(h) $J_u = 0$ is a stationary point which forms the necessary condition of optimality. This is represented pictorially




by keeping $J_u = 0$, we are always at the minimum of $J$

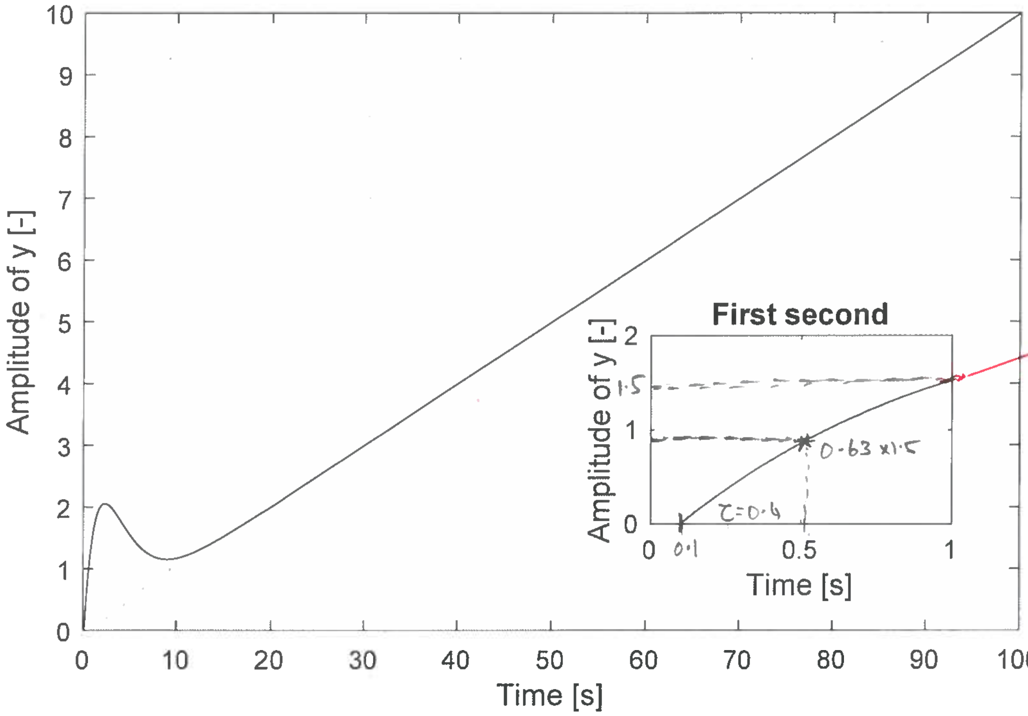
(i) Never control the cost directly. This will lead to either infeasibility ✓ or suboptimality (where there are two ss points and would lead to some strange operation)



4/10

**Problem 2 (10%)**

Consider the following step response from $u$ to $y$. The step change in $u$ is given by $\Delta u = 1$.




You want to use $u$ for the control of $y$.

We consider two cases

i) The measurement of $y$ is perfect.

ii) The measurement of $y$ has an additional delay of 20s.

$e^{-\theta s} = \frac{1}{(\theta s + 1)}$

$e^{-\theta s} \approx \frac{1}{\theta s + 1}$ $\quad \frac{1}{20s+1} \approx e^{-20s}$

a) You want to achieve tight control of the process and hence consider $\tau_c = \theta_{total}$. For each of the two cases, fit a first order + time delay process (or integrator + time-delay process) to this process and apply the SIMC rules.

b) Extra points: What do you think the process model is (transfer function g(s))?

(i) $\theta_{total}$ = any time delay + $\frac{1}{2}$ ~~big~~ first neglected time constant ~~+~~ higher order time constant + meas. delay

$\theta_{total} \approx 0.1$ ✓ and we are interested in the first second of $y$.

$\tau \approx 0.4 = 0.63 \times 1.5$

$G'(s) = \frac{1.5\, e^{-0.1s}}{(0.4s+1)}$ $\quad$ simc rules give

4

$K_c = \frac{1}{k}\frac{\tau}{(\tau_c+\theta)} = \frac{1}{1.5}\frac{0.4}{(0.1+0.1)} \approx 1.334$ ✓

$\tau_c = 0.1$

?,?,? No way it is this small.

