Emnemodul: Advanced Process Control

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.

Problem 1 (20%)

2.5 for each but 22.5 max 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 $F = GJ_{uu}^{-1}J_{ud} + G_d$. 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?

(a)

(b) It is a complement to both RTO and MPC

(c) $dy_{opt} = G\,du_{opt} + G_d\,dd \Rightarrow \frac{dy_{opt}}{dd} = G\frac{du_{opt}}{dd} + G_d$ F

Here we have $J_u = J_{uu}^*\underbrace{(u-u^*)}_{du_{opt}} + J_{ud}^*\underbrace{(d-d^*)}_{dd}$. Assume nominal optimum ($J_u^*=0$).

Then to remain optimal we need $J_u = 0 \Rightarrow J_{uu}^*du_{opt} + J_{ud}\,dd = 0 \Rightarrow \frac{du_{opt}}{dd} = -J_{uu}^{-1}J_{ud}$

q.e.d.

(d) $HF=0$

(e) No noise and $n_y \geq n_d + n_u$. $c = Hy \Rightarrow c_{opt} = H(G^y u_{opt} + G_d^y d) = HFd$. Need $HF=0$ to remain optimal for all d.

(f)

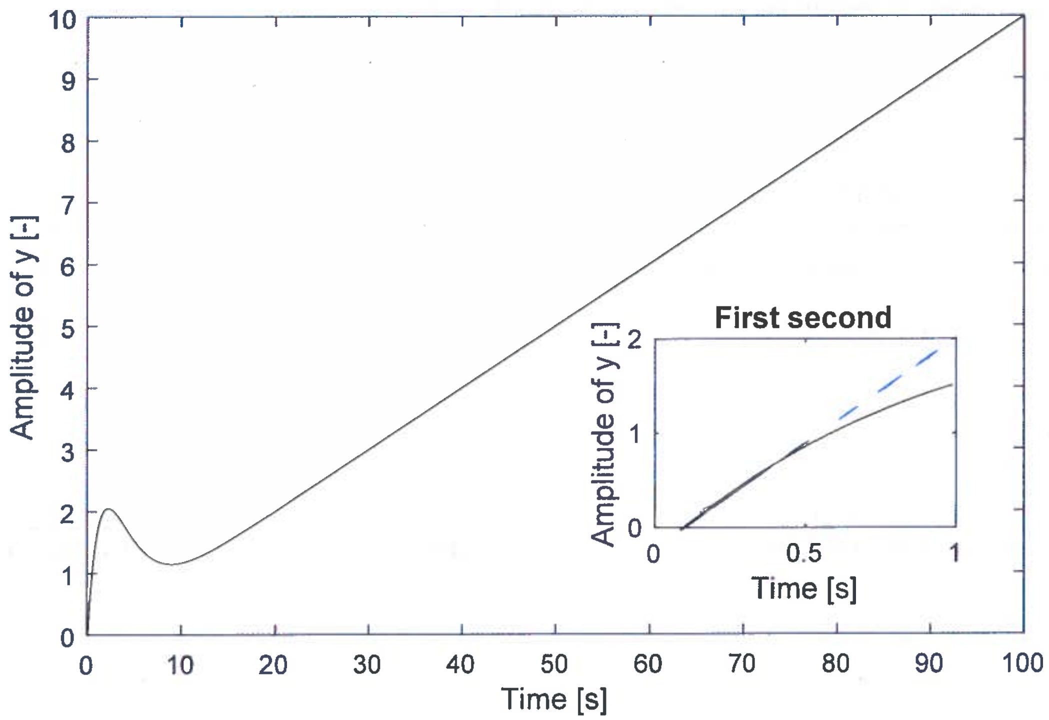
(g) includes meas noise and does not have the restriction $n_y \geq n_u + n_d$.

(h) Because when $J_u = 0$ we are at the optimal point.

(i) No. Can give infeasibility if we fix J.

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.

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))?

