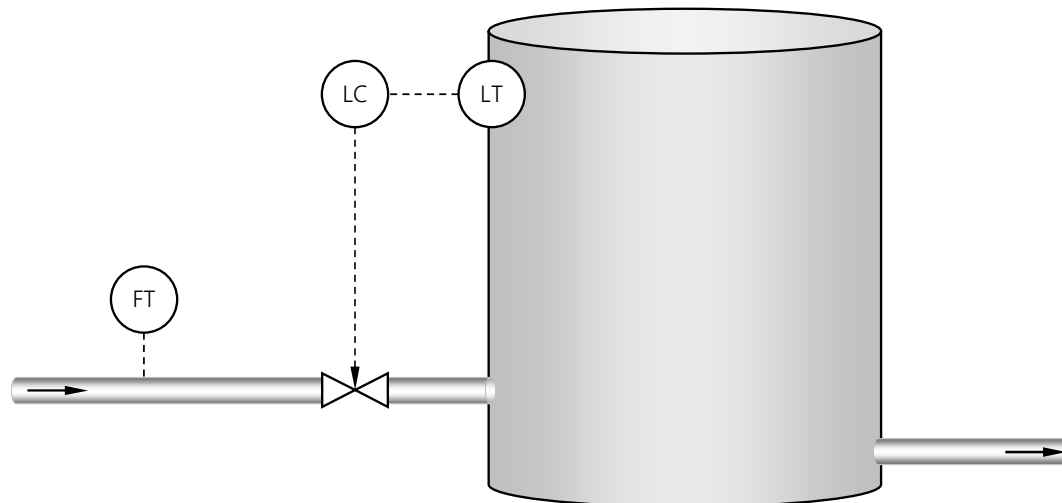


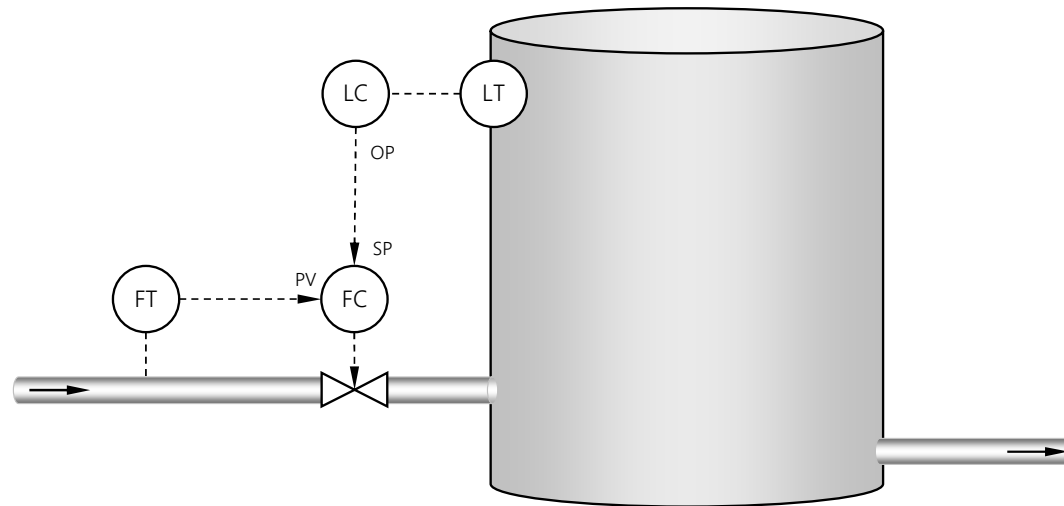
Cascade control

Example: Level control with improvement potential

- It is important to keep this level close to its setpoint.
- The tank level varies too much, due to large pressure variations in the feed line.
- We can't make the controller more aggressive, because then it becomes unstable.
- Can we still improve control performance?

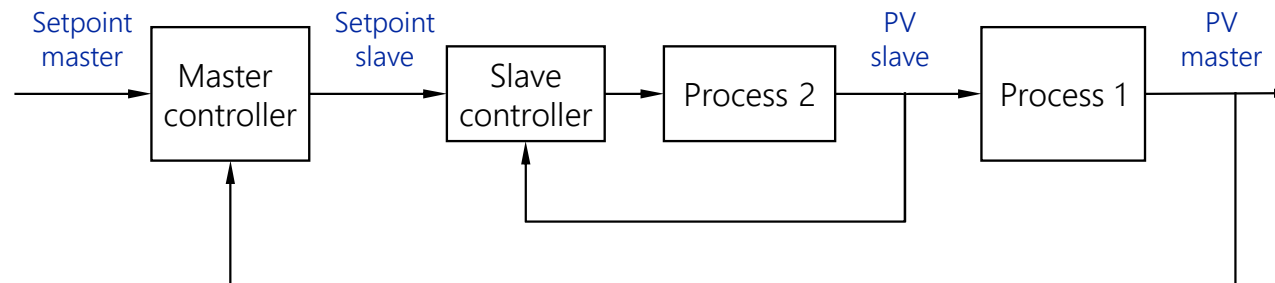


Solution: Control the flow too!



Cascade control: One controller provides the SP to another

- The control signal of the master controller (u_1) is the setpoint of the slave controller (r_2).

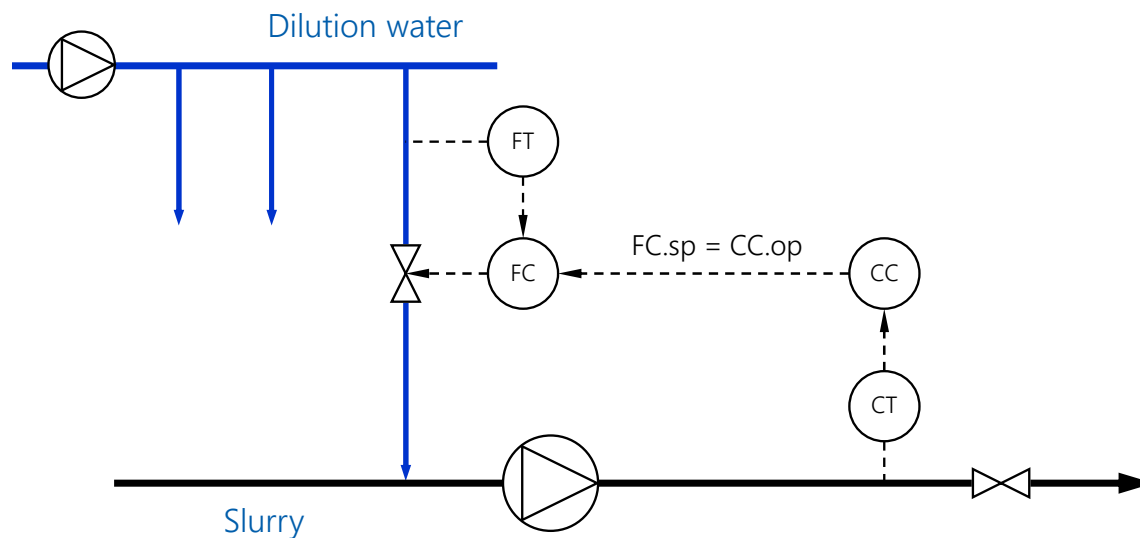


Typical examples for cascade control

Master	Slave
Level	Incoming or outgoing flow
Temperature HEX	Cooling water flow
Concentration	Dilution water flow
Column temperature	Reflux flow
Evaporator density	Steam flow
pH	Caustic addition

Cascade control: Slave disturbance “paradox”

- Scenario: Concentration control (master) with dilution water as slave
 - A decrease in water header pressure causes the flow of dilution water to decrease.
 - Outline how dilution water flow and concentration vary over time.
The master and slave controller are both active.
- Hint: Start by drawing in the middle graph



Give the answer as 4 trends

Conc

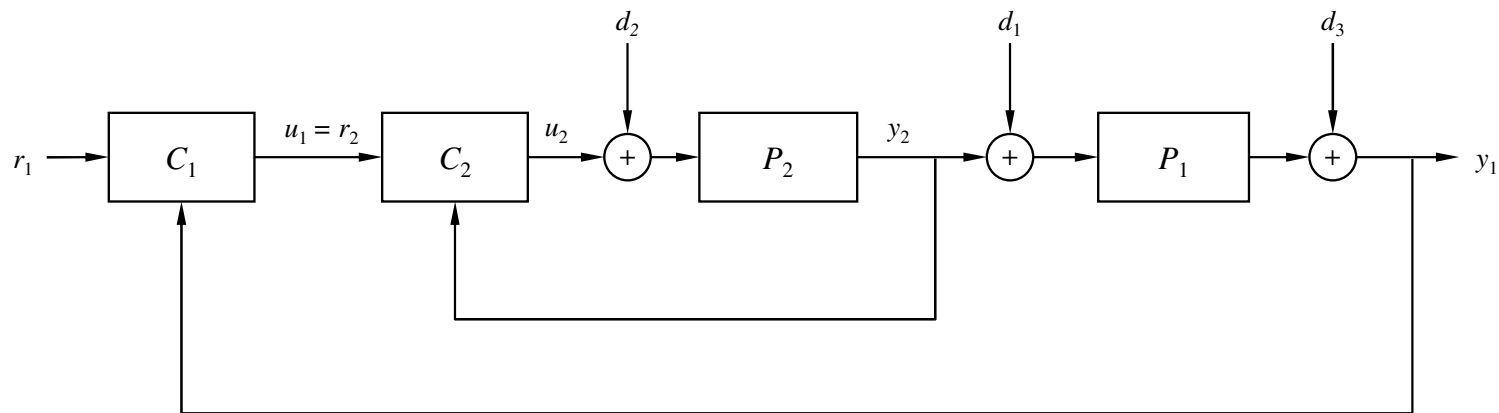
Dilution water, SP and PV

Valve position

Question: Where does the disturbance enter?

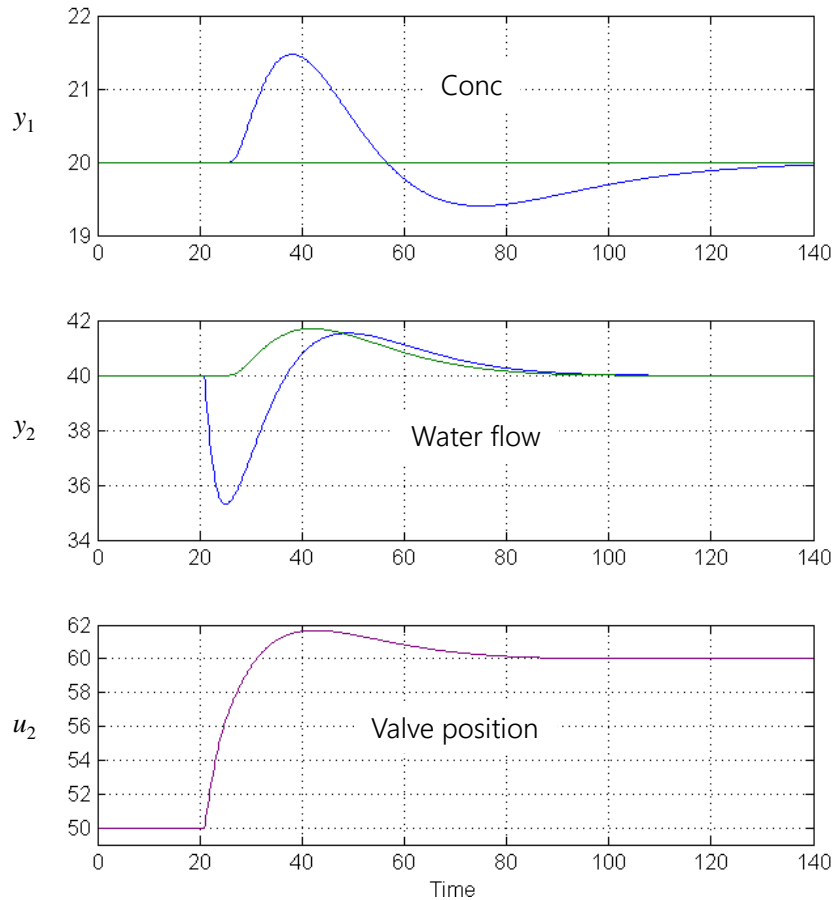
In the above scenario: Which is the best model for pressure disturbances?

Do they enter as d_1 , d_2 or d_3 in the block diagram below?



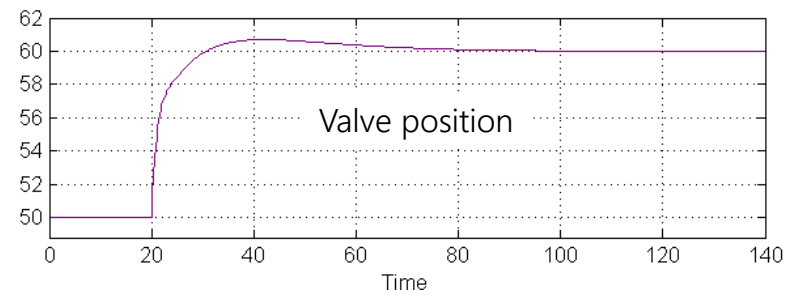
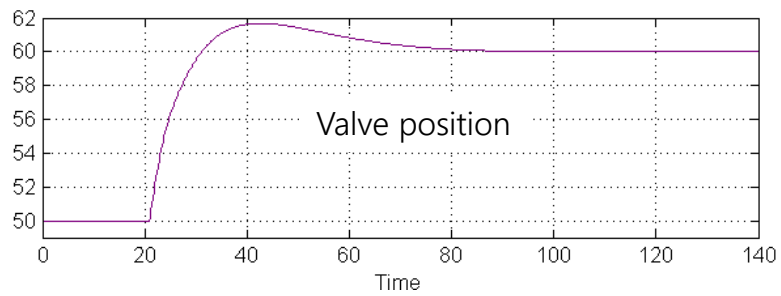
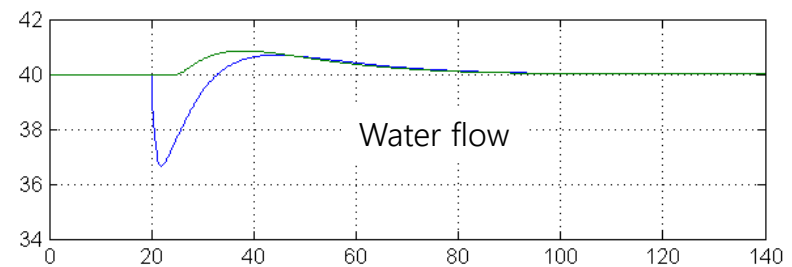
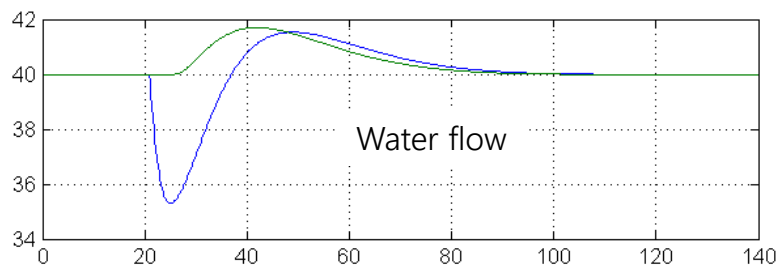
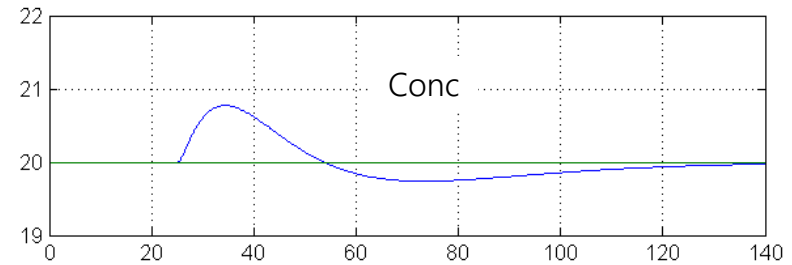
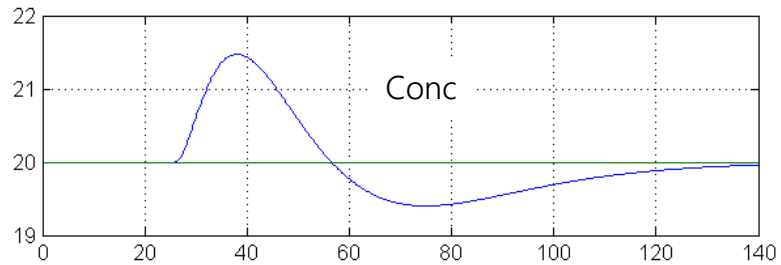
Answer: d_2

Solution: S-shaped disturbance in concentration



- The decrease in water flow affects concentration, even though the flow controller is active.
- The master controller acts on this, changing the water flow SP.
- In some sense this is unnecessary. This disturbance should be handled by the slave controller.