$\tau_I = \min(\tau, 4(\tau_c+\theta)) \Rightarrow 0.4,\ 4(0.2)$ ~~≠~~ $\underline{0.4}$ (✓)

~~When it has Time delay = 20s, we are interested~~ in
~~the longer time s~~

But if we assume the first second is an integrating process,
we then get

$$k' = \frac{\Delta y}{\Delta u \cdot \Delta t} = \frac{1.5}{1} \Rightarrow 1.5 \qquad G'(s) = \frac{1.5\, e^{-0.1s}}{s}$$

$$K_c = \frac{1}{1.5} \frac{1}{(\tau_c + \theta)} = 3.334$$

$$\tau_I = 4(\tau_c + \theta) = 0.8$$

Yes.

5/6

But most likely the prev sol^n would work better ^{then..}

When time delay = 20s, we are interested in the longer time scale, ~~then~~ it is an integrating process.

$k' = \frac{\Delta y}{\Delta u \cdot \Delta t} \Rightarrow \frac{10}{1 \times 100} \Rightarrow 0.1 \approx \frac{0.1}{1 \times 99.9}$

$\theta = 20 + 0.1 \Rightarrow 20.1$ $\tau_c = 20.1$

$G'(s) = \frac{k' e^{-\theta s}}{s}$ Simc rules give $K_c = \frac{1}{k'} \frac{1}{(\tau_c + \theta)}$

$\Rightarrow \frac{1}{0.1} \cdot \frac{1}{(20.1) \times 2} = \underline{\underline{0.248}}$ ✓

$\tau_I = 4(\tau_c + \theta) = \underline{\underline{160.8}}$

(b) I would guess, it has an integrator and a zero in the LHP (and maybe a ~~...~~ $(\tau s+1)$ in the denominator?)

$g(s) = k \cdot \frac{(Ts+1)}{s\,\cancel{(\tau s+1)}} e^{-\theta s}$

an integrating pole zero process (IPZ)

where $g(s) = \frac{K}{s} \frac{(Ts+1)}{(\tau s+1)} e^{-\theta s}$

Not quite.

SIMC rules for this give

$K_c = \frac{1}{k \cdot T} \frac{\tau}{(\tau_c + \theta)}$ $\tau_I = \min(\tau, 4(\tau_c + \theta))$

10/10

**Problem 3 (10%)**

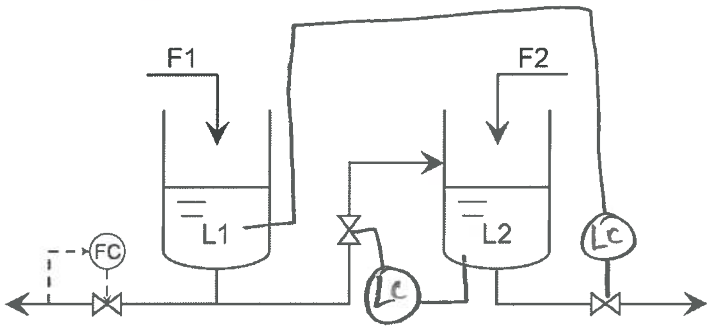
Consistency is crucial property for process control and should be fulfilled at all times

a) Define ("global") consistency and local consistency?
b) Why is consistency a highly desired property?
c) Propose for the following process (i) a control structure which is only globally consistent and (ii) a control structure which satisfy local consistency. Reason for your choice. Assume that F1 and F2 are given and outside of our control range.

(a) When the ~~t~~ mass balance is achieved (no accumulation of mass) in any part of the process, both in individual units and overall system → then it is said to be consistent. In addition if all the inventories are controlled by its local inflow and outflow then it is locally consistent ✓

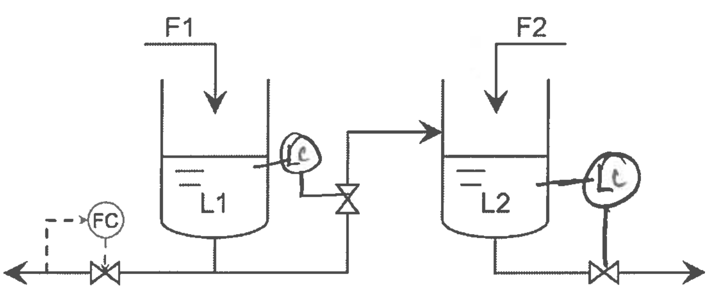
(b) because we do not want accumulation of mass (unbalanced mass) which may lead to oscillatory behaviour and different controllers working against each other ( ~~coun~~ trying to counteract the effect of one another)
✓