6 each but max 10

(a) (i) Integrating + delay model. $G = \frac{k'}{s} e^{-\theta s}$ with $k' = 2.5$ and $\theta = 0.1$

SIMC-rule. $K_c = \frac{1}{k'} \frac{1}{\tau_c + \theta} \overset{\tau_c = \theta}{=} \frac{1}{2.5} \frac{1}{0.2} = 2$

$\tau_I = \min(\tau_1, 4(\tau_c + \theta)) = 4(\tau_c + \theta) = 0.8$

(ii) In this case $k' \approx \frac{10}{100} = 0.1$ and $\theta = 20.1$. $K_c = \frac{1}{k'} \frac{1}{\tau_c + \theta} \overset{\tau_c = \theta}{=} \frac{1}{0.1} \frac{1}{40} = 0.25$, $\tau_I = 160$

Note: Simulations show that both controllers work OK. Try!

Extra max 5

(b) The model is $G = \left[\frac{10 s}{(2s+1)^2} + \frac{0.1}{s}\right] e^{-0.1 s}$

Problem 3 (10%)

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

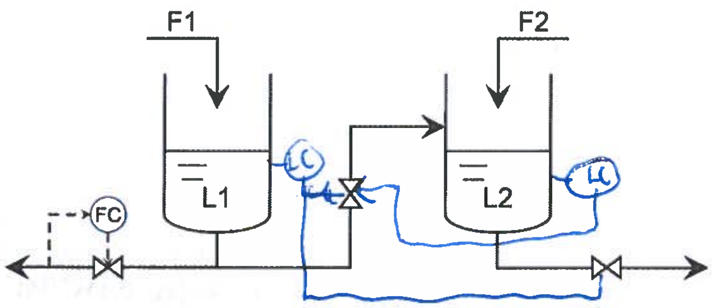
a) Define ("global") consistency and local consistency?
b) Why is consistency a highly desired property?
c) Propose for the following process (i) a control structure which is only globally consistent and (ii) a control structure which satisfy local consistency. Reason for your choice. Assume that F1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

4 (a) Consistency = Mass balances satisfied at steady-state.
Local " : Using only local loops

2 (b) Otherwise need "high-level" controller to make it work or it simply fails

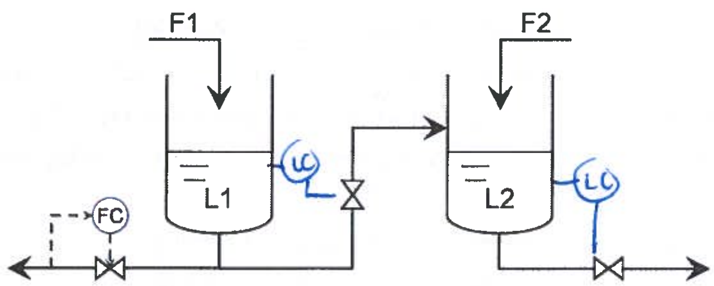
(i) Only global consistent control structure:

2



(ii) Local consistent control structure:

2



Problem 4 (10%)

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

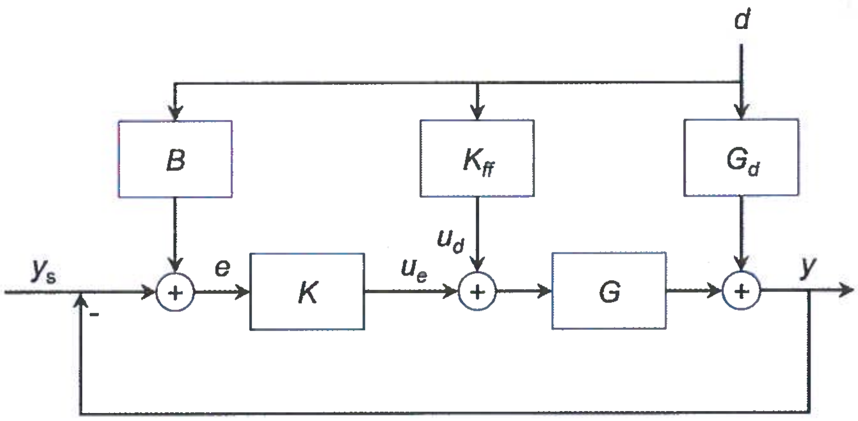
6 a) Draw a block-diagram of a cascade control system and explain which disturbances can be easily rejected by cascade control and for which there is less advantage. Dc