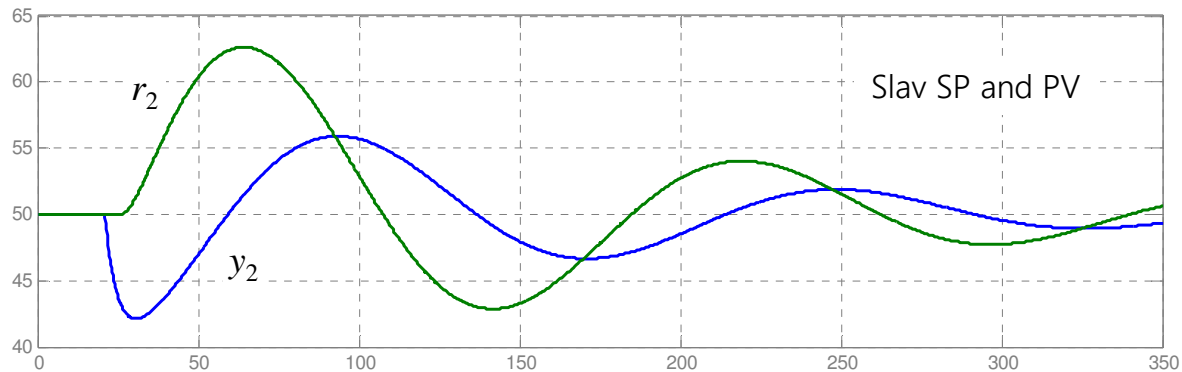
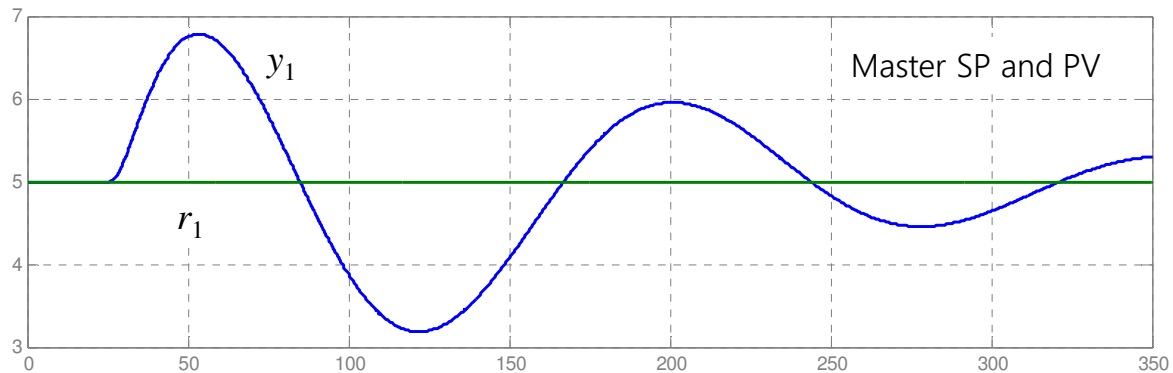
Conclusion: Important that the slave is much faster than the master



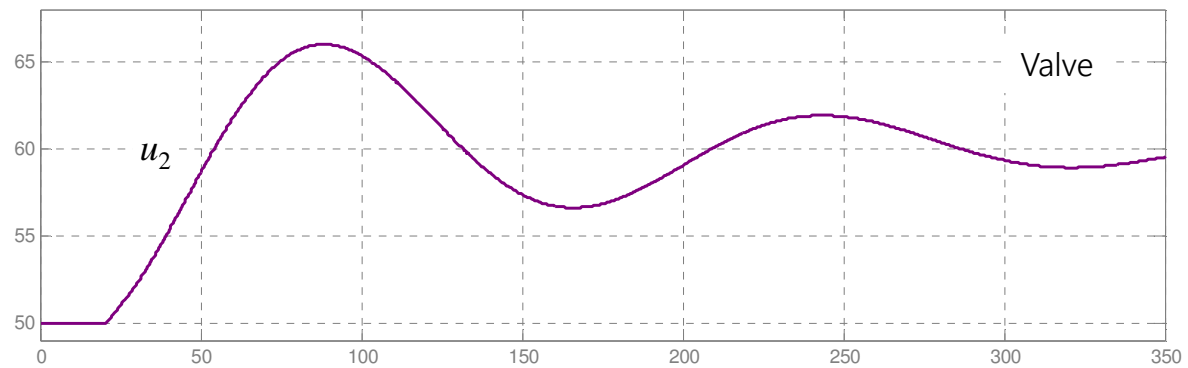
Here somebody tried to make the master faster than the slave.

The disturbance is thrown back and forth between master and slave

On the borderline to unstable.



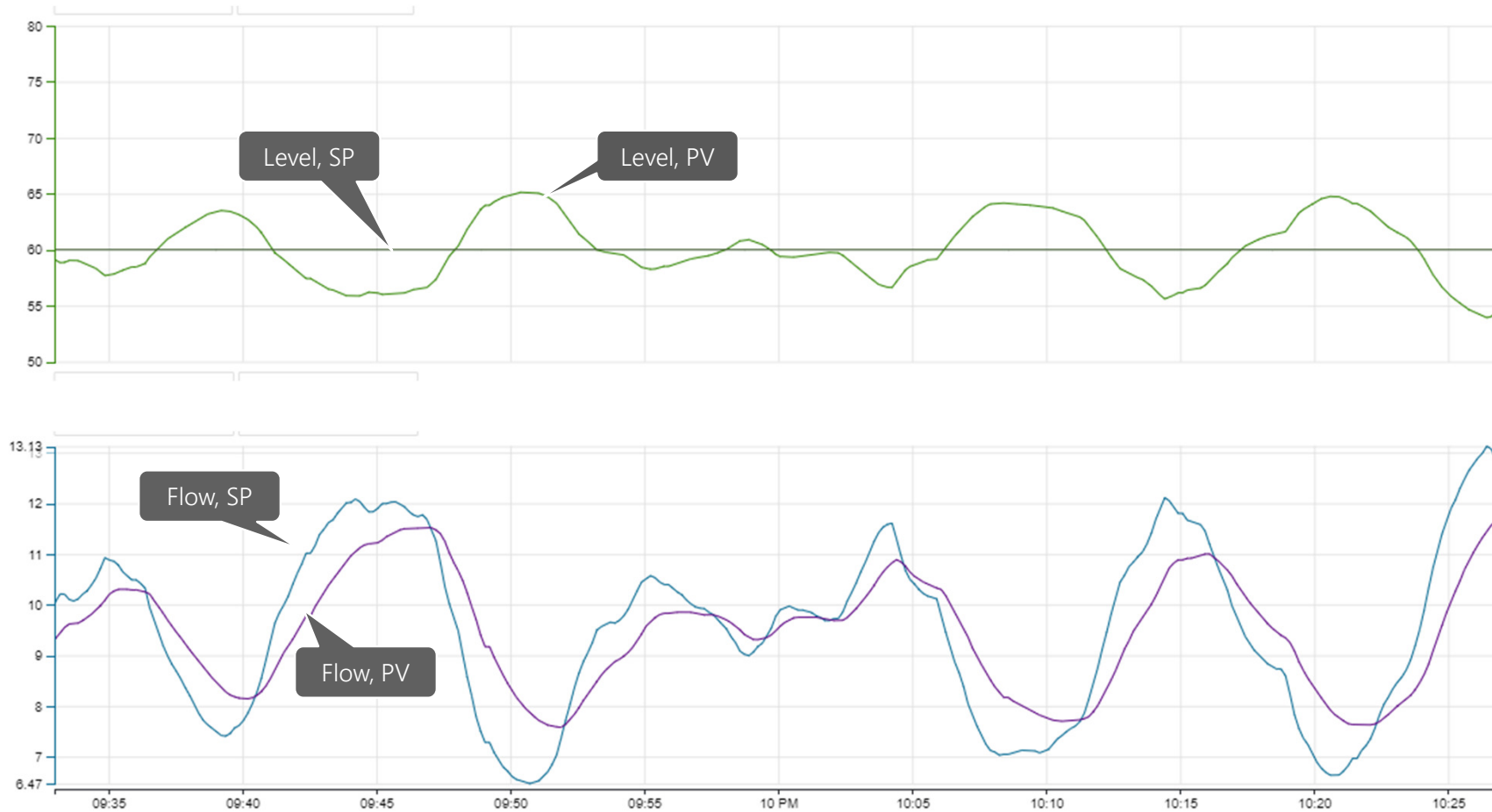
" $\lambda_1=15$ $\lambda_2=40$ "



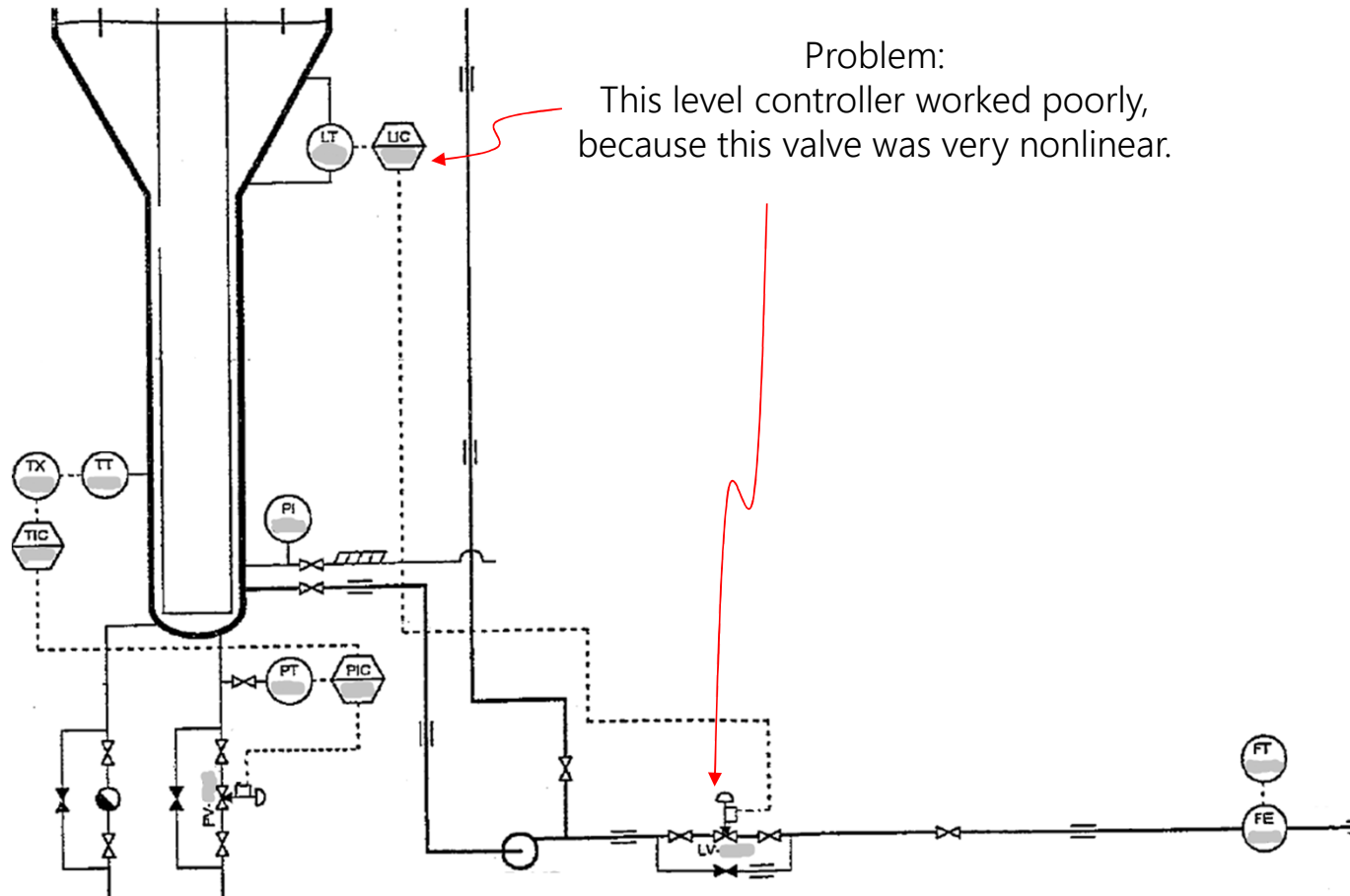
When to use cascade control?

- The slave loop dynamics should be considerably faster than the master loop dynamics.
- The motivation for cascade control is in the slave controller. It should handle disturbances entering before process 2.
- Can cascade control be motivated even if you don't have disturbances at the lowest level?
 - The answer is "yes", and we will soon see why.

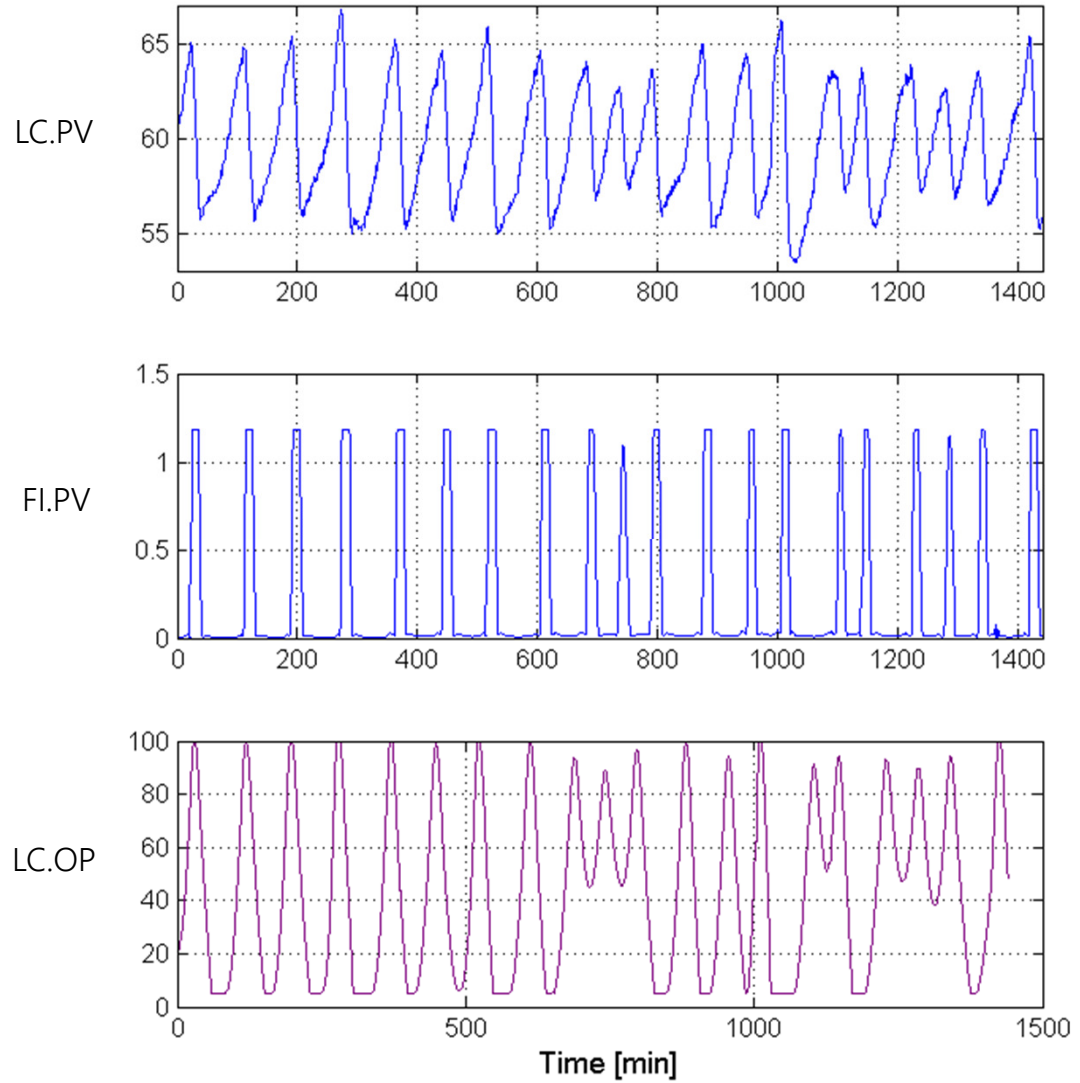
Real example of poor frequency separation



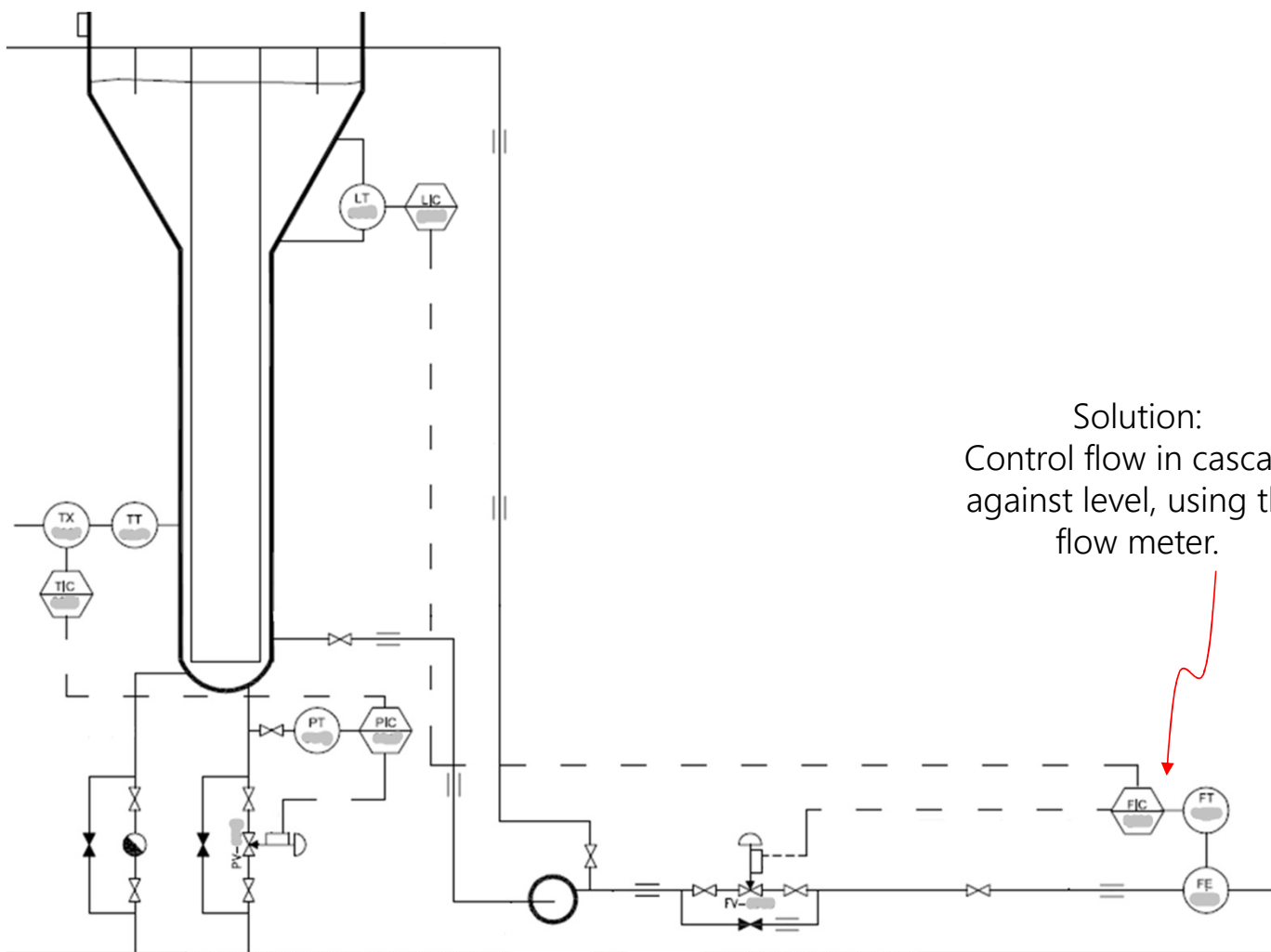
Evaporator: Improvement opportunity



Before improvement:



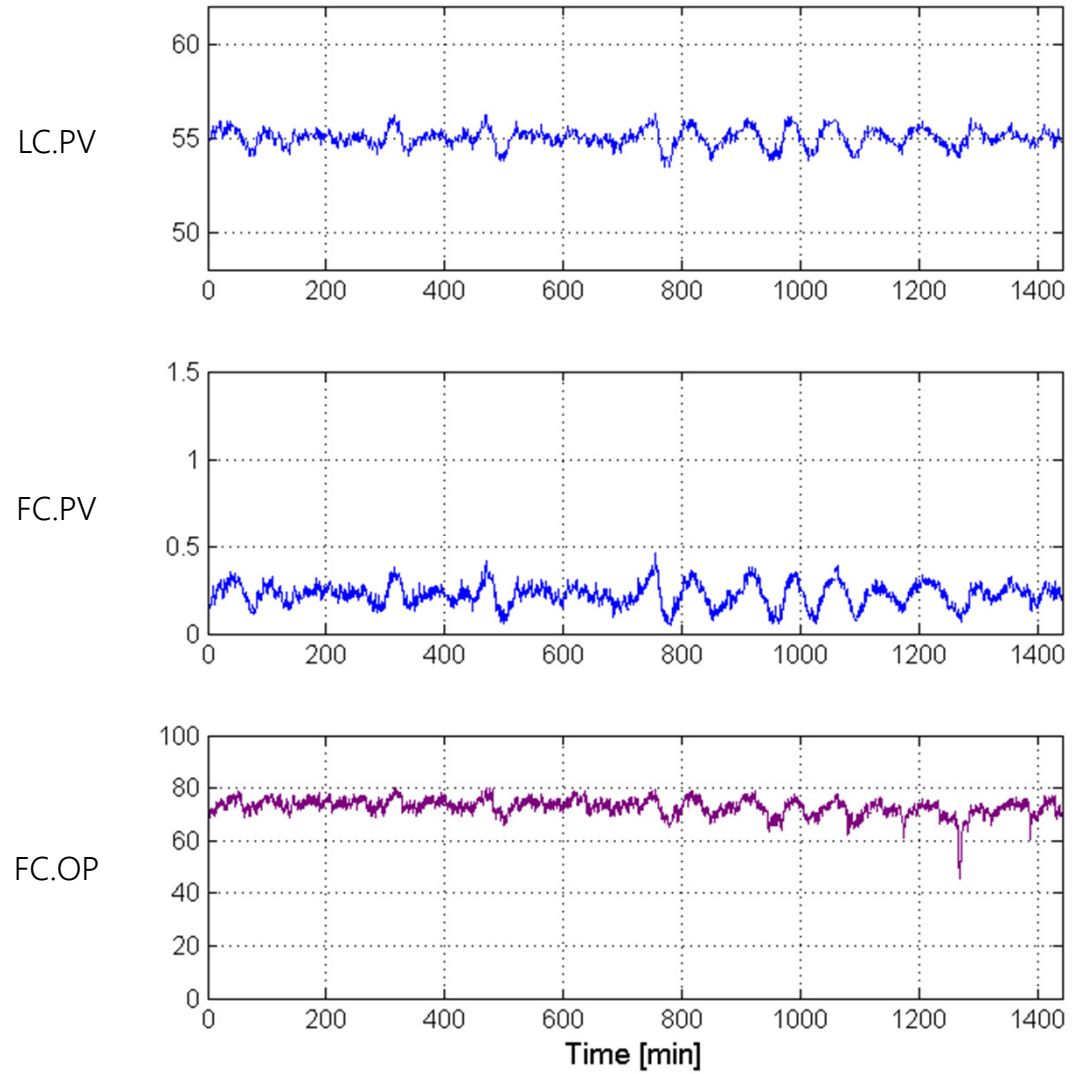
Very nonlinear valve



Solution:
Control flow in cascade
against level, using this
flow meter.

Cascade control eliminates the effect of the valve nonlinearity.

No more level cycling.



Cascade control linearizes

- Why did the slave process become linear in the example above?
- The slave process is a control loop. From SP to PV.
- In a closed loop the gain (from SP to PV) is always 1.
 - In steady state $SP = PV$, if the controller has integral action.

Additional advantage:

- In some applications it's convenient for the operator to be able to run the flow in Auto, with setpoint, instead of manipulating the control valve.

Policy: Motivate all additional complexity

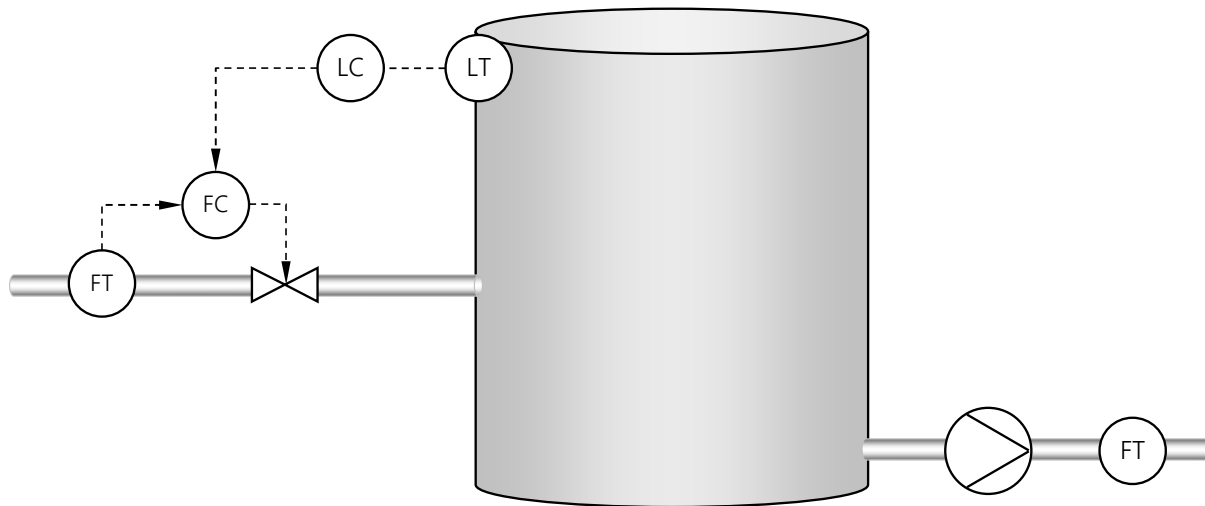
General principle:

- If you are using something else than a simple feedback you should motivate why SISO is not enough.
- In the first case above, pressure disturbances were that motivation.

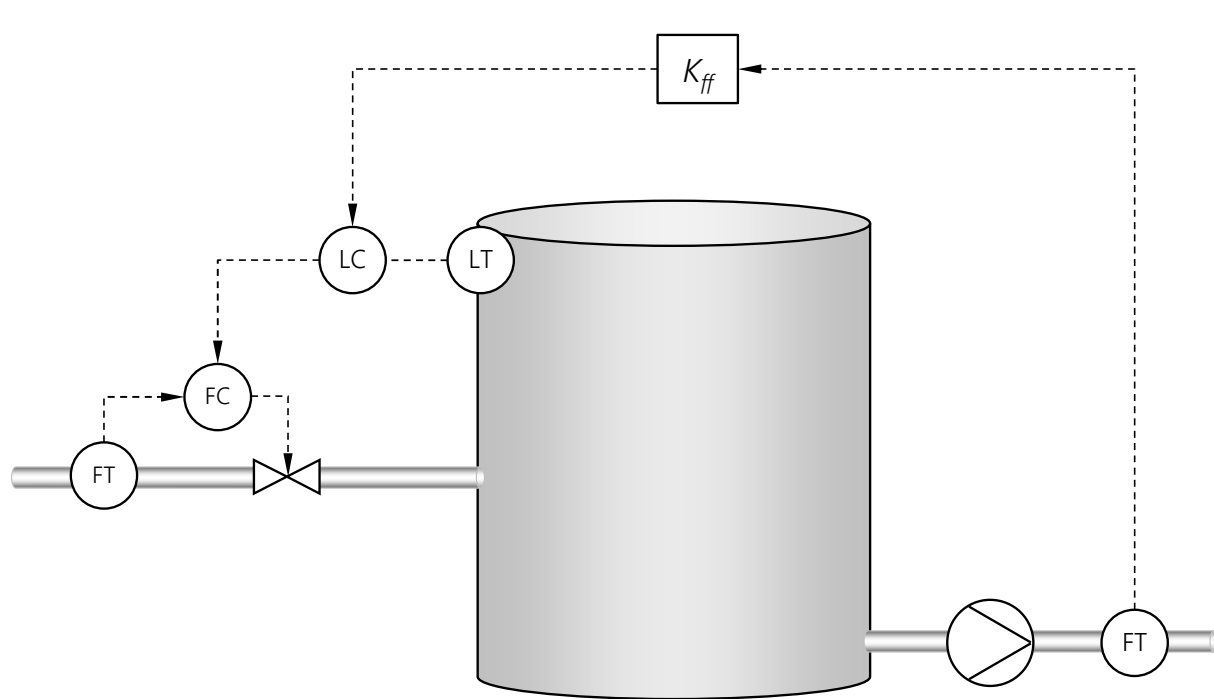
Feedforward

Ex: Level control with improvement potential

- It is important to keep this level close to its setpoint.
- The tank level varies too much, because of large variations in the outgoing flow.
- The LC cannot be tuned more aggressively: then it becomes unstable.
- Can we still improve control performance?



Feedforward: Warn the controller "in advance"



K_{ff}

= multiplication with a factor K_{ff}

In this case probably $K_{ff} = 1$

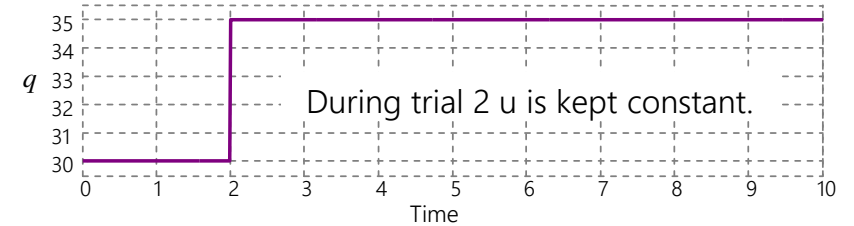
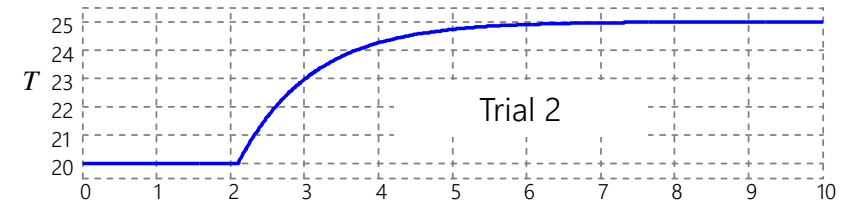
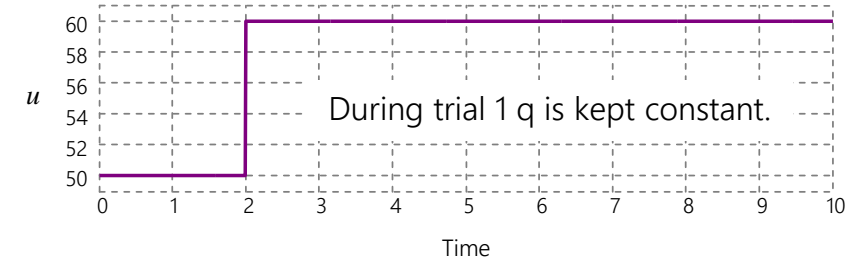
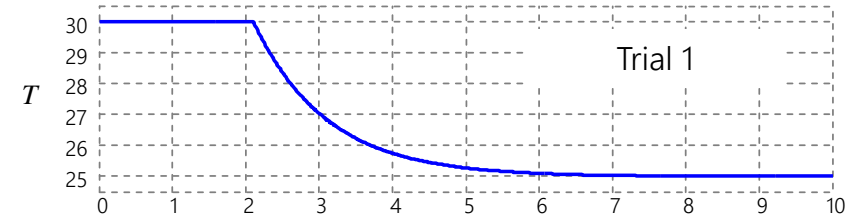
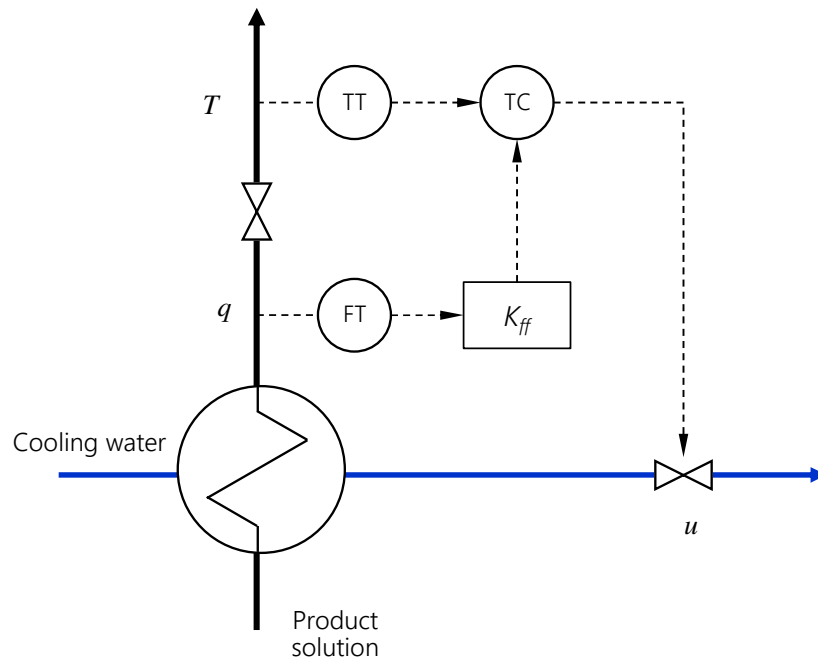
In words:

If the outgoing flow increases by 5 m³/h we should immediately increase the incoming flow by the same amount.

We don't need to wait for the level to deviate from its setpoint.

Exercise: FF gain for HEX

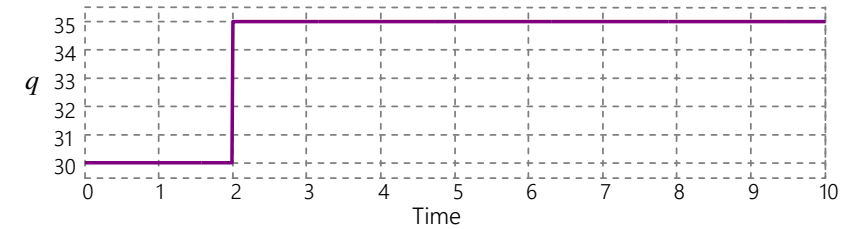
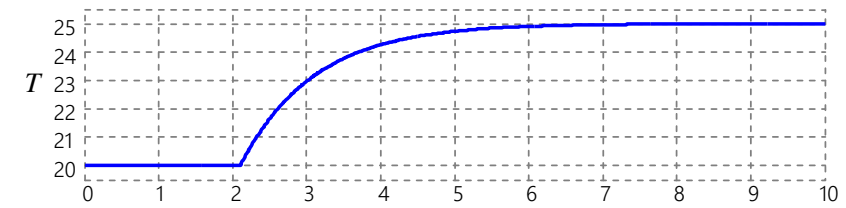
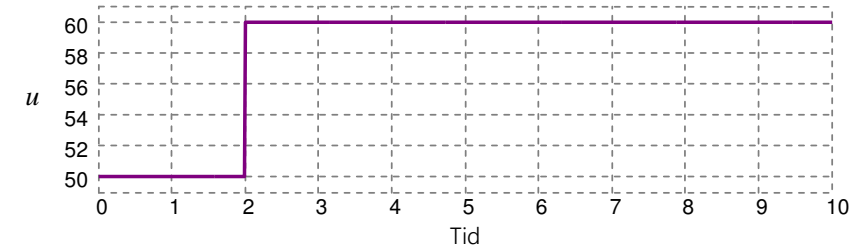
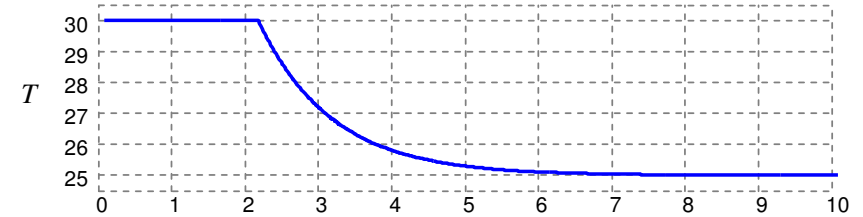
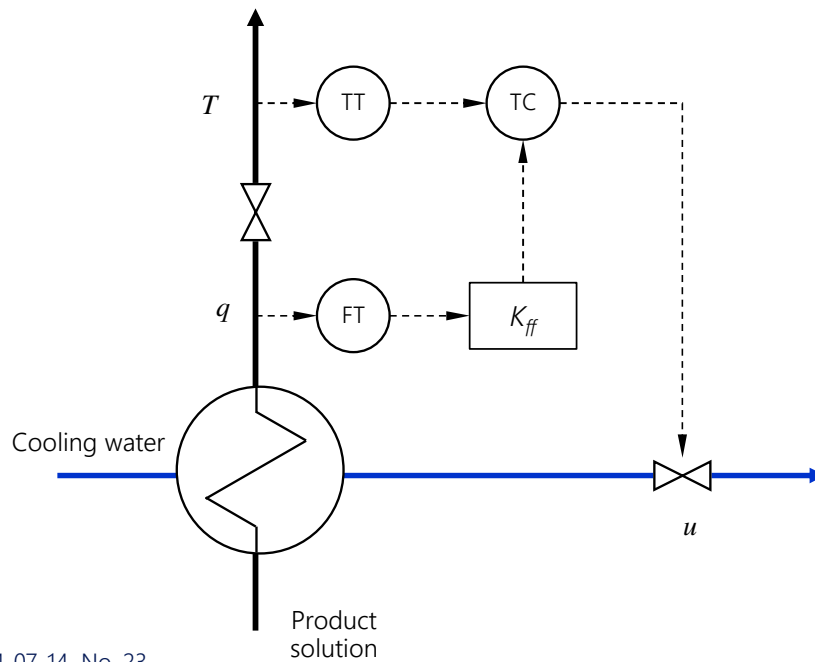
Determine the correct feedforward gain.
Use data from the two trials showed to the right.