(i) Only global consistent control structure:




Levels not controlled by local inflow & outflow and radiation rule not ok! therefore only global not local consistency ✓

(ii) Local consistent control structure:




✓

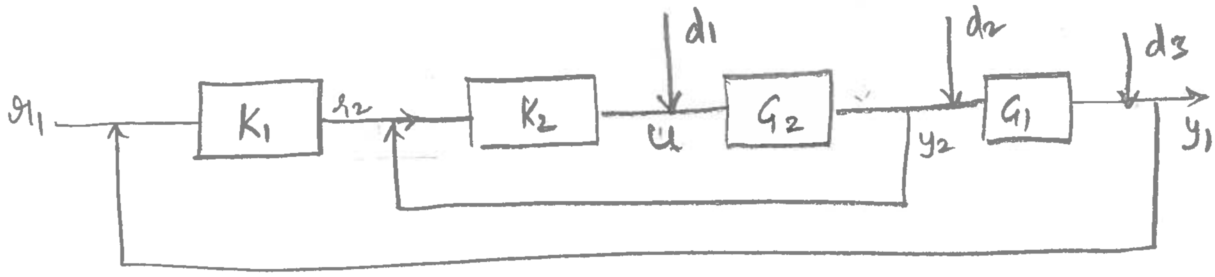
radiating rule ok! and Levels controlled by local outflow Therefore global & local consistency

**Problem 4 (10%)** 10/10

Cascade controllers are frequently used to improve control performance. Answer the following questions regarding control performance.

a) Draw a block-diagram of a cascade control system and explain which disturbances can be easily rejected by cascade control and for which there is less advantage.

b) Give the main reasons for the application of cascade control.




(a) disturbances in $d_1$ can be easily rejected by cascade. but not $d_2$ and ~~$d_3$~~ even worse for $d_3$ ✓

(b) Cascade control is applied when

→ disturbances are large in magnitude ($d_1$) ✓

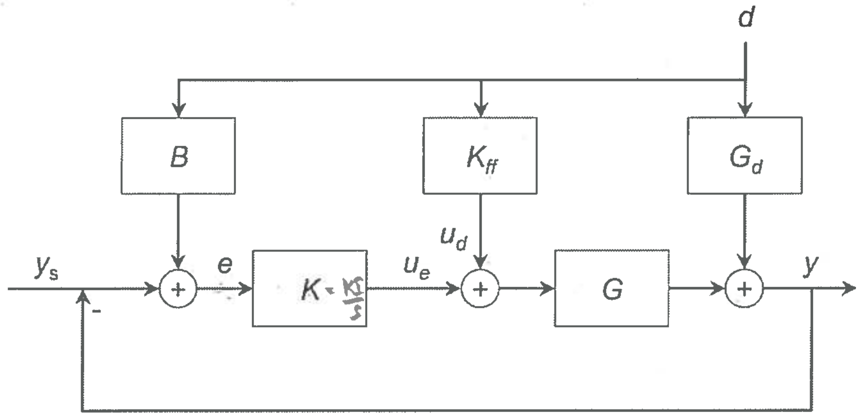
→ there is a large effective delay in $G_1$ ✓

→ when $G_2$ is nonlinear ✓

Wow!

**Problem 5 (20%)** 20/20

Feedforward control can be advantageous if the influence of a disturbance on our measured variables has either a high time constant or time-delay. A general feedforward structure can be represented as the following block-diagram. Here both $K_{ff}$ and $B$ are to be selected.




Consider feedforward control with constant setpoint, $y_s = 0$. The most common way to implement a combined feedback and feedforward controller is

$u = K_{ff}d + Ke$

where $K_{ff}$ is the feedforward controller. This corresponds to $B = 0$ in the block diagram.