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

- Disturbances before slave measurement
- Nonlinearity inside slave
- Delay in meas. of CV. ~~...~~ but slave is fast

Problem 5 (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 $G = 3e^{-s}, G_D = 2$

Set $K_c = 0, \tau_I = 0$, but $k_I = K_c/\tau_I$ finite, $\tau_c = \theta$

Pure integral controller with $K_I = \frac{1}{k}\frac{1}{\tau_c + \theta} = \frac{1}{3}\frac{1}{2} = 0.17$

max 20

7 a) Design a feedback controller K using the SIMC rules.

7 b) Design a feedforward controller K_{ff} (you may here assume K=0) . $y = (G_d + G K_{ff})d = 0 \Rightarrow K_{ff} = -G^{-1}G_d = -\frac{1}{3}\cancel{e^{s}} \cdot 2 = -0.67$

7 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.

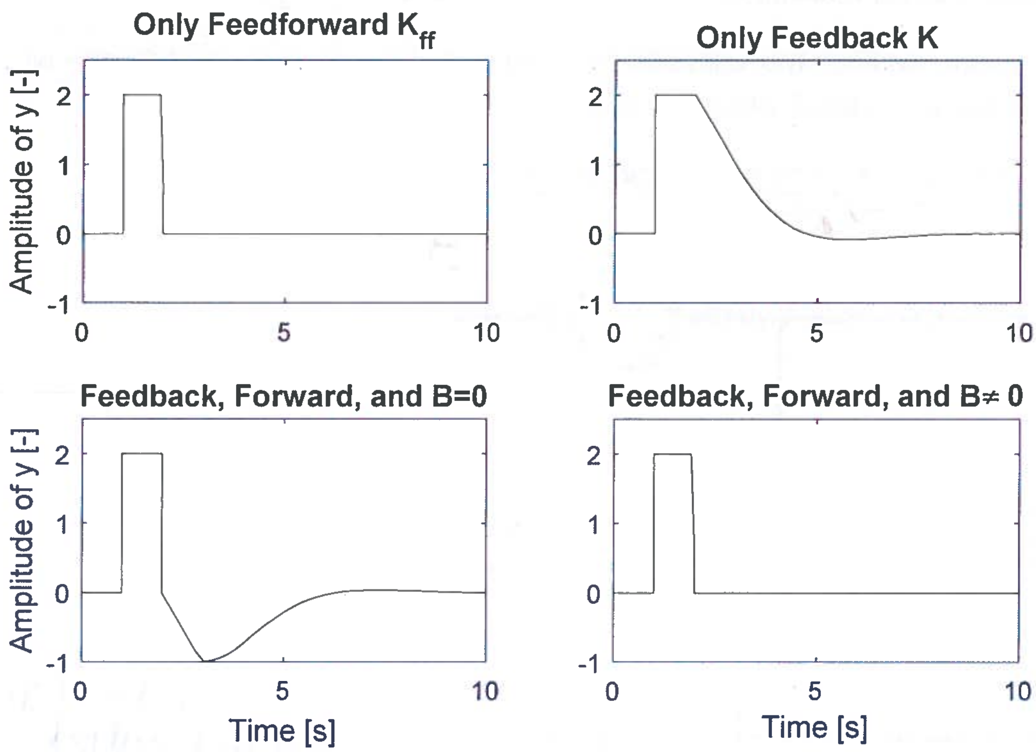
   i) Derive an expression for B which gives this (with symbols only).

   ii) Calculate B for the above mentioned case (with your K and K_{ff}).

CANNOT HAVE PREDICTION!

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.



