Emnemodul: Advanced Process Control

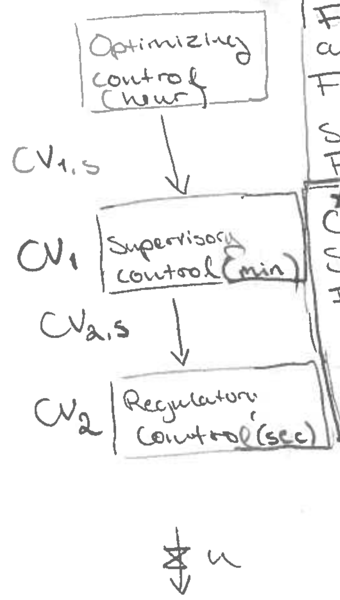
01. Dec. 2017. Time: 0915 – 1200.
Answer as carefully as possible, preferably using the available space.
If possible, do not write on the backside of the exam.
You may answer in Norwegian; however, English is preferred.

Problem 1 – General Questions (15%)

a) Typically, there exists a control hierarchy in chemical plants. State the different layer of this control hierarchy.
b) What are the tasks in each layer?
c) What is the principle of self-optimizing control (SOC)?
d) In which section of the control hierarchy would you position self-optimizing control?
e) Is it possible to combine SOC with model predictive control (MPC)? Reason your answer!
f) Give two advantages and two disadvantages of MPC.
g) List two properties of good controlled variables.

Problem 1

a) and b)



* Optimizing control
RTO (Real-time optimization)
Find cost function and constraints.
Find CV_1 (economic control variables) and pair with MV_1.
Set $CV_{1,s}$ setpoints and
Find optimal operation point

* Supervisory control
Control CV_1 with MV_{1s}.
Switch between regions of active constraints.
Find CV_2 and MV_2 for every region (Regulatory controlled variables) and give setpoints, CV_{2s}, to regulatory layer.

* Regulatory control
Stabilize the plant by controlling CV_2.

c) The principal of SOC is to minimize the loss of known disturbances. SOC uses degrees of freedom left after controlling the active constraints to controll variables that are insensitive to disturbances and therfore minimizes the loss.

d) I would place SOC in the supervisory control layer. SOC is a part of the economic control variables, CV_1. The setpoints of CV_1's are found in the optimizing control layer, but are controlled in the supervisory control layer

e) Yes, MPC can be combined with SOC.
MPC can be used for real time optimization and supervisory control.
· MPC can decide what SOC should be used in different active constraints regions.

f) MPC:

Advantages:

+ Can easily switch between active constraints regions
+ Can take into account interactions between different inputs

Disadvantages:

- Need a good, updated model of your process plant
- Need more CPU compared to decentralized control

g) Good controlled variables

1) Easy to measure
2) Insensitive to disturbances
3) Sensitive to change in input, meaning a large gain $\Delta G = \frac{\Delta c}{\Delta u}$. (Dominant variables).

Problem 2 – PID controller tuning (15%)

Consider a process given by the following process model

$$G_1(s) = \frac{300}{(100s+1)(10s+1)} e^{-5s}$$

Additionally, you will have a measurement model with time delay θ_m given by

$$G_m(s) = e^{-\theta_m s}$$

The aim in this task is to tune a controller using the SIMC rules with $\tau_c = \theta$ (effective delay) for the following two cases:

a) The time delay in the measurement function is given by $\theta_m = 1$.
b) The time delay in the measurement function is given by $\theta_m = 10$.

Consider the following arbitrary first order process

$$G_2(s) = \frac{k}{\tau_1 s + 1}$$

c) Derive and simplify (!) the closed-loop transfer function from the setpoint y_s to the measurement y $\left(T = \frac{y}{y_s}\right)$ using a PI controller tuned with the SIMC rules and $\tau_c = \tau_1$.

a) PID - controller: (τ_2 - is dominant)

$\tau_1 = 100$

$\tau_2 = 10$

$\theta = 5 + 1 = 6$

$\underline{K_c} = \frac{1}{k} \frac{\tau_1}{(\tau_c + \theta)} = \frac{1}{300} \frac{100}{(6+6)} = \frac{1}{36} = \underline{\underline{0.027}}$