Consider an example with $G = 3e^{-s}, G_D = 2$

a) Design a feedback controller $K$ using the SIMC rules. pure Integral control
b) Design a feedforward controller $K_{ff}$ (you may here assume $K = 0$)
c) Krister Forsman mentioned in his lecture that the feedback controller ($K$) may try to counteract the effect of the feedforward controller ($K_{ff}$), leading to an undesired overshoot for y when there is a disturbance (see lower left subfigure of the figure below). This may be avoided with a good choice for $B$. The idea is to make the feedback action independent of $d$. This happens if the transfer function from $d$ to $e$ is 0.
   i) Derive an expression for $B$ which gives this (with symbols only).
   ii) Calculate $B$ for the above mentioned case (with your $K$ and $K_{ff}$).

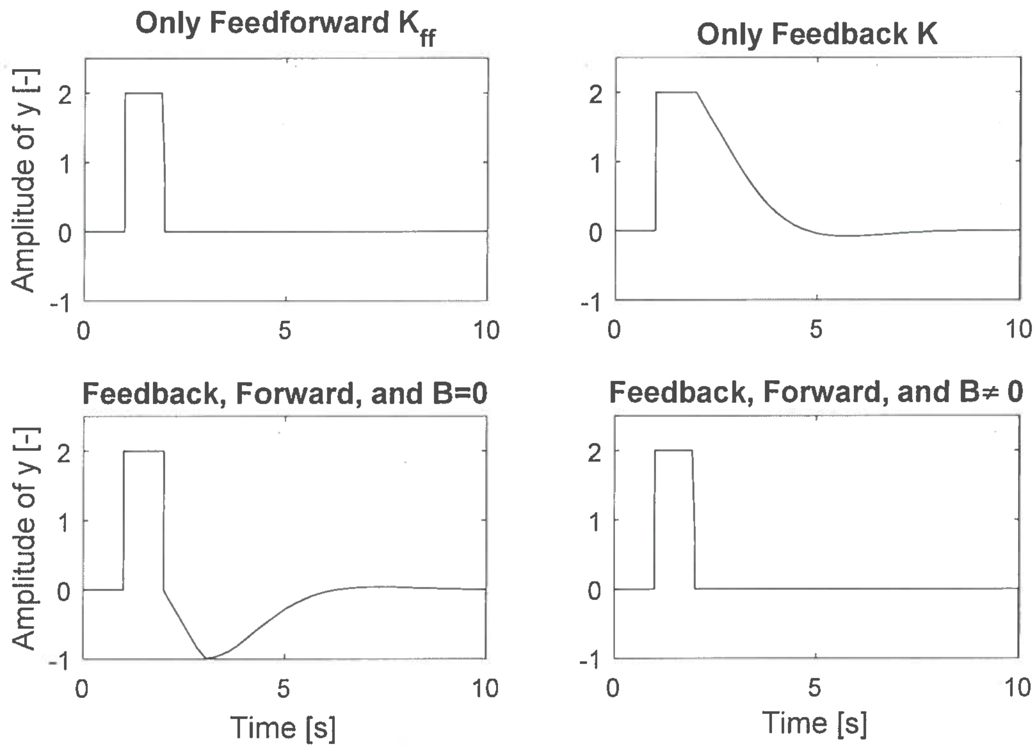
(a) SIMC rules for $G(s) = 3e^{-s}$ (pure time delay process)

for a pure time delay process, we may want to use only pure I control where $K_I = \frac{1}{k}\frac{1}{(\tau_c+\theta)} \Rightarrow \frac{1}{3}\cdot\frac{1}{(\tau_c+1)}$ ✓

& where $\tau_c$ is a tuning parameter and for such a process we may prefer to use tight control with $\tau_c = \theta$ then $K_I = 0.1667$ ✓ and $(K_c = 0)$ Yes

Comment:

The following plots show the advantage of using a matrix $B$ to remove the counteraction of the feedback controller on the effect of the feedforward controller for the above mentioned system.




(b) $K_{ff}$ is generally given by $\cancel{g_d/g}$ $-G_D/G$

$\therefore K_{FF_{ideal}} = \dfrac{-2e^{s}}{3}$

$y = G_d d + G K_{ff} d = 0$

$\therefore (G_d + G K_{ff}) d = 0$

$\Rightarrow K_{ff} = -\dfrac{G_d}{G}$

But it cannot have prediction

$\therefore K_{FF} = \dfrac{-2}{3} \Rightarrow -0.667$ yes

(c) (i) ~~$B = F_r G K_{ff}$~~ where $F_r = \dfrac{1}{(\tau_r s + \theta)}$ is any suitable filter.