(i) Want to make transfer function from d to e zero. Then the feedback control will not change the feedforward action.

Get (set $K=0$)

$$e = Bd + GK_{ff}d + G_d d = 0$$

$$\Rightarrow B + GK_{ff} + G_d = 0 \Rightarrow B = +GK_{ff} + G_d \qquad (*)$$

(ii)

$$B = +3e^{-s}\left(-\frac{2}{3}\right) + 2 = -2e^{-s} + 2 = \underline{\underline{-2(e^{-s}-1)}} = \underline{\underline{2(1-e^{-s})}}$$

Note: The use of a B-term is probably not always a good idea for feedforward control, because it only makes a difference when the feedforward is not perfect (with perfect feedforward B=0 in (*)) and when feedforward is not perfect we may of course have benefit of feedback!

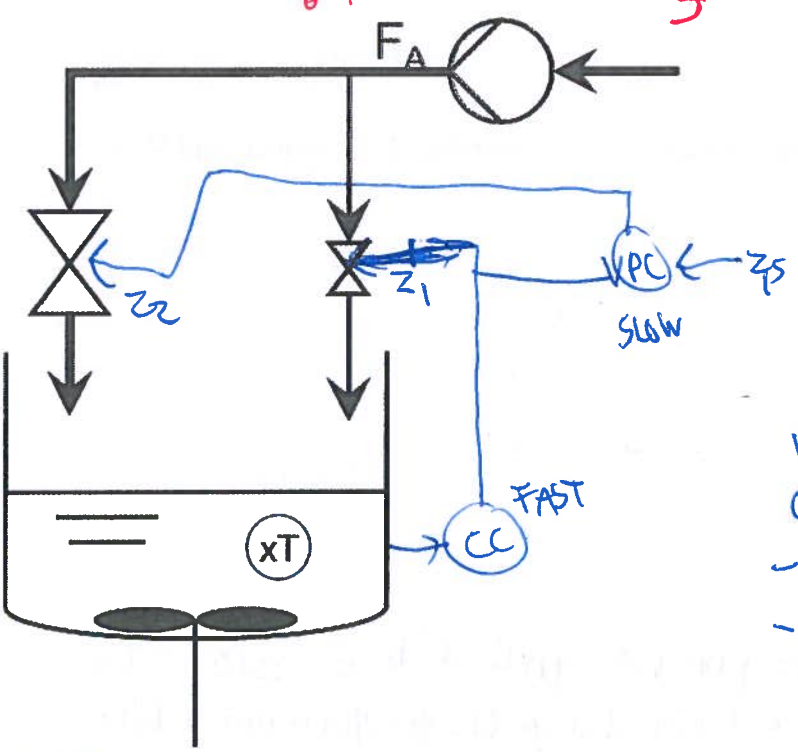
Problem 6 (10%)

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?