Solution

The effect of q on temperature is "twice as strong" as the effect of u :
 5 units in q has the same effect as 10 units on u , but in the opposite direction.

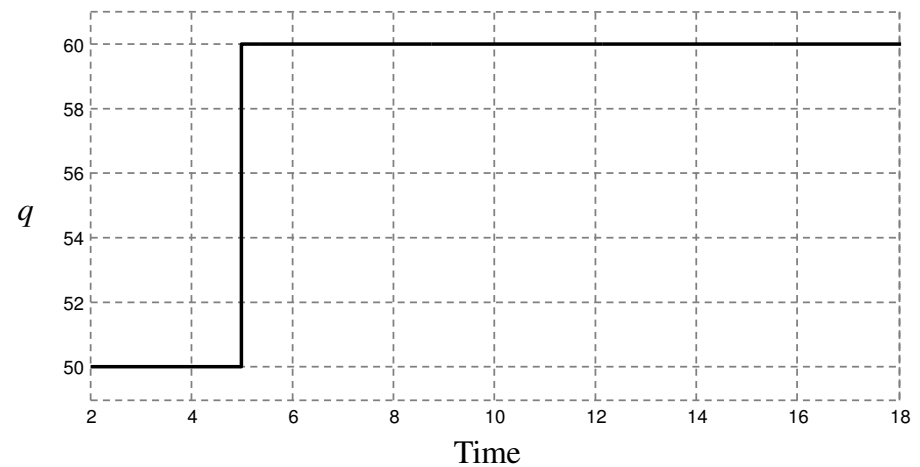
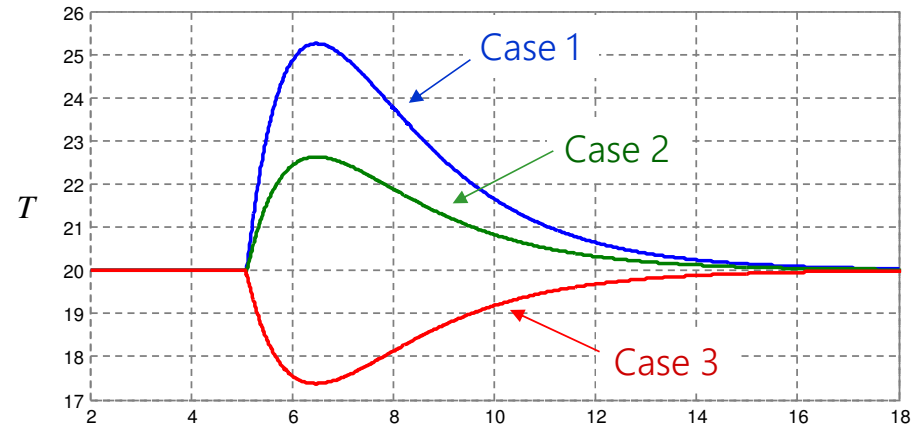
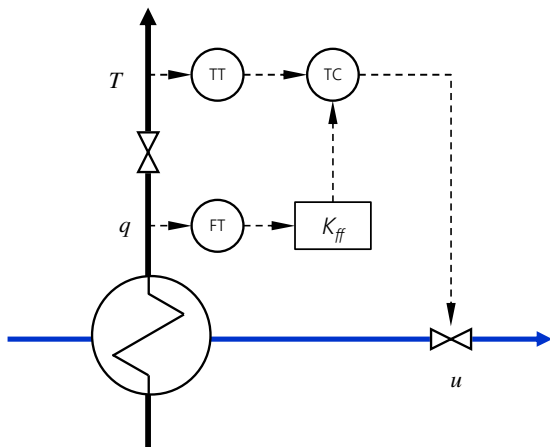
Thus: To counteract a disturbance from q , u has to change twice as much as q .
 So: $K_{ff} = 2$



Exercise: Wrong K_{ff} for HEX in the example above

Which behavior corresponds to

- $K_{ff} = 0$ Case 1
- $K_{ff} = 1$ Case 2
- $K_{ff} = 3$ Case 3



Conclusion: Feedforward is not robust

- Feedback control: If the process model used for tuning is not correct, the feedback algorithm compensates for that.
- Feedforward control: A model error can cause more damage:
 - If the FF gain is completely wrong, the FF can do more harm than good.

Use a "chicken factor"



Feedforward exercise

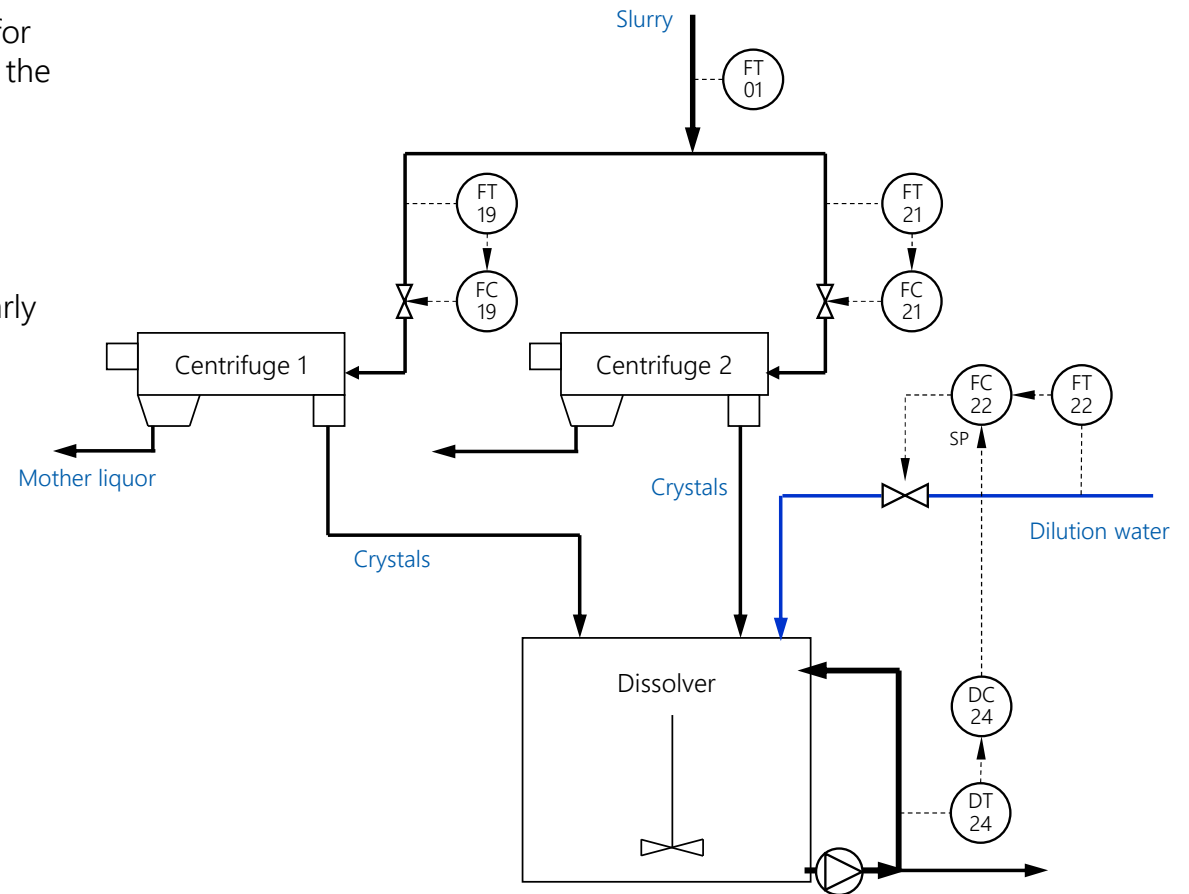
Density control in a dissolver

Problem: Every shift one of the centrifuges is shut off for cleaning, during one hour. Then the flow of crystals to the dissolver decreases by 50% (but the solids flow is not measured).

Consequence: a large upset in dissolver density.

The density controller DC24 cannot be tuned particularly aggressively because the process dynamics is slow.

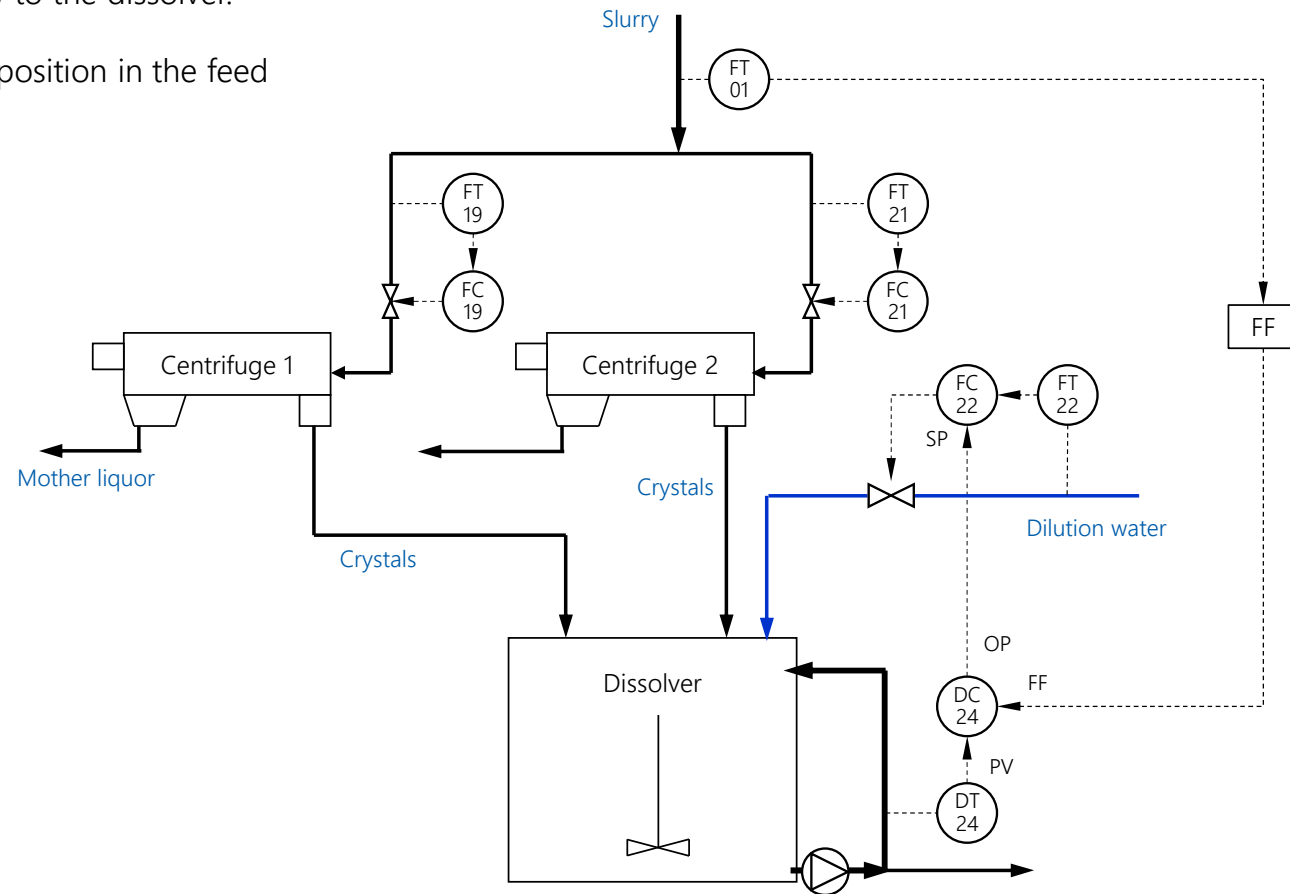
How can we improve density control?



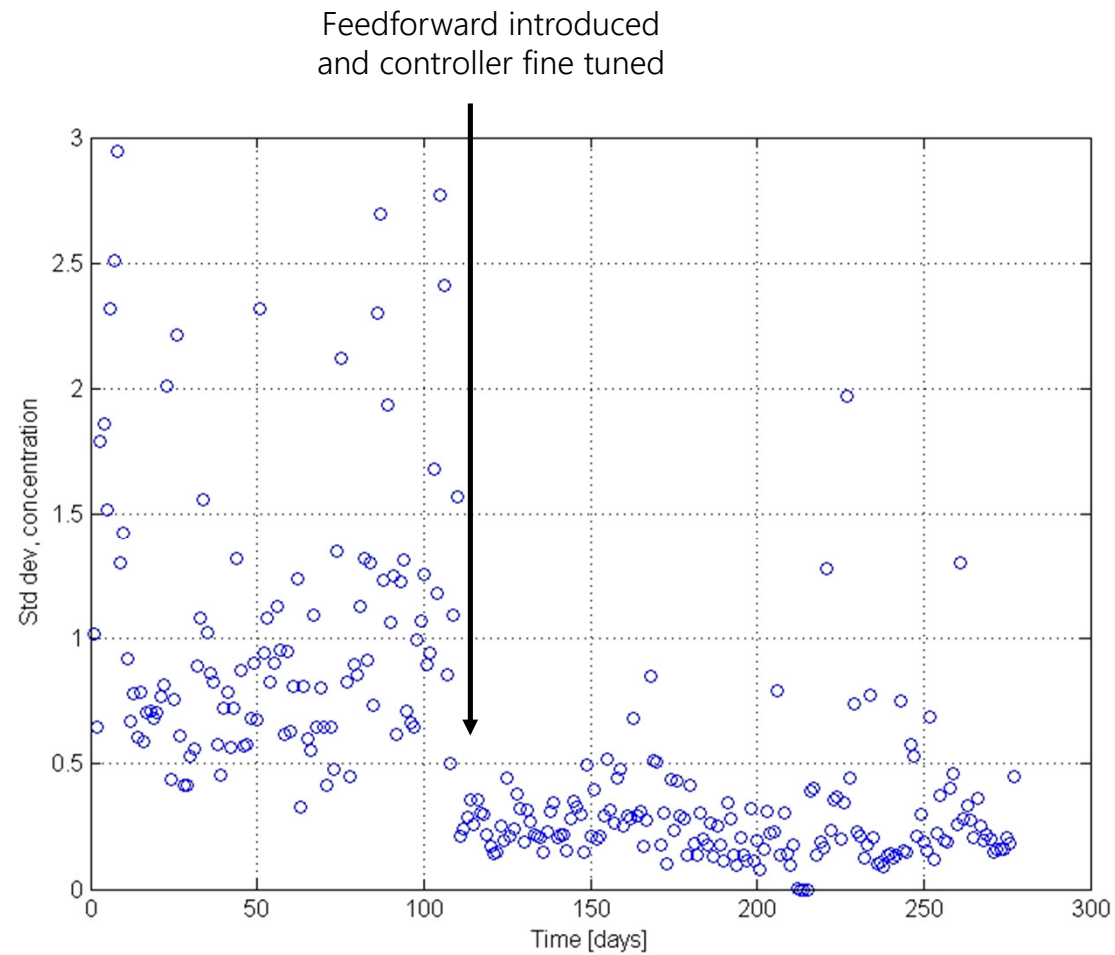
Solution

Introduce a feedforward from centrifuges total feed to the density controller.
A change in feed corresponds to a change in solids flow to the dissolver.

The feedforward gain will not be perfect, since the composition in the feed slurry varies. But in practice it works very well.