$e = Bd + y_s - y \quad (y_s = 0)$

(c) $e = Bd - y \Rightarrow Bd - [G_d d + G(K_{ff} d + Ke)]$

$e = Bd - G_d d - G K_{ff} d - GKe$

$\Rightarrow e(I + GK) = (B - G_d - GK_{ff})d$

$\frac{e}{d} = (1+GK)^{-1}(B - G_d - GK_{ff})$ = transfer function from $d \to e$ must be equal to zero.

$(1+GK)^{-1}(B - G_d - GK_{ff}) = 0 \Rightarrow B - G_d - GK_{ff} = 0$

(i) $\boxed{B = G_d + GK_{ff}}$ ✓

$2 - 3e^{-s} \times \frac{2}{3}$

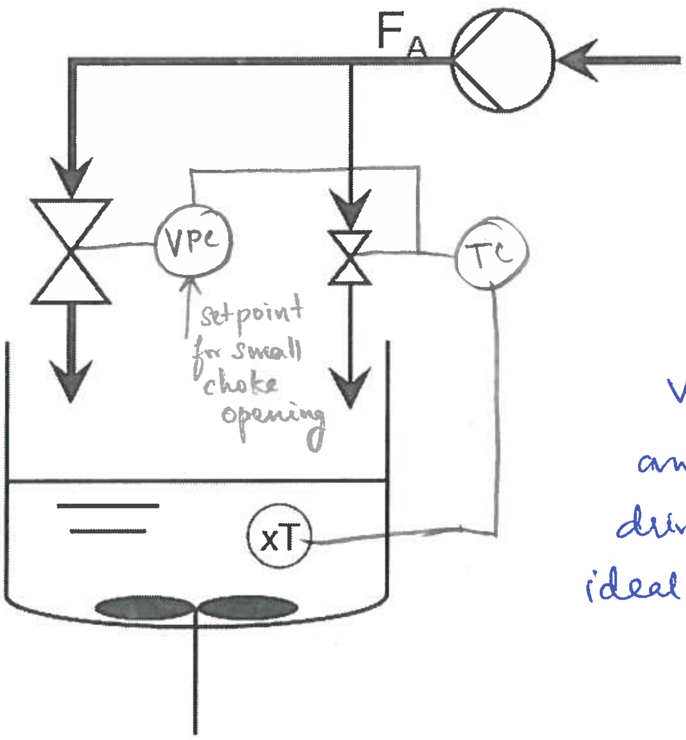
(ii) $\boxed{B = 2 - 2e^{-s}}$ ✓

**Problem 6 (10%)** 10/10

Consider the semi-batch reactor shown below. In this process, the reactant A is constantly fed from a not shown tank. Due to the danger of a thermal runaway, we need tight control of the concentration of our reactant A.

In this system, we have two valves for the feed of A, a large coarse (poor) valve which has discrete steps and a small, accurate valve.

Propose a control structure for this process. What is it called?