$\underline{\tau_I} = \min(\tau_1, 8\theta) = \min(100, 8 \times 6) = \underline{\underline{48}}$

$\underline{\tau_D} = \tau_2 = \underline{\underline{10}}$

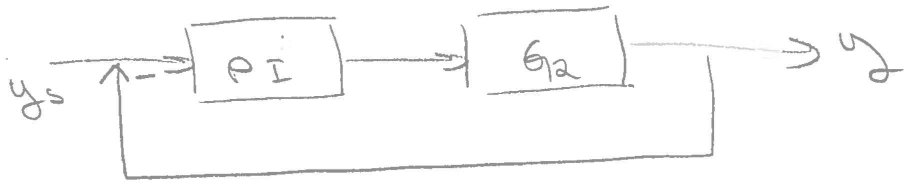
b) $\tau_1 = 100$, $\tau_2 = 10$ $\theta = 5 + 10 = 15$

$\underline{K_c} = \frac{1}{(300)} \frac{100}{(15+15)} = \frac{1}{90} = \underline{\underline{0.011}}$

$\underline{\tau_I} = \min(\tau_1, 8\theta) = \min(100, 8 \times 15) = \min(100, 120) = \underline{100}$

$\tau_D = \tau_2 = 10$

c)



$$G_2 = \frac{K}{\tau_1 s + 1}$$

<u>SIMC rules</u>: $\tau_c = \tau_1$

$$K_c = \frac{1}{K}\frac{\tau_1}{(\tau_c + \theta)}$$

$$\tau_I = \min\left(\tau_1, 4(\tau_1 + \theta)\right) = \tau_1$$

<u>PI</u> controller: (series form)

$$K_c\left(1 + \frac{1}{\tau_I s}\right) = K_c\left(\frac{\tau_I s + 1}{\tau_I s}\right)$$

$$\{\tau_I = \tau_1\}$$

$$G = PI \times G_2 = K_c\left(\frac{\cancel{\tau_I s + 1}}{\tau_I s}\right)\left(\frac{K}{\cancel{\tau_1 s + 1}}\right) = \frac{K_c K}{\tau_1 s}$$

Closed loop response

$$\frac{y}{y_s} = \frac{G}{1+G} = \frac{\left(\frac{K_c K}{\tau_1 s}\right)}{1 + \left(\frac{K_c K}{\tau_1 s}\right)} = \underline{\underline{\frac{K_c K}{\tau_1 s + K_c K}}}$$

Problem 3 – Advanced Control Structures (15%)

The relative gain array (RGA) is a tool one can use to decide on controller pairing in multivariable systems. Additionally, it gives you information about the influence on coupling. Consider the following steady state RGA:

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

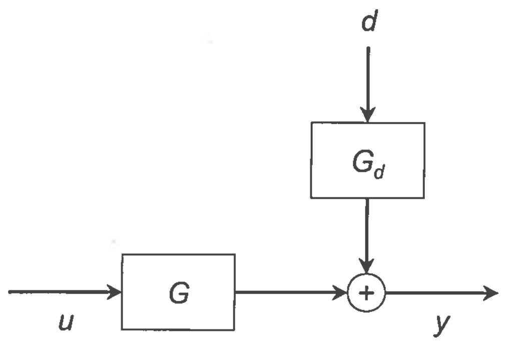
a) How would you pair the inputs with the controlled variables?

b) What are the implications if you pair on the following RGA values $\lambda_{i,j}$?

1. $\lambda_{i,j} < 0$
2. $0 < \lambda_{i,j} < 1$
3. $1 < \lambda_{i,j}$

Feedforward control is frequently used in process control. It may however lead to problems if not tuned properly.

c) Draw a feedforward control structure in the following block diagram.



d) In which situation is it advisable to incorporate a feedforward controller?

e) Why can you not always use a perfect feedforward controller?