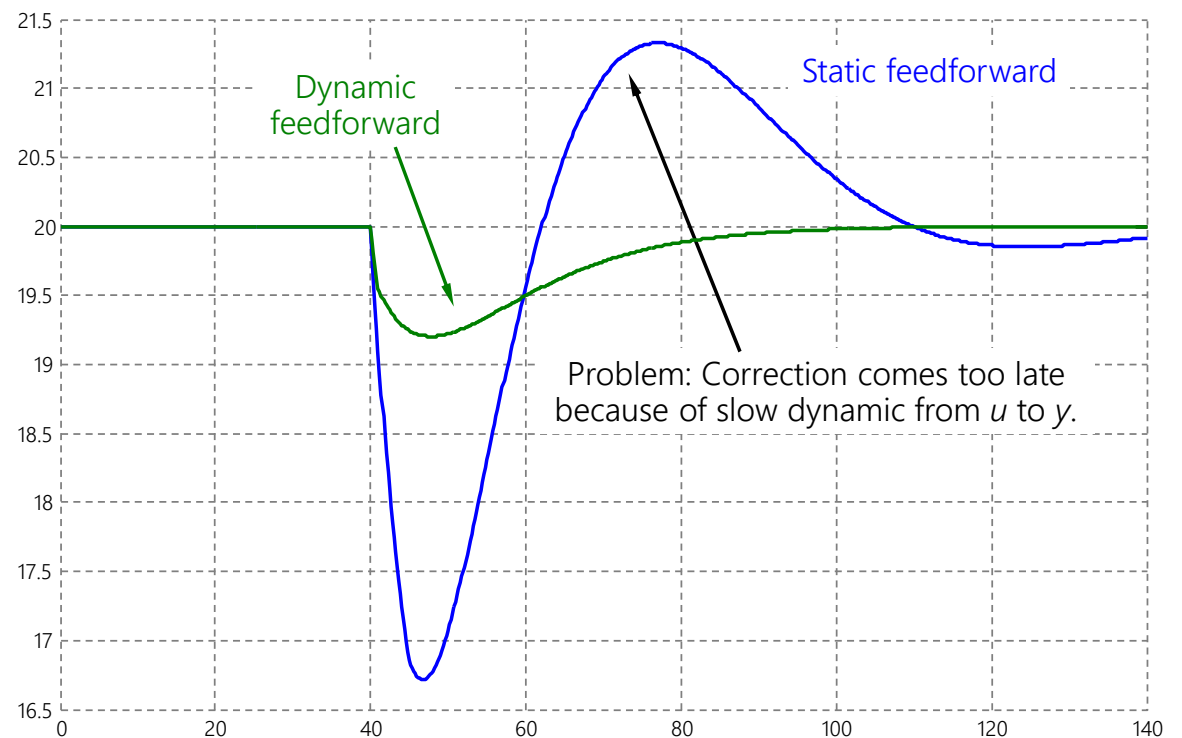


Feedforward and tuning reduced variations



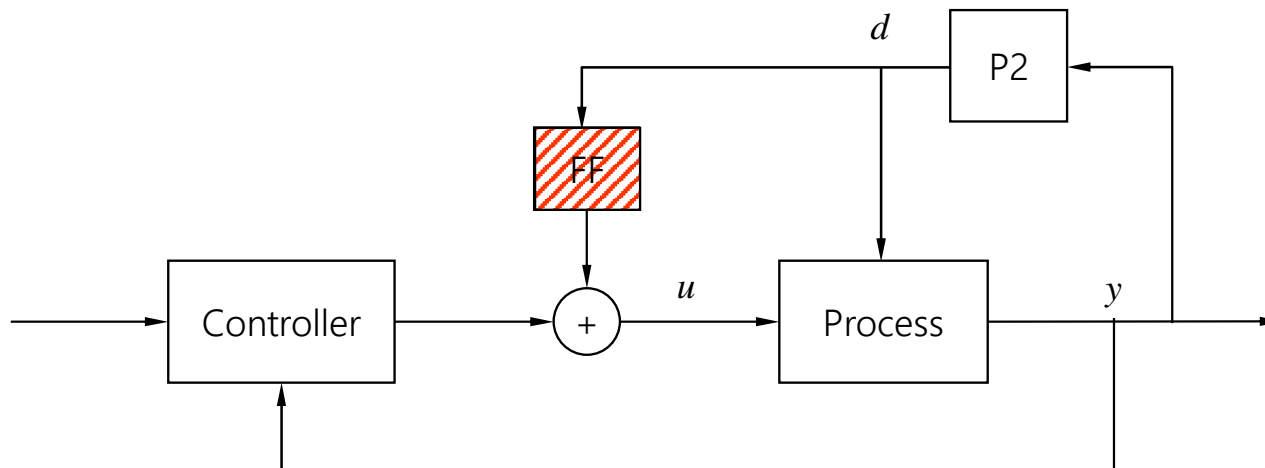
Careful if the two dynamics differ

- In some cases we have control loop where the disturbance effect on PV has a different dynamics than the effect from controller output.
- Then it may be a bad idea to introduce a static feedforward. Why?

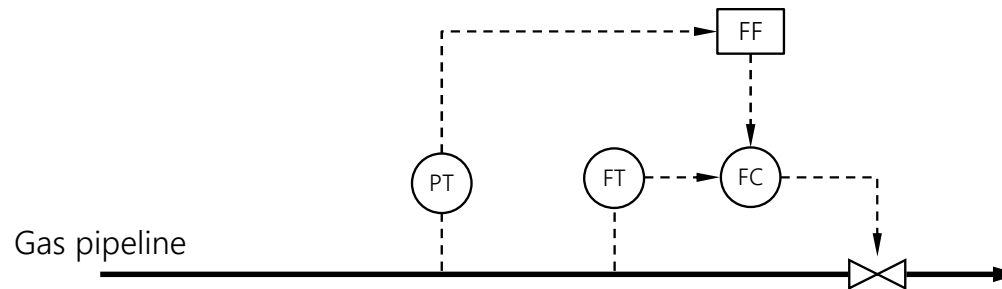


Avoid unintentional feedback!

- If the disturbance variable depends on the manipulated variable you have to be careful.
 - If you try to feed forward you introduce an extra feedback!
- d is not an external variable
- It may work, but the previous analyses do not apply.
- Normally a feedforward cannot make the loop unstable, but in this case it may.



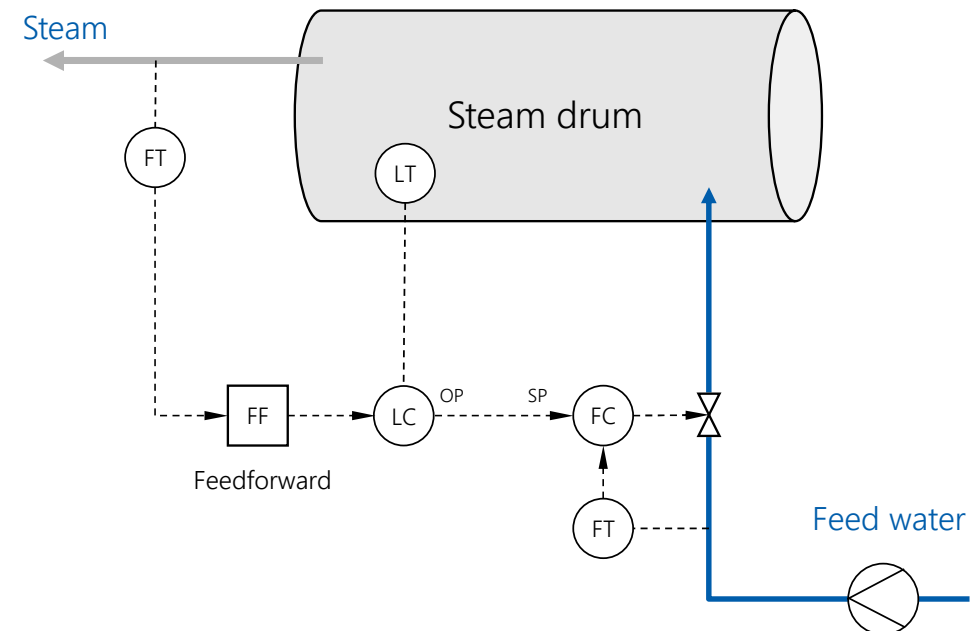
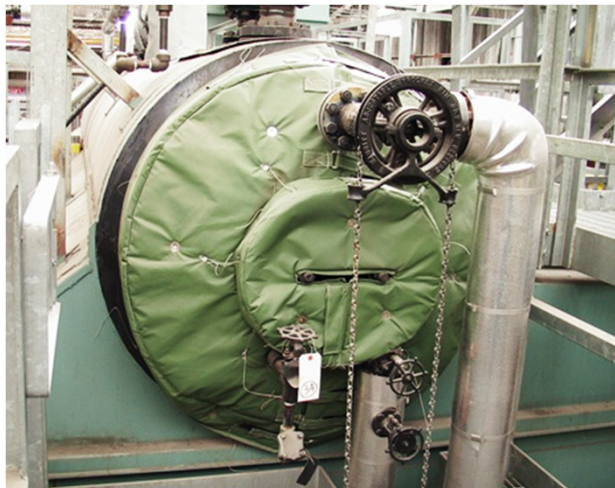
Example of unintentional feedback



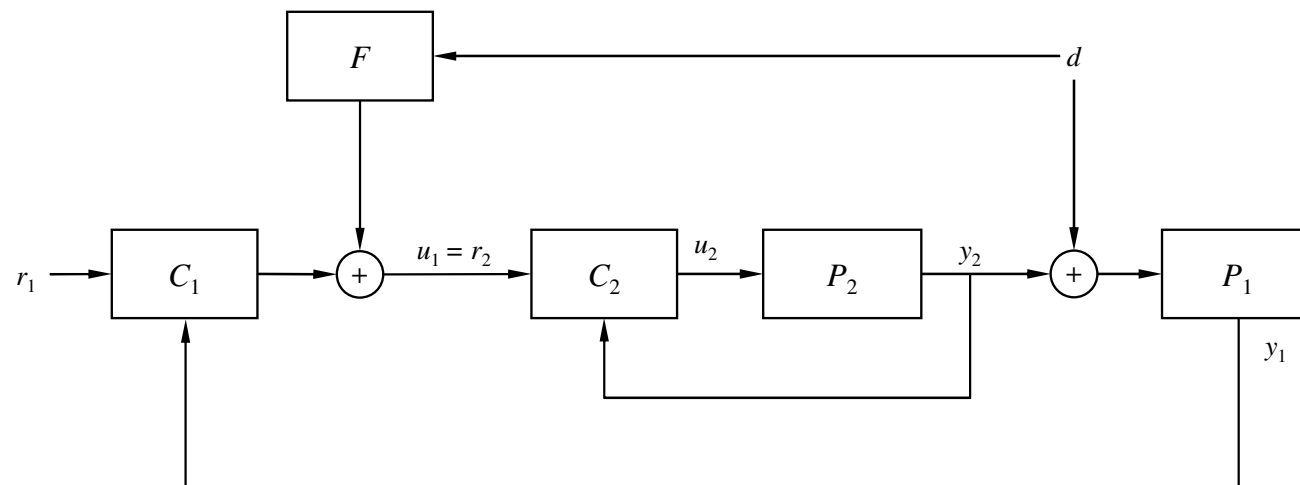
If this is a large consumer, the valve affects both flow and pressure.
Then the FF-block is not a normal feedforward, but an extra feedback.

Drum level example: FF + cascade

- A common way of controlling the drum level in a steam boiler is to combine feedforward and cascade control.
- The LC gets a feedforward from steam consumption. Feedwater flow in cascade against level.



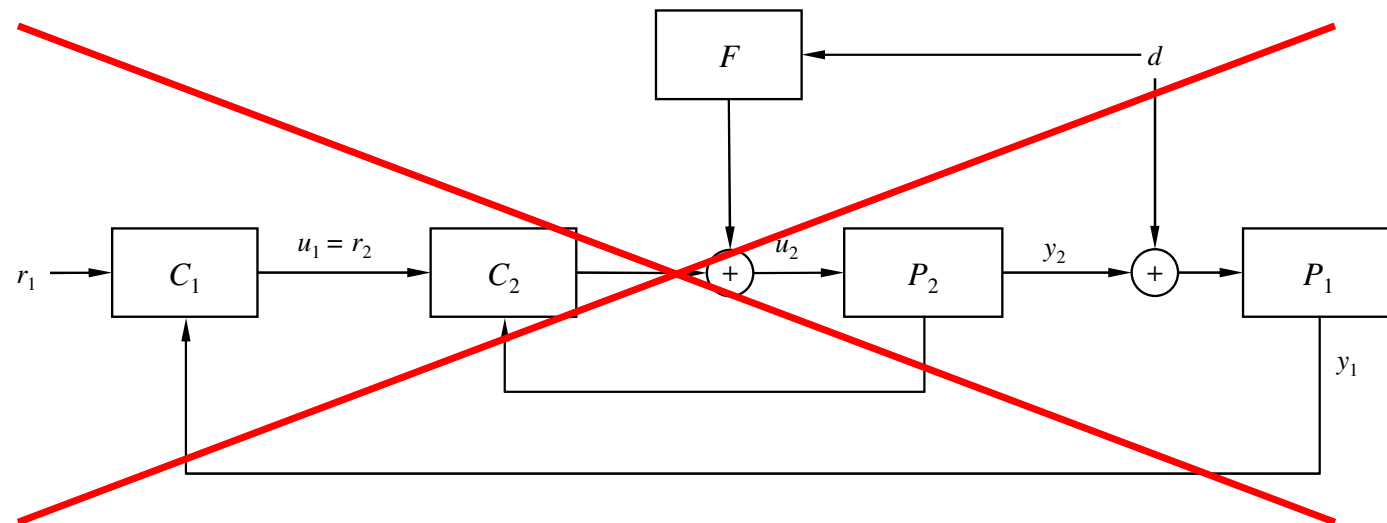
FF and cascade: Block diagram



Comments on the FF/Cas combination

- In fact we already saw several examples of this combo.
 - The first example with feedforward for tank level, the example with dissolver density control, etc
- If you are not careful, the implementation can go wrong.
Below is a common mistake, feeding forward to the slave controller instead of the master.
 - This is completely pointless: d doesn't even affect y_2 .
The FF term will be immediately counteracted by the slave controller integral part.

Warning: DON'T do like this!

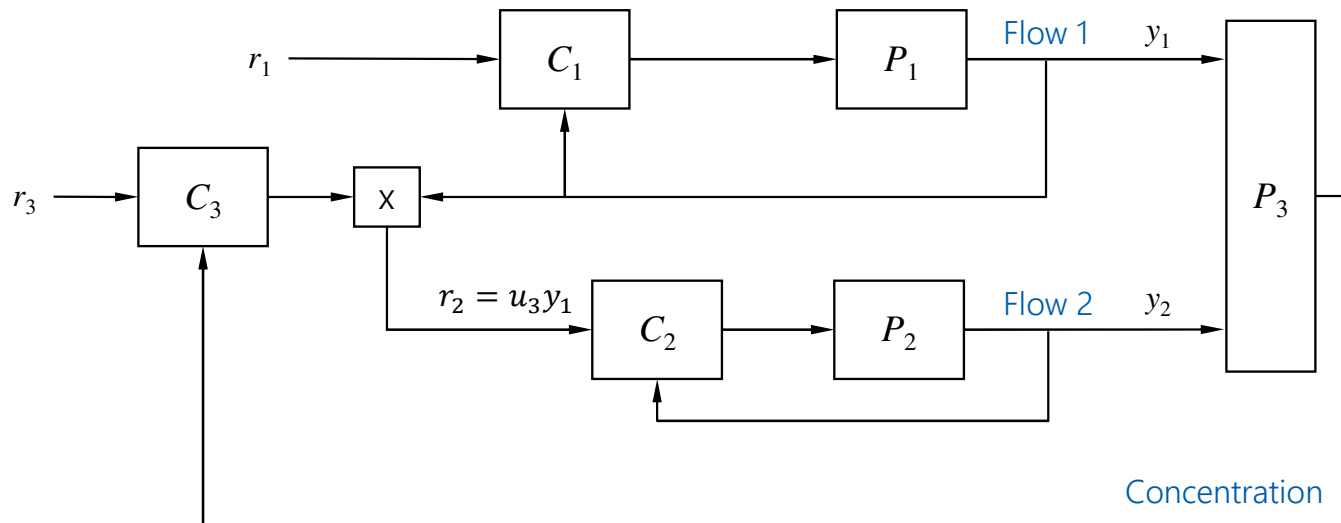


Ratio control

Ratio in cascade

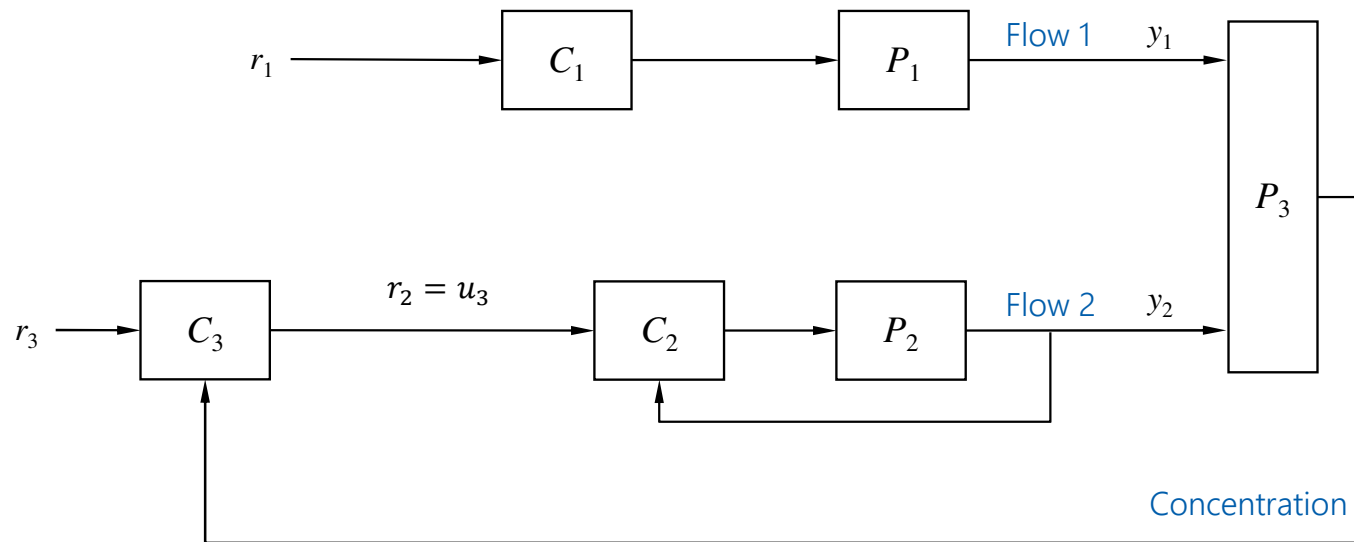
Typical examples: reactor, mixing, dilution

This is an important and useful structure that is not well known by practitioners.