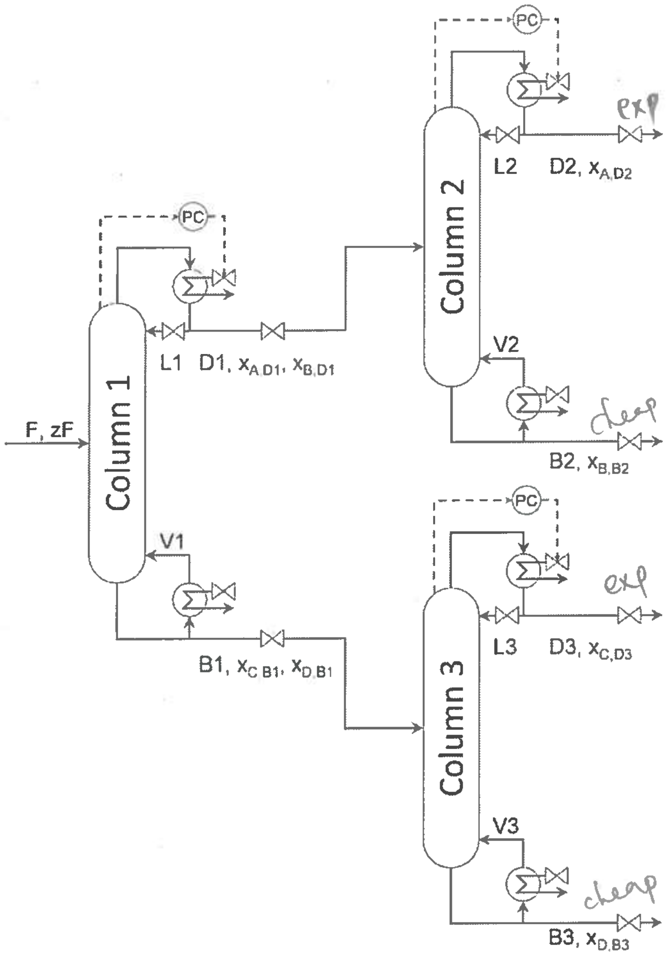



We cannot get tight control with the large poor valve so, we use the small accurate valve for tight ~~temperature~~ composition control and use the big valve to drive the small valve to an ideal openening (e.g 5% or something small)

Valve position control / Input resetting or midranging control

**Problem 7 (20%)** 20/20

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

$p_A = 5\$/\text{mol}$
$p_B = 1\$/\text{mol}$
$p_C = 5\$/\text{mol}$
$p_D = 1\$/\text{mol}$

A and C are expensive

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:

$p_V = 0.0001\$/\text{mol}$ cheap energy

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?

(a) 15 dynamic degrees of freedom. The liquid levels in the coolant tanks and column have to be controlled

∴ 15 − 6 ⇒ 9 SS DOF

in addition, the pressure in each column is controlled

⇒ 9 − 3 ⇒ 6 DOF left ✓

(b) The expensive products will always be active (to avoid product giveaway)

(selg gråstein som gull)

$X_{A,D2} = 97\%$ A ✓

$X_{C,D3} = 97\%$ C

in addition, due to low energy prices, the vapor flow rates will be active, since we want to overpurify the cheap components to get more of the valuable products (A, C) ✓

∴ $V_1 = 5$ mol/s

$V_2 = 3$ mol/s

$V_3 = 4$ mol/s

∴ We will have 5 active Constraints

(c) Explanation of control structure:

→ Temperature control on all columns to stabilize the Temp profile
it gives additional benefits like better level control in condensate level, better disturbance rejection & less interactions.

TC is also placed in the section of the most valuable product
∴ TC is controlled using L1, L2 and L3
The temp for TC is then the DOF.

→ Column levels are controlled using the B1, B2 and B3 respectively due to "pair close" rule ✓

→ We have chosen LV configuration, due to the benefits of LV configuration over DV configuration. Therefore coolant tank levels are controlled using $D_1$, $D_2$ and $D_3$ ✓