Problem 3

a) I would suggest the following pairing

$$K = \begin{bmatrix} 0 & 1 & 0 \\ 0 & 0 & 1 \\ 1 & 0 & 0 \end{bmatrix}$$

We want to use steady state RGA close to 1.

b) RGA, $\lambda_{ij} = \dfrac{(y_i/u_j)_{OL}}{(y_i/u_j)_{CL}}$

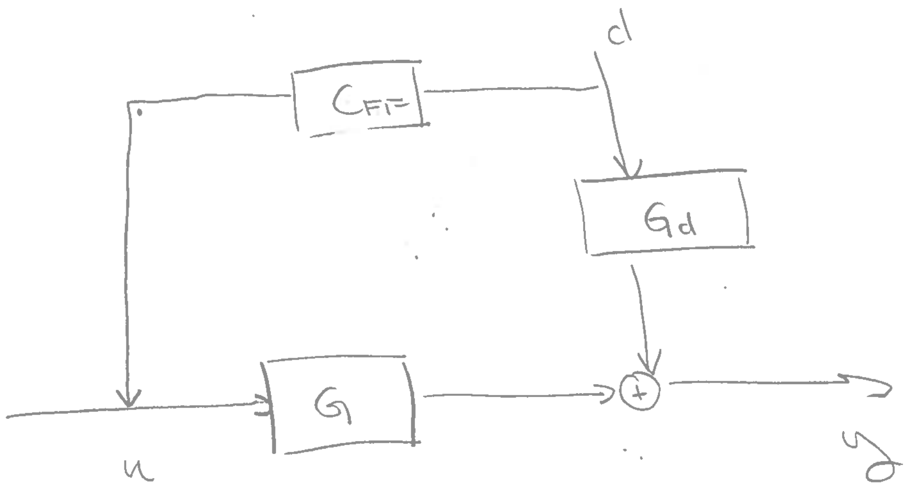
1. $\lambda_{i,j} < 0$ =
2. $0 < \lambda_{i,j} < 1$
3. $1 < \lambda_{i,j}$ =>

1. Here the closed loop gain is working in the opposite direction compared to the open loop gain. With this pairing we can drive the prosess in the wrong direction.

2. The closed loop gain is larger than the open loop gain. This can be dangerous and difficult to tune. Not very robust.

3. The closed loop gain is smaller than the open loop gain. This pairing will not be optimal since the gain might be too small.

c) Feed forward block diagram



d) Use feed forward controller when

1. The disturbance is known and can be measured.
2. Feed forward requires a good model of the process, G and the disturbance, Gd
3. C_{FF} has enough time to act if the effective time delay in Gd is large.

e) Do not use feed forward controller when the model Gd is unknown. The feedforward controller can also increase the gain of the process too much, if the gain is increased with a factor of 2, do not use feed forward.

f) You can not always use a perfect feed forward controller if the models G and Gd are unknown. It is therefore common to use a static feed forward controller. To make sure that the feed forward controller does not overreact, the optimal gain is multiplied with a factor 0.8.

Problem 4 – Consistency (15%)

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

a) What is the definition of consistency and the more stringent local consistency?
b) What is the "so-called" throughput manipulator (TPM)?
c) What does the radiation rule say?

a) Consistency: The overall mass balance should be fulfilled. A prosess is consistent if the inlet or the outlet flows depend on the inventory.

Local consistency: For every process unit

i) One of flows (inlet or outlet) should depend on the inventory

ii) If the process unit includes more than one phase: ~~For every phase~~ one of the flows (inlet or outlet) should depend on the inventory of that phase.

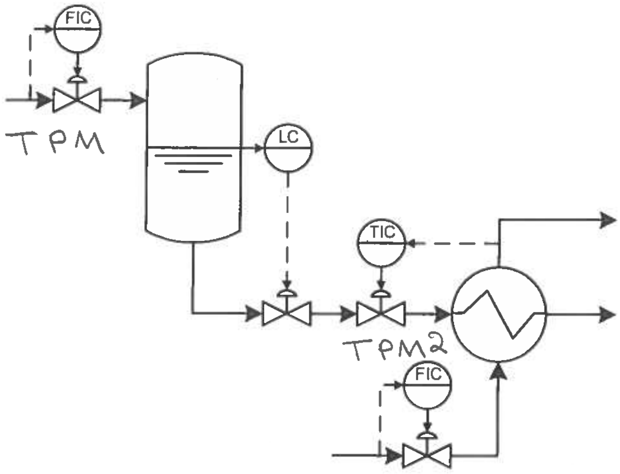
iii) If the process unit includes more than one component: For each componentent, one of the flows (inlet or outlet) should depend on the inventory of that component.

c) TPM: Sets the total throughput of the whole process.

d) The regulatory controllers (level, pressure) should radiate away from the TPM. This will make your process consistent.