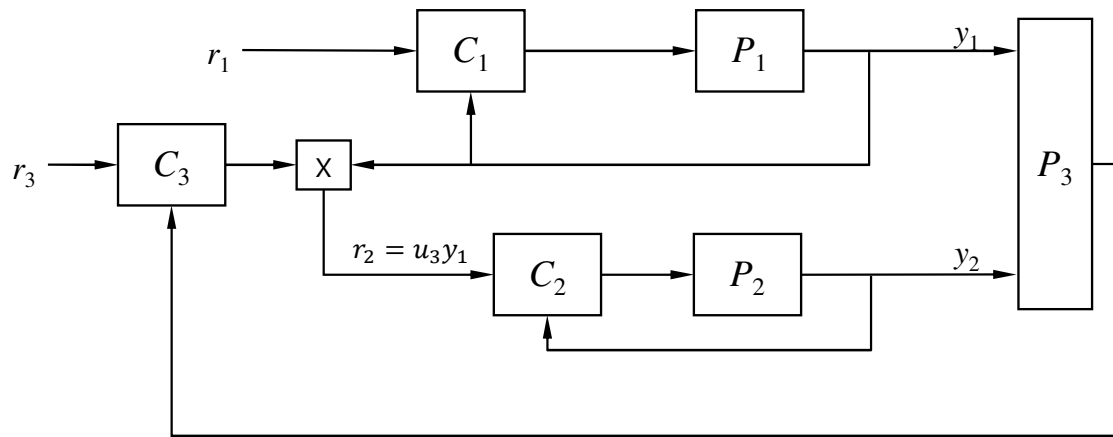


Cascade control, for comparison

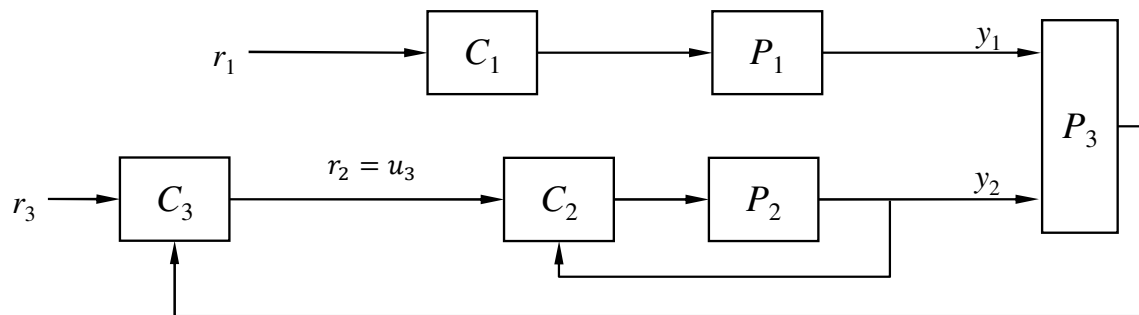
This structure is an alternative solution to the same problem.
Exercise: In which way is this structure inferior to "ratio in cascade"?



Compare structures



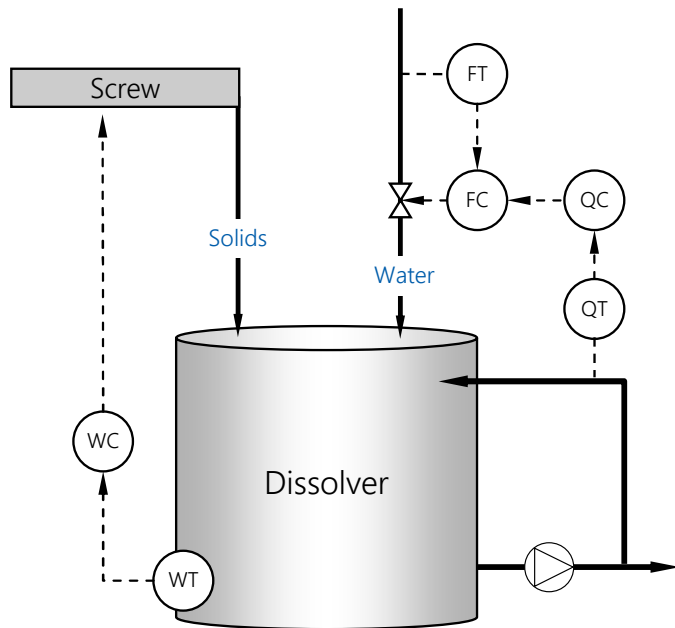
Exercise: In which way is cascade control inferior to "ratio in cascade"?



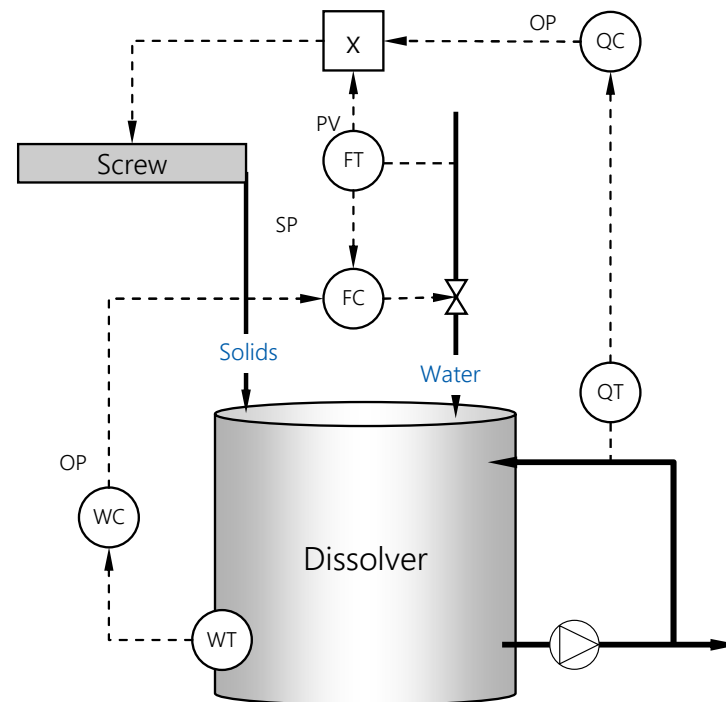
Answer: With ratio control any change in the master flow immediately changes the secondary flow SP, without going through P_3 .

Ratio in cascade reduces interaction

Direct cascade control:
Concentration and level controllers
disturb each other



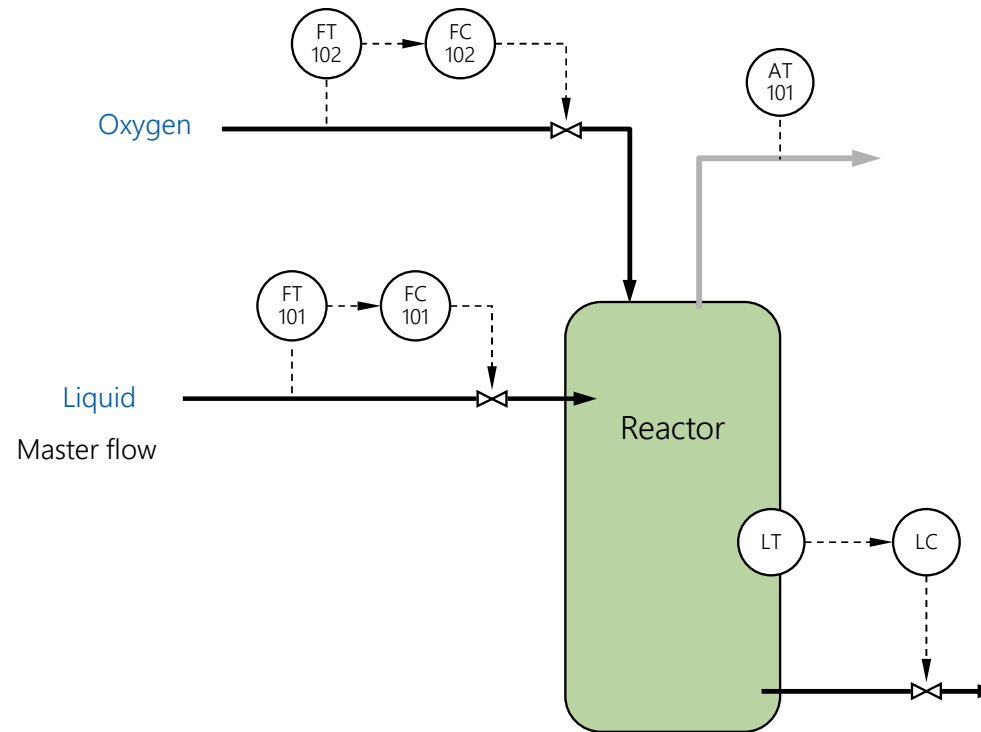
Ratio in cascade:
Level controller does not disturb
concentration controller



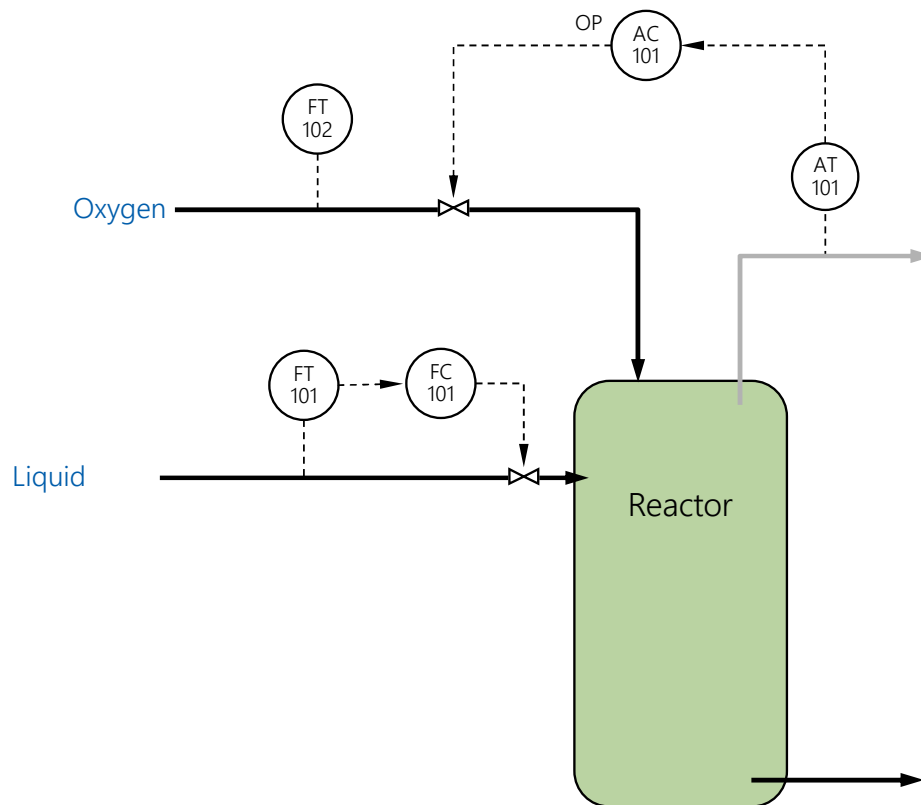
This is an example of **decoupling**

Exercise: Suggest control structures

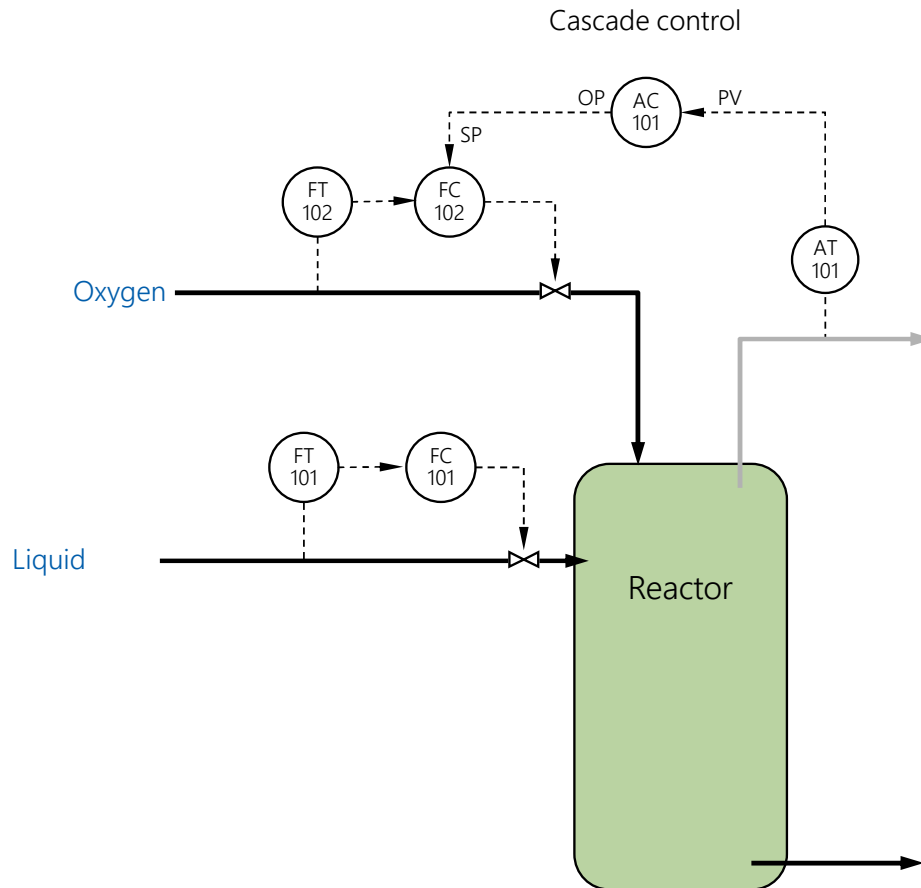
- Process: Two phase reactor (liquid – oxygen); oxidation
- Oxygen excess in the reactor gas is measured (AT-101), and should be controlled.
- Suggest different ways of controlling the O₂ content, and compare them. Which solution is superior in which situation?
 - Don't care about the level control.
 - Liquid is master flow.



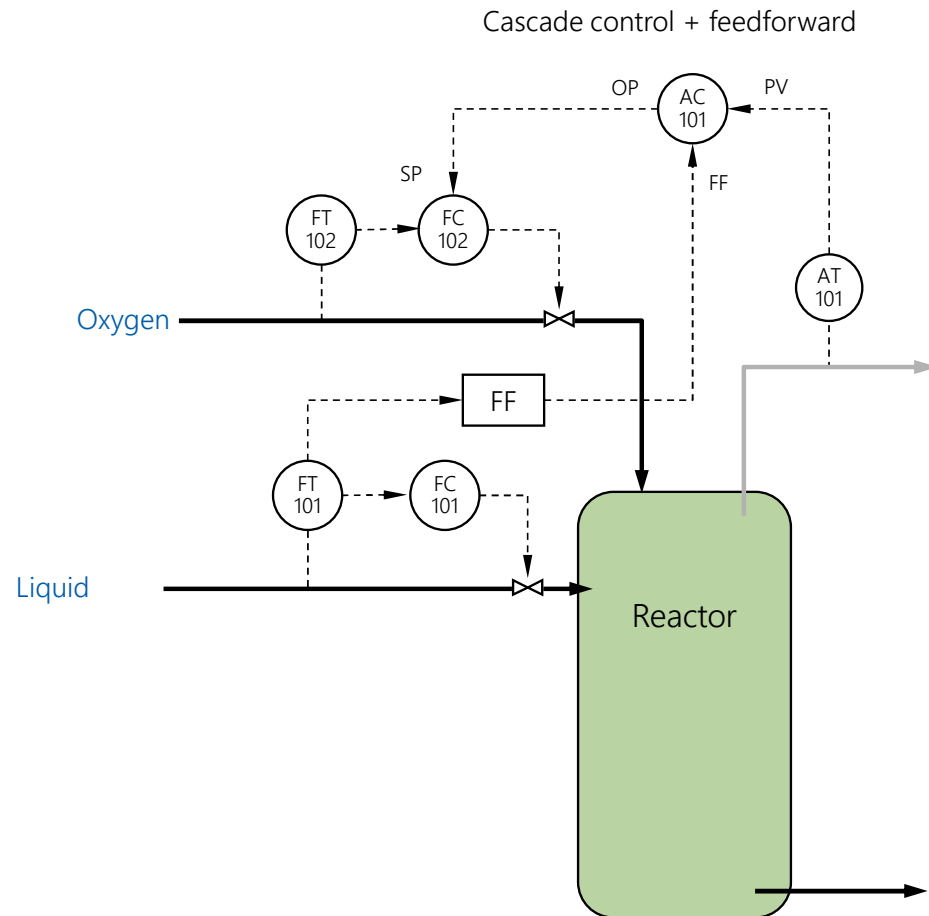
Option 0: Residual oxygen control manipulating valve directly



Option 1: Oxygen feed in cascade against residual O₂

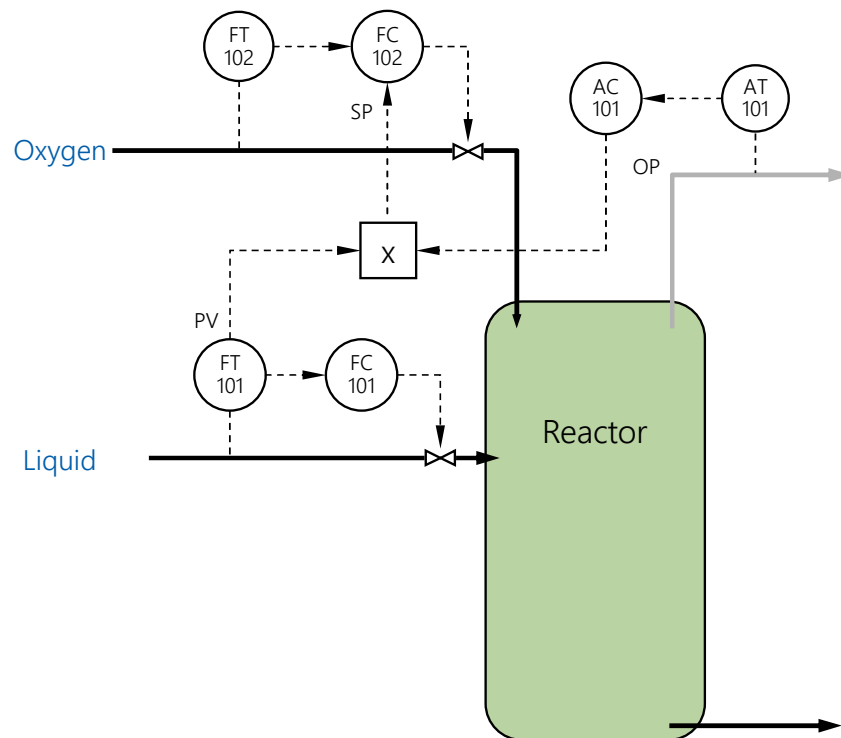


Option 2: O₂ feed in cascade + feedforward from aldehyde



This structure is better than cascade only if the liquid feed varies, e.g. during start-ups.