7 3

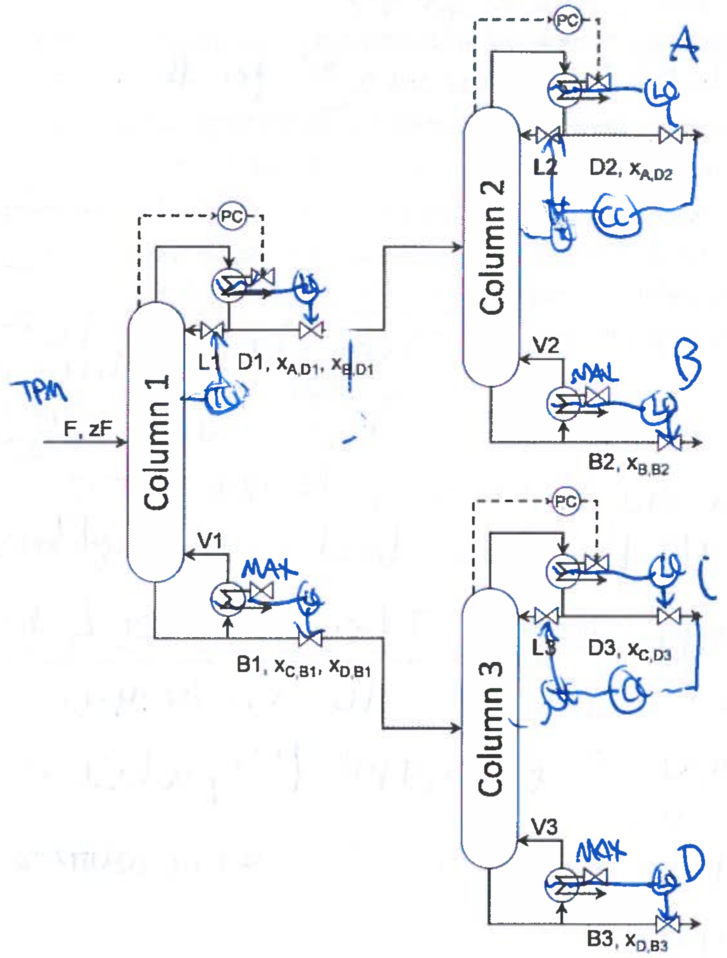


We have two inputs and one output.
- want z_1 to return to z_{1s}.
- z_2 is used for steady-state control.

Names:
Valve position control, Midranging control, input resetting.

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

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

p_A p_B p_C p_D

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

In which the product prices are given by

$p_A = 5$ \$/mol

$p_B = 1$ \$/mol

$p_C = 5$ \$/mol

$p_D = 1$ \$/mol

D2 D3

A and C compositions active to avoid product give-away.

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:

$p_V = 0.0001$ \$/mol (cheap energy → constraints on V_1, V_2, V_3 active

Based on the above information, answer the following questions.

12 (with PC closed and feed given) 6 (6 levels with no steady-state effect)

2+2 a) How many dynamic and steady-state degree of freedoms does this system have?

4 b) Based on your experience and engineering know-how, which constraints will be active for the above mentioned system. Justify your answer. See prev. page

6 c) Propose a control structure for this case and draw it in the figure on the next page. Explain your choice. One unconstrained DOF (L1). Probably use a TC for this

3 d) If you have degree of remaining freedoms that are not used for controlling active constraints or stabilizing inventories, then propose possibilities on how to use them.

3 e) What happens, if the feed rate is increased? Would you suggest moving the TPM? See Figure

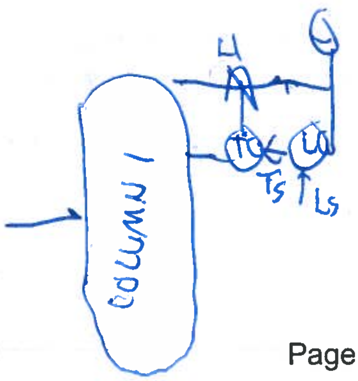
3 f) The product streams of A and C shall be sold directly to the customers, hence the product composition constraints are hard constraint and may not be violated at any point. Can the idea of squeeze and shift be used? How would you apply it? Make sure these two composition controllers are tightly tuned. Should be easy if we also add TC

5 red 20

(e) When F is increased, first one of x_B or x_D becomes active and then the other of the two. The last is the bottleneck.