→ Control active constraints

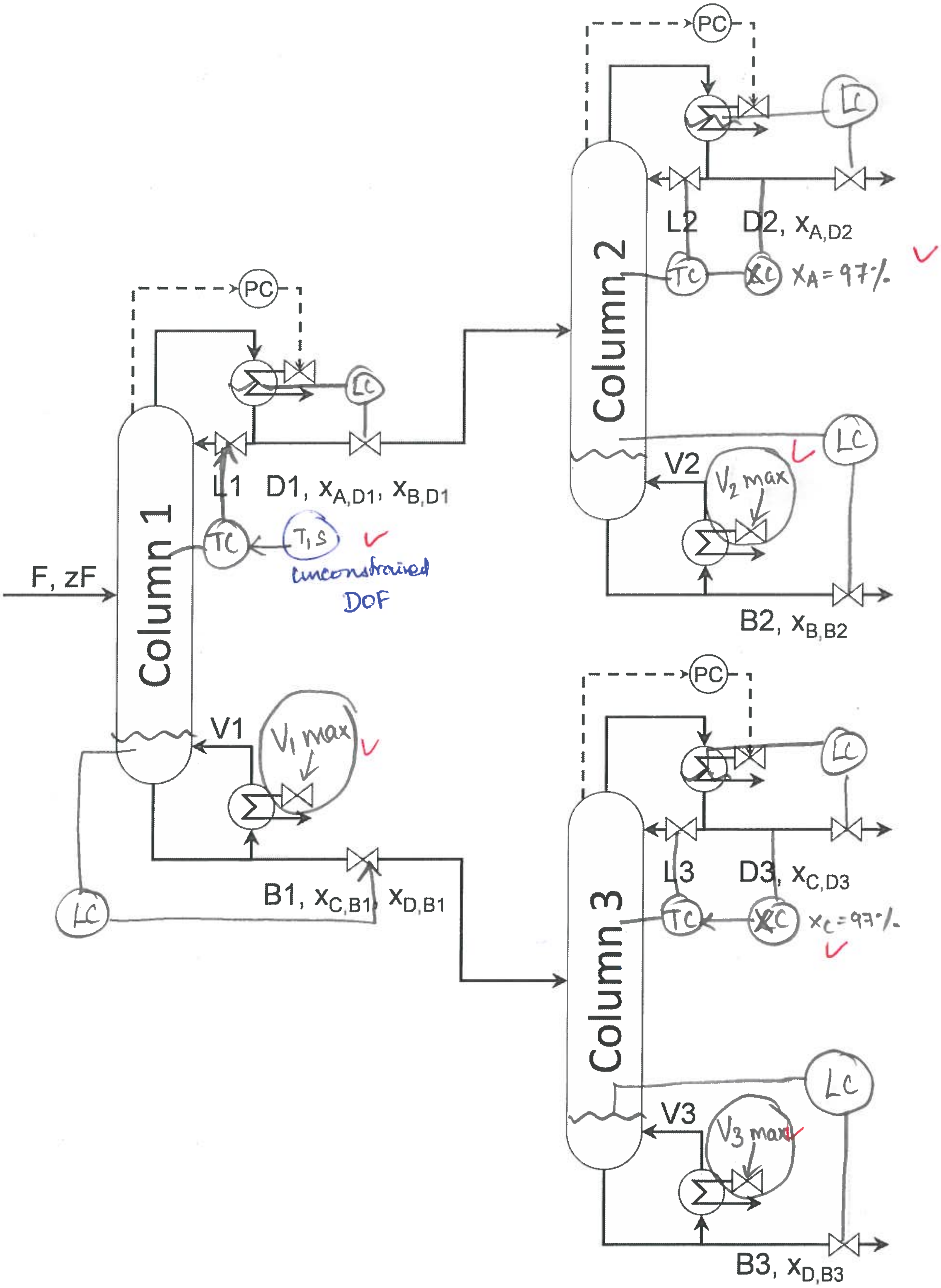
Control $V_1 = V_1 max = 5$ mol/s

Control $V_2 = V_2 max = 3$ mol/s ✓

Control $V_3 = V_3 max = 4$ mol/s

Control $X_A$ composition using $T_2$, setpoint at 97% A
Control $X_C$ composition using $T_3$ setpoint at 97% C
} Cascade Controllers

→ $T_1$, setpoint remaining unconstrained degree of freedom.
need to find self optimising variable
can keep $L_1$ constant or keep $\frac{L}{F}$ constant


PC
LC
L2
D2, xA,D2
TC
XC
XA=97%
Column 2
PC
LC
L1
D1, xA,D1, xB,D1
TC
T1,s
unconstrained DOF
F, zF
Column 1
LC
V2
V2 max
B2, xB,B2
PC
LC
V1
V1 max
B1, xC,B1, xD,B1
LC
L3
D3, xC,D3
TC
XC
xc=97%
Column 3
LC
V3
V3 max
B3, xD,B3

(d) Can control $L_1$ at constant setpoint or
Can control $L/F$ at a constant setpoint.

or T at constant setpoint.

Alternatively, can use null space method / Exact Local method to compute optimal measurement combination.

Yes.

(e) if feed increases, TPM could be moved to the bottleneck i.e. the last constraint which becomes active when the feed increases.

f) If they are hard constraints, then we need to apply back off which is the safety margin to remain feasible.

back off = measurement error + control error. (for hard constraints

By having a tighter control, we can reduce the variance and hence move the setpoint closer to the constraint. This is the squeeze and shift method.

This is implemented using cascade controllers where TC are used for tight control and XC are used for controlling the $X_A$ and $X_C$ compositions by providing setpoints to the TC. See ~~b~~ answer to question C, where TC are used in cascade (instead of controlling the compositions of $X_A$ and $X_C$ directly using $L_2$ and $L_3$ which would have ~~will~~ require large back off)