Opt. 3: O₂ flow in ratio against liquid, cascade against residual O₂



This structure is even better than cascade + feedforward.
"Adaptive feedforward"

In which situations is it better than FF?

Answer:

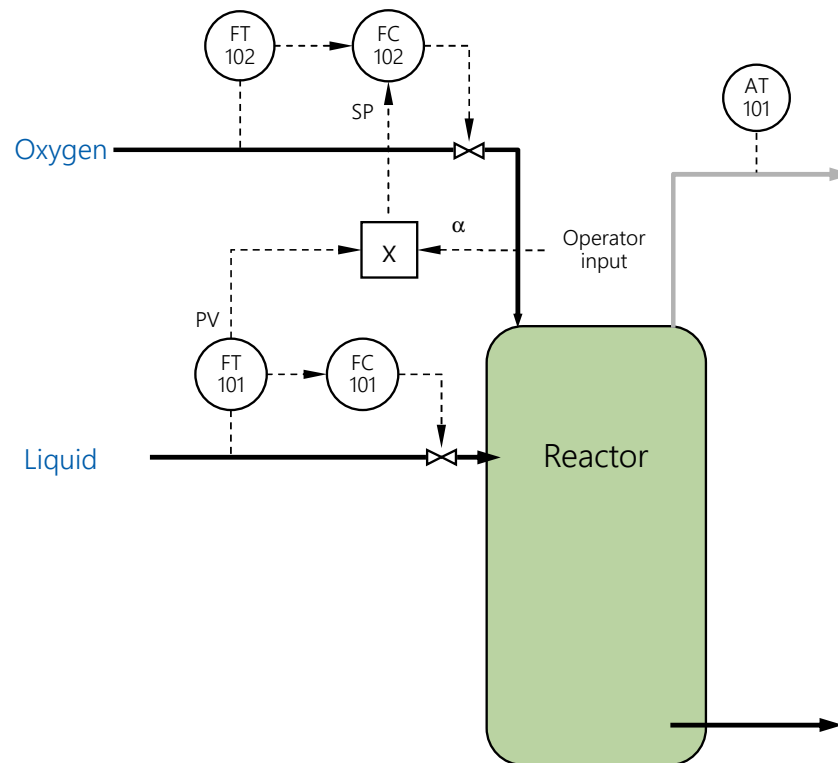
If the concentration in the liquid flow changes, then variations in feed flow are best handled by this structure.

Scenario: The concentration in the liquid feed changes permanently. The structure showed here only needs to compensate for this once. At the next production rate change there is no transient. But for "cascade + FF" we get the same transient at every change in production rate.

Also: With feedforward the FF factor would need to be changed depending on the setpoint of the AC.

Not a solution: O₂-content is not controlled

Ratio control without cascade does not solve the task.
Here the oxygen content is not controlled.

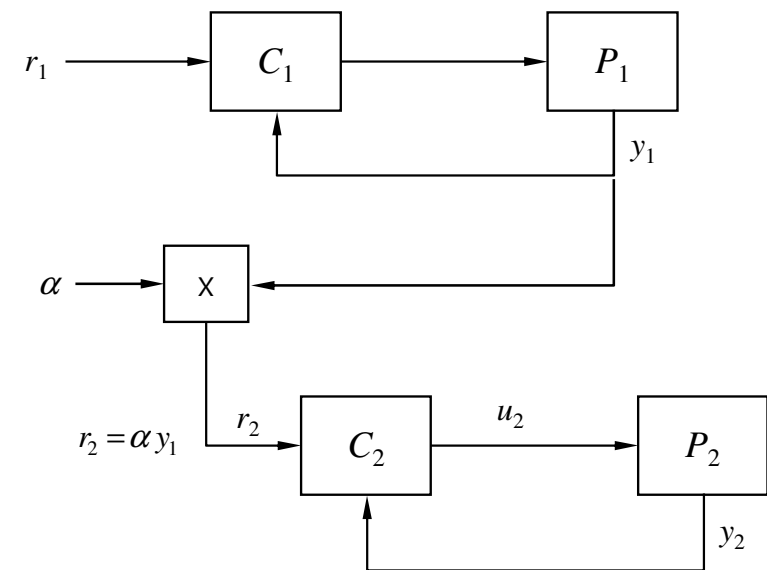


Protection against slave controller saturation

- A potential problem in ratio control is that if the slave controller (manipulated variable) saturates, e.g. having output $u=100\%$, then the control is lost.
- Below we propose a solution to this problem.

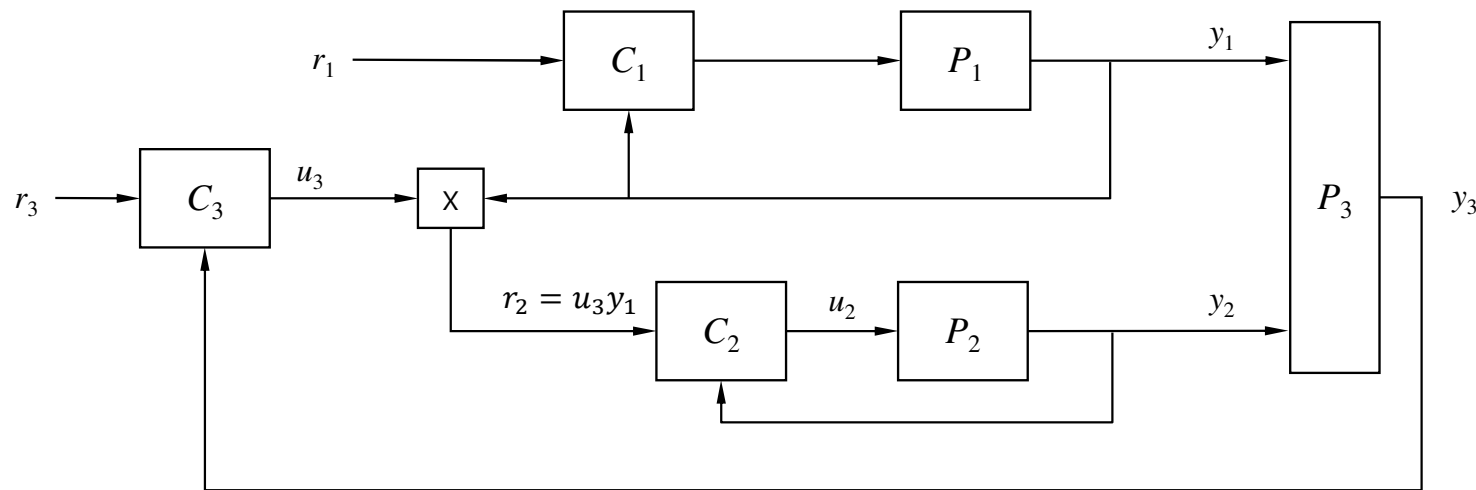
Traditional ratio control, without feedback

- The desired ratio α is set by the operator.
- If u_2 reaches its max value, then the ratio will deviate from the desired value in steady state.

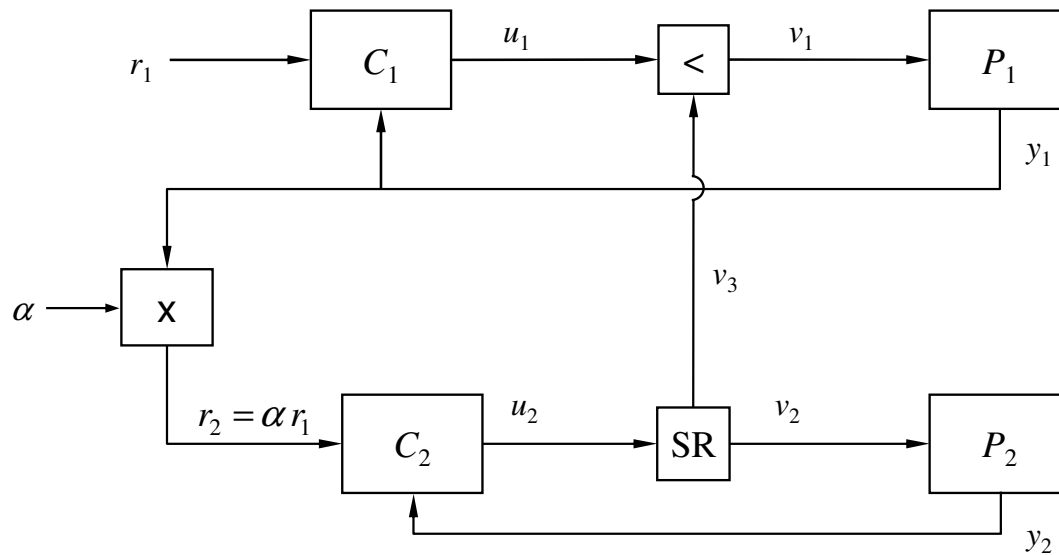


Ratio control with feedback; same issue

- This structure can be used when there is a measurement y_3 dependent on the ratio of flows 1 and 2.
- However, it suffers from the same problem when u_2 saturates.



Solution: New structure using SRC and selector

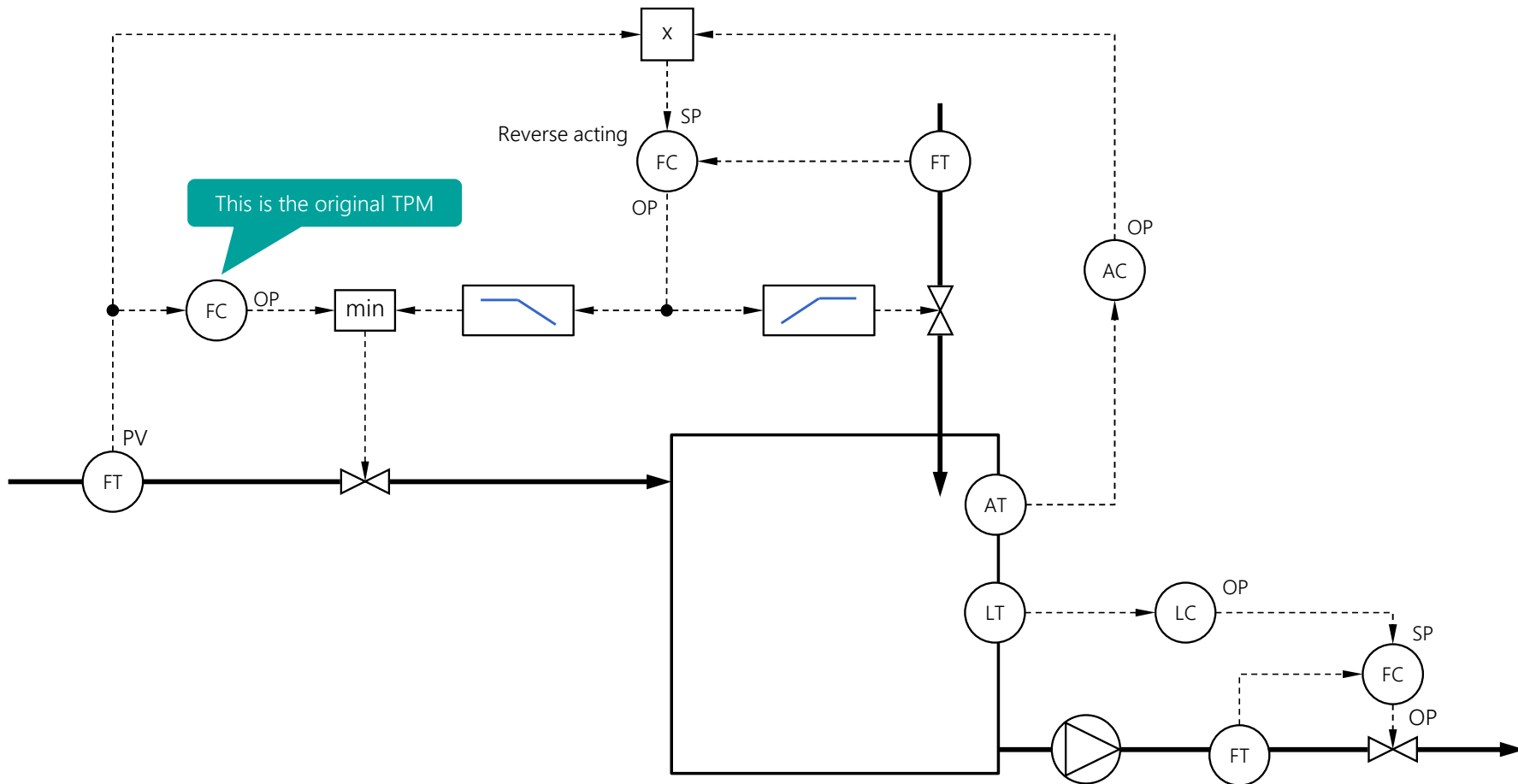


$$v_1 = \min(u_1, v_3)$$

SR-table
(example)

u_2	v_2	v_3
0	0	100
50	100	100
100	100	0

Application example: Blending



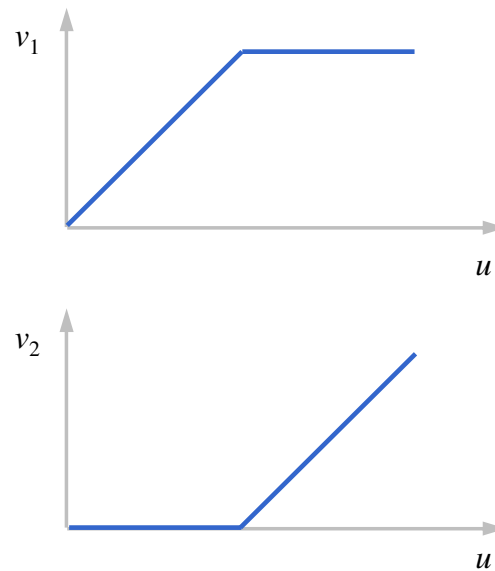
Split-range

(SRC: Split-range control)

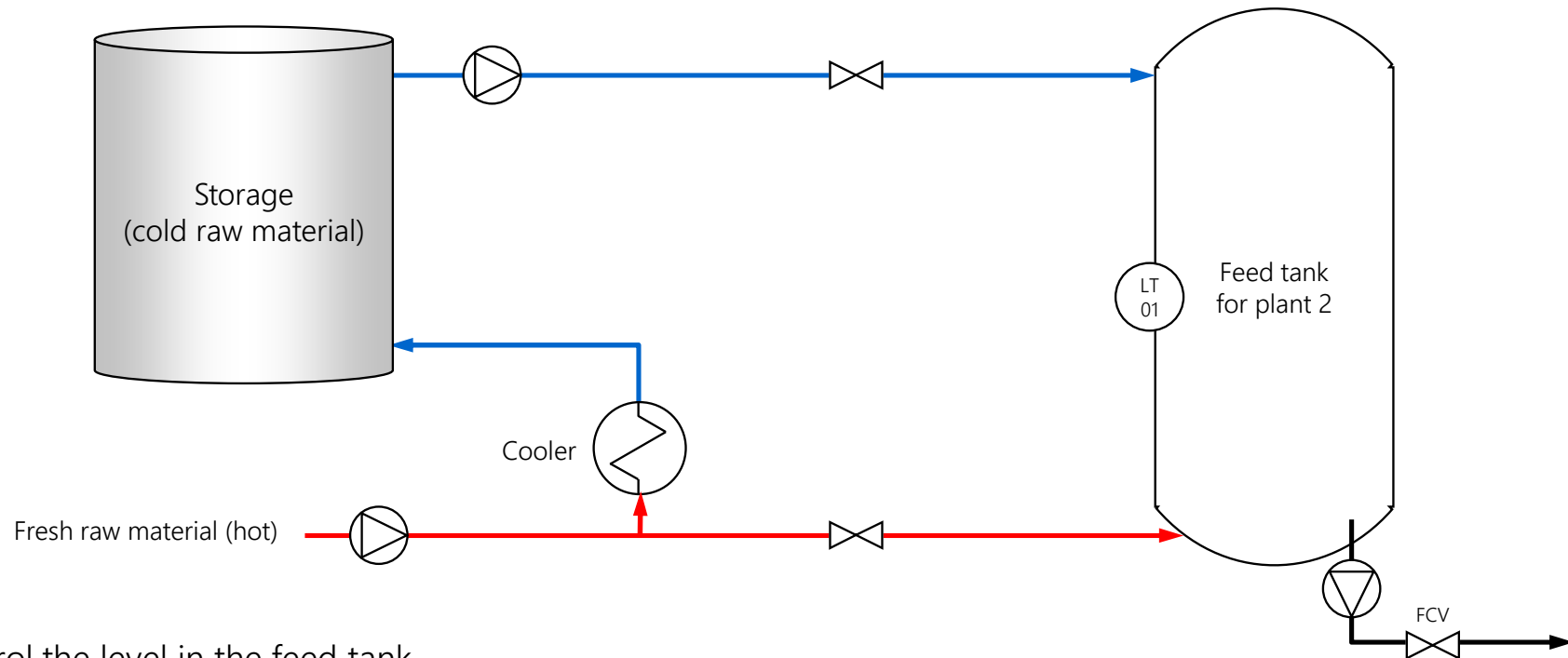
Split-range: two valve – one controller

- In some applications we have two MVs and want the controller to use one “first” and the other one next.
- The most common solution is to send the controller OP to two tables; one per valve.
For example:

OP	v_1	v_2
0	0	0
50	100	0
100	100	100



Level control with two manipulated flows



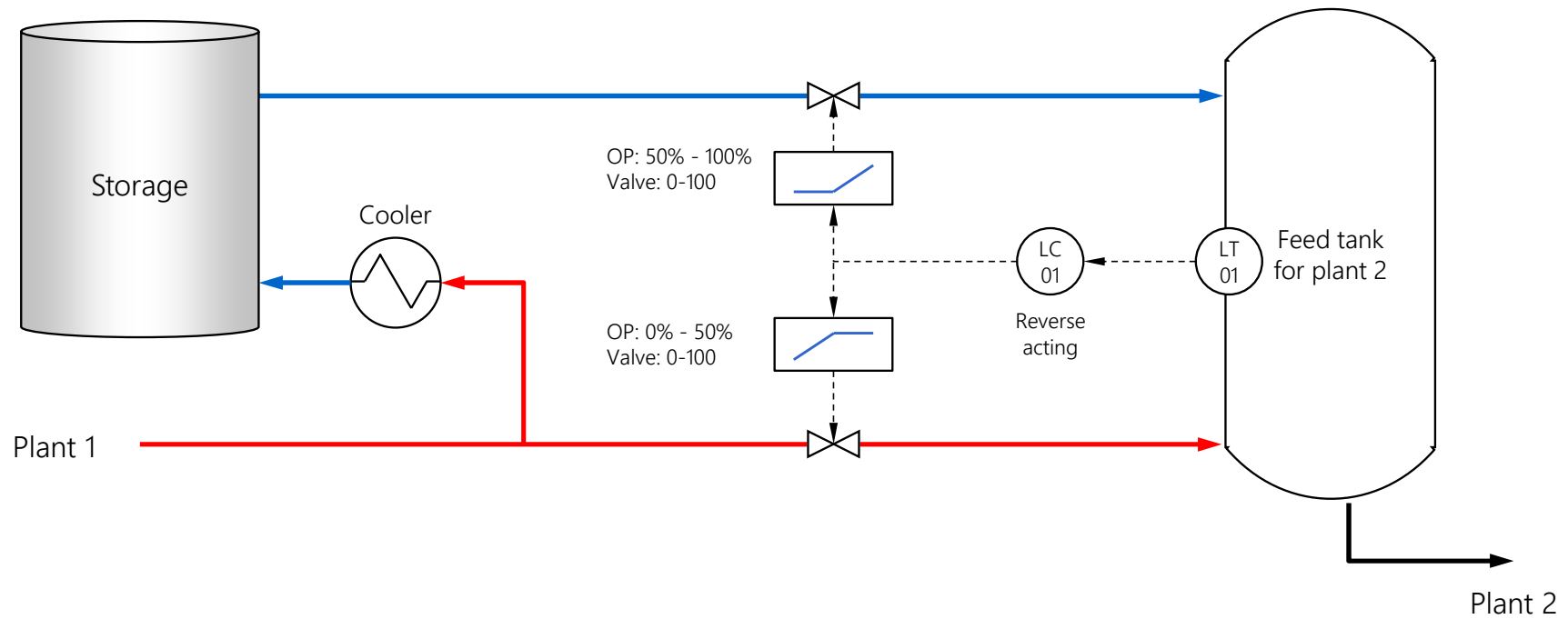
Main task: Control the level in the feed tank.

Use the extra degree of freedom for energy optimization: Primarily take fresh, hot, raw material.

If that doesn't suffice, take from storage.

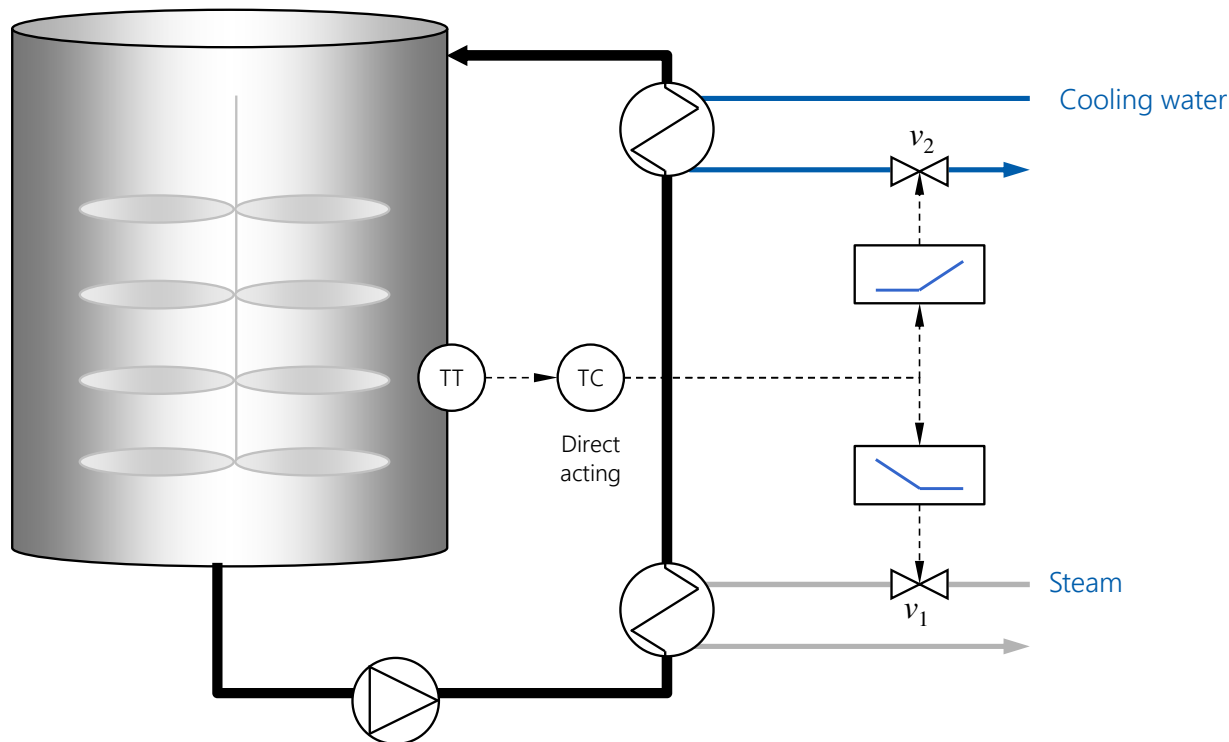
If we don't have optimizing control we may cool unnecessarily, and then re-heat.

SRC application: level control with optimization



Another SRC application: Cooling - Heating

- Another example of split-range: cooling or heating for temperature control.
 - Here the MVs are “qualitatively different” and likely to have different dynamics etc. Could be hard to tune.
 - E.g. exothermal reaction in a continuous reactor. Heating for start-ups, cooling during normal operations.

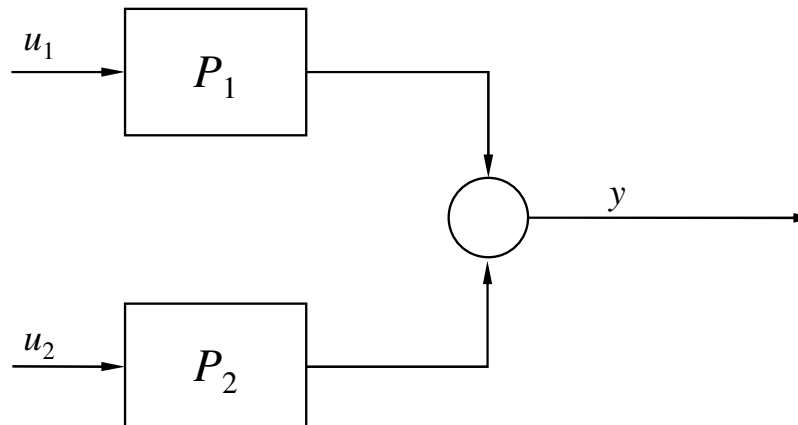


OP	Steam	C.w.
0	100	0
50	0	0
100	0	100

Valve position control (VPC)
("Mid-ranging")

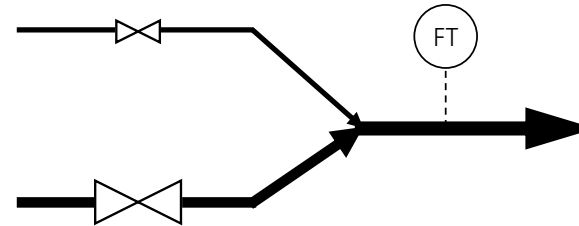
Extra degree of freedom: A coarse and a fine MV

- Scenario: We have two different MVs at our disposal to control one process variable.
- The dynamics from u_1 may be different from the one from u_2 . Typically one has larger gain and slower response than the other.
- How do we utilize this extra DoF?

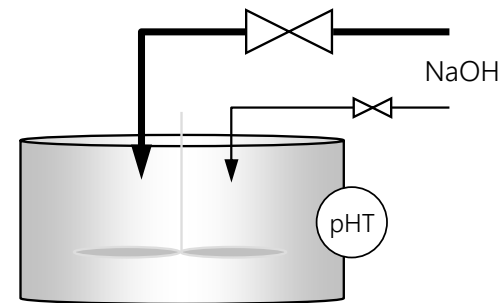


Typical processes with extra DoF

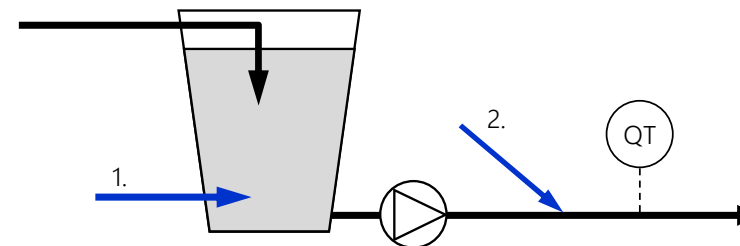
- Ex 1: Two parallel valves of different sizes manipulate the same flow:



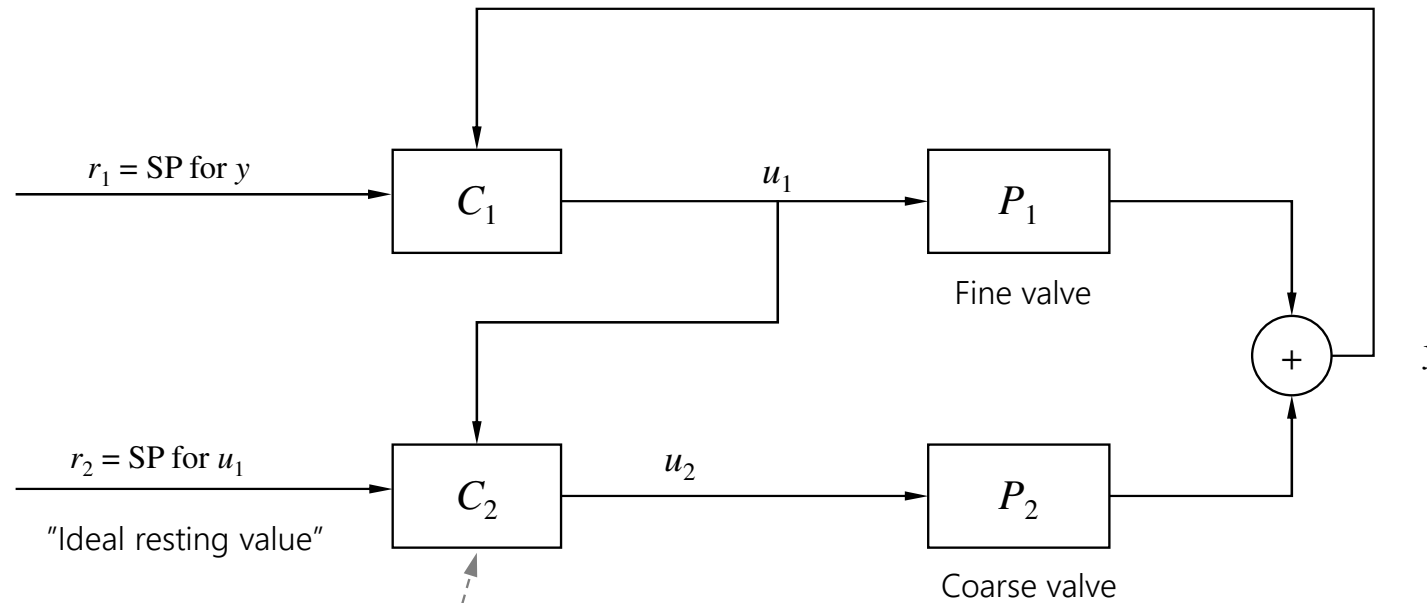
- Ex 2: Control pH in a tank, by manipulating caustic addition. Fine and coarse valve



- Ex 3: Dilution in two steps: in tank and pipe



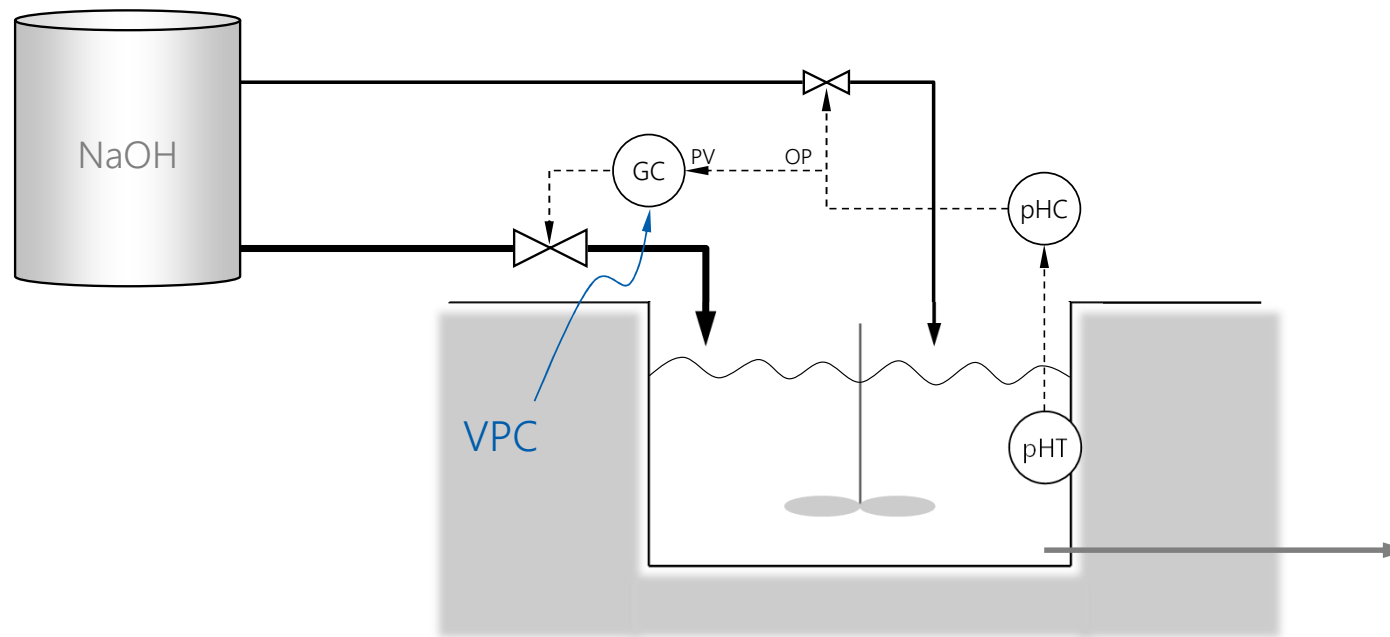
VPC: New controller with u_1 as controlled variable




C_2 is a valve position controller (VPC)
This controller should be considerably slower than C_1

Example: Sewer pH control; Mid-ranging

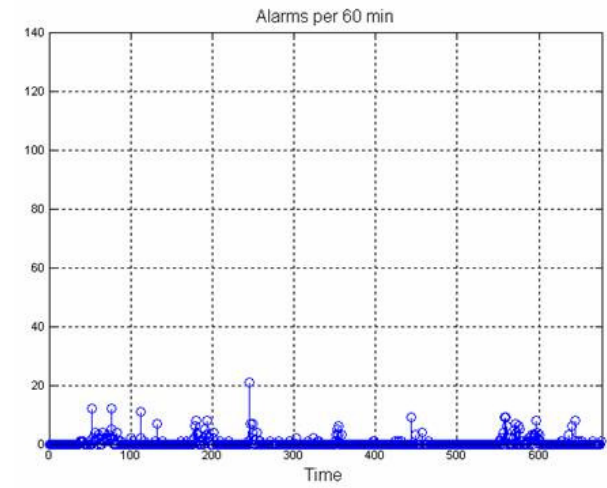
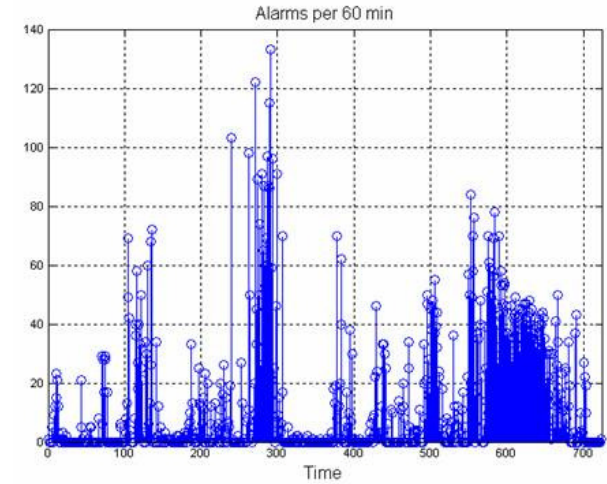