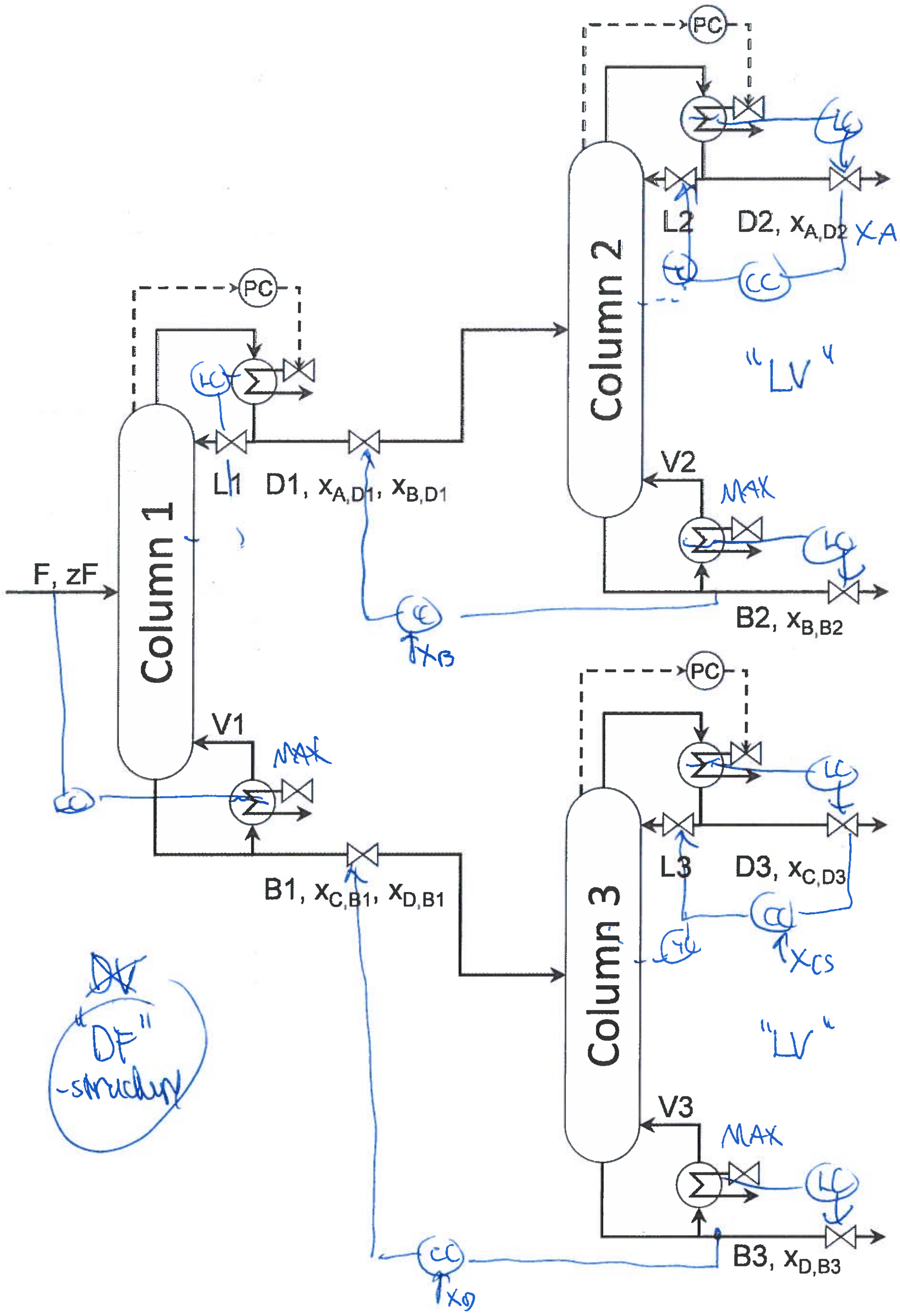
Assume x_B first becomes active. Then we use L_4 to control x_B, which is a "long" loop. Later x_D becomes active and we must use F to control this, which is another "very long loop": So move TPM to feed of column 2 or column 3. See Figure

Problem: We have no control of the "composition profile" or temperature in column 1. We could maybe cascade the level controller to a temperature controller, although I have never seen this done.



Hm...

Or better: Use LV in column 1 and let CC for x_B set setpoint for TC


PC
PC
PC
Column 1
Column 2
Column 3
F, zF
L1
D1, xA,D1, xB,D1
V1
B1, xC,B1, xD,B1
L2
D2, xA,D2
V2
B2, xB,B2
L3
D3, xC,D3
V3
B3, xD,B3
"LV"
"LV"
MAX
MAX
MAX
"DF"
-structure