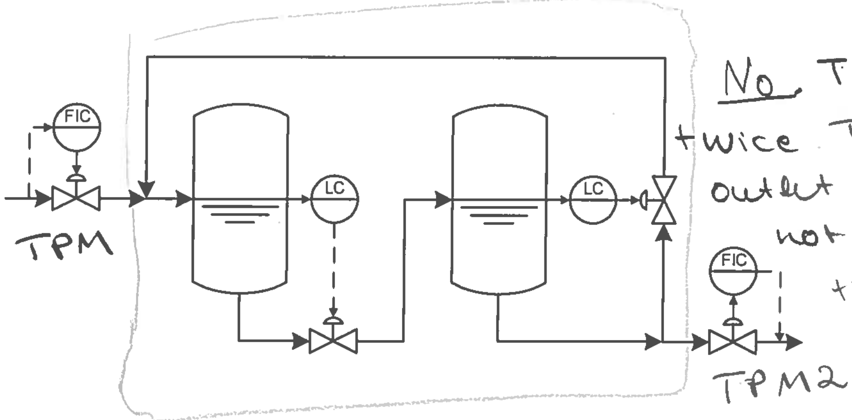
d) Are the following control structure consistent and what is (are) the TPM(s)? Justify your answers for global consistency. Do not consider gas hold-up!

i)



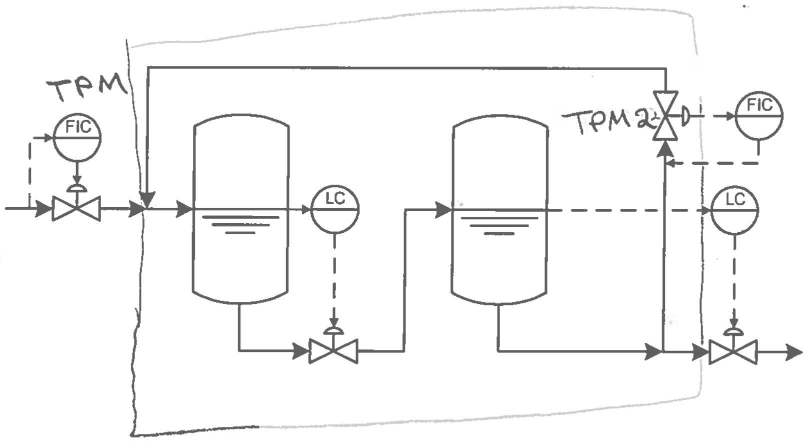
No.
The TPM is set twice. Neither the outlet flow or the inlet flow depend on the inventory.

ii)



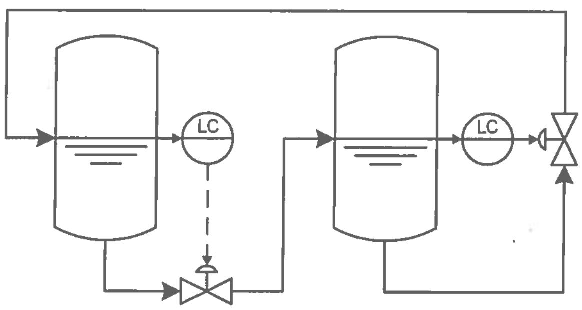
No. TPM is set twice. The inlet and outlet flow does not depend on the inventory

iii)



Yes.
The second FIC is on since the flow is split.
The outlet flow depends on the inventory

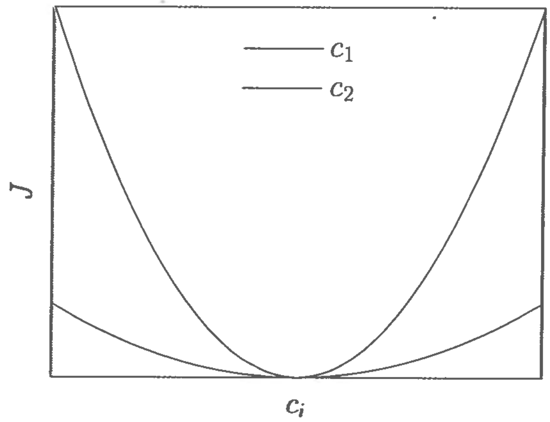
iv)



No
In a closed-loop process, one of the units should be left uncontrolled.

Problem 5 – Self–Optimizing Control (20%)

a) Consider the following plot showing the cost function J as a function of two different controlled variables c_i. Which of the two controlled variables would you implement? Why?



b) The nullspace method is one method, which can be used in the calculation of a selection matrix **H**. For this method, answer the following questions:
1. How many measurements are required for the nullspace method?
2. How do you calculate the selection matrix **H**?
3. 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 generalization of the nullspace method and a second method to calculate the selection matrix **H**.
1. How is the selection matrix **H** calculated in this method?
2. What is the advantage of the exact local method compared to the nullspace method?

a) I would have implemented $\underline{C_2}$.
We want a C_2 that is insensitive to disturbances and measurement noise, and with a large gain

$$G_2 = \frac{\Delta C_2}{\Delta u}$$

b) 1. $n_y \geq n_u + n_d$ measurements are required
2. $HF = 0$ where F is the sensitivity matrix. $F = \frac{\partial y}{\partial d}\Big|_{u_{opt}}$

c) The nullspace method is optimal when there is no noise in our measurements $W_n = 0$

$C = Hy$

$$\frac{\Delta C_{opt}}{\Delta d} = H \frac{\Delta y}{\Delta d} \Rightarrow \Delta C_{opt} = HF\Delta d = 0$$

$HF = 0$ when there is no measurement noise, meaning that we have perfect controll $\Delta C = 0$

c) Exact local method

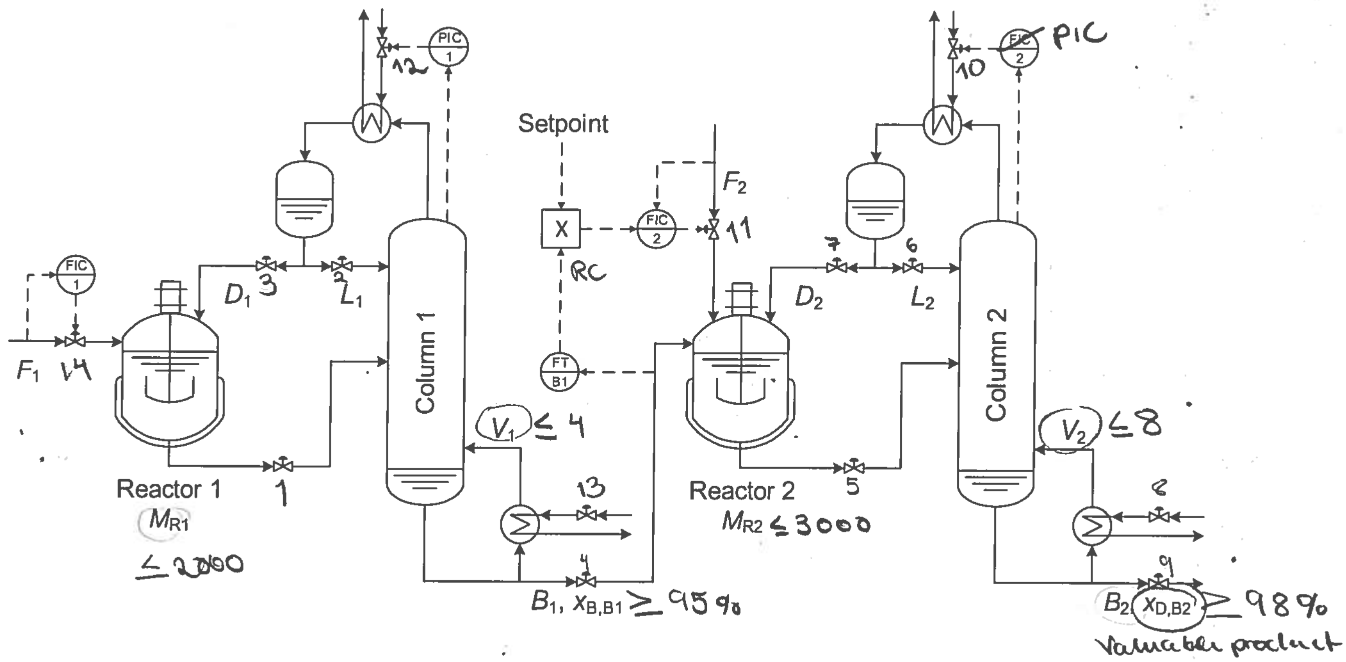
$$Y = [F W_d \quad W_n]$$

$$H^T = (Y Y^T)(G^y)^{-1}$$

d) The exact local method includes noise, W_n, and does not have any restrictions when it comes to number of measurements.

Problem 6 – Applied Plantwide Control (20%)

Consider the following process with two serial reactor and distillation columns.



In Reactor 1 (with hold-up M_{R1}), chemical A (fed through F_1) reacts to chemical B according to the following reversible reaction:

$$2A \rightleftharpoons B$$

Chemical A is then separated from chemical B in Distillation column 1 and refed to Reactor 1 through D_1. The separated chemical B (stream B_1) is then fed together with a fresh stream F_2 containing chemical C in excess of chemical B to Reactor 2. Chemical B and chemical C react then to chemical D in Reactor 2 (with hold-up M_{R2}) according to

$$B + C \rightarrow D$$

Chemical D is then separated in column 2 and chemical B and C recycled to Reactor 2 through the distillate D_2.

As unwanted side reaction,

$$A + C \rightarrow E$$

Is simultaneously occurring in Reactor 2 for conversion of remaining chemical A.

This process was running now for several years and due to increased competition in the market, our objective is now to minimize the operating costs, which are given by:

$$J = -\text{Profit} = p_{F_1}F_1 + p_{F_2}F_2 + p_V\left(V_1 + V_2\right) - p_{B2}B_2$$

Where p_X denotes the price for quantity X, and the capital letters denote the streams.

The values for the prices are:

$p_{F_1} = 1\ \$/mol$
$p_{F_2} = 0.5\ \$/mol$
$p_{B_2} = 5\ \$/mol$
$p_V = 0.001\ \$/mol$

The price for energy can herby be considered as very low as the plant is located in Iceland with cheap, geothermal energy. In addition, several constraints have to be held during operation. These are given by

$V_1 \leq 4$ mol/s
$V_2 \leq 8$ mol/s
$x_{B_1,B} \geq 95\%$ B
$x_{B_2,D} \geq 98\%$ D
$M_{R_1} \leq 2000$ kmol
$M_{R_2} \leq 3000$ kmol

Do not change any of the controllers (FIC and PIC) drawn in the flowsheet!

a) Give the number of dynamic and steady state degree of freedom.
b) Which operational constraints will be most likely active based on your engineering experience? Reason your answer.
c) Add the missing controllers in the flowsheet on the following page such that you obtain a consistent control structure for the active constraints and inventory control. If you have any degrees of freedom left, propose a possible way to utilize them.
d) What is snowballing? Can it occur in this system and if yes, where can it occur?
e) What is the control structure controlling the feed flow rate to Reactor 2 (F_2) called?

a) Dynamic DOF = 14 (all the valves that can be manipulated)
Steadystate DOF = $N_{valves} - N_{0y} - N_{specs}$
$N_{valves} = 14$ (total number of valves)
$N_{0y} = 4$ (the inlet flow is given, 2 pressure controllers, ratio control)
+ 6 (level controllers for reactors, condensers and reboilers)
Steady state DOF = ~~14 - 4 - 6~~ = 4 (left for active constraints and soc)

b) Active constraints: are the most valuable product, x_{B2} and the reboiler flows, V_1 and V_2, since the energy is cheap.

c) Variables left for SOC

$= \#$ steady state DOF $- 3$(active constraints)

$= 4 - 3 = 1$

Pairing for MR_1 and MR_2 = pair close

Pairing for X_{B2} = Pair close

SOC = Use SOC to controll the amount of $X_{A,S}$ sent into the distillation column.

d) Snowballing can occur in recycle loops when all the flows are used for inventory control. If the total throughput is increased, the recycle streams might build up.

In this system, we have 2 recycle loops where snowballing ~~where snowballing can~~ occur.

e) The control structure controlling F_2 is called ratio control. The set point gives the optimal ratio F_2 / B_1.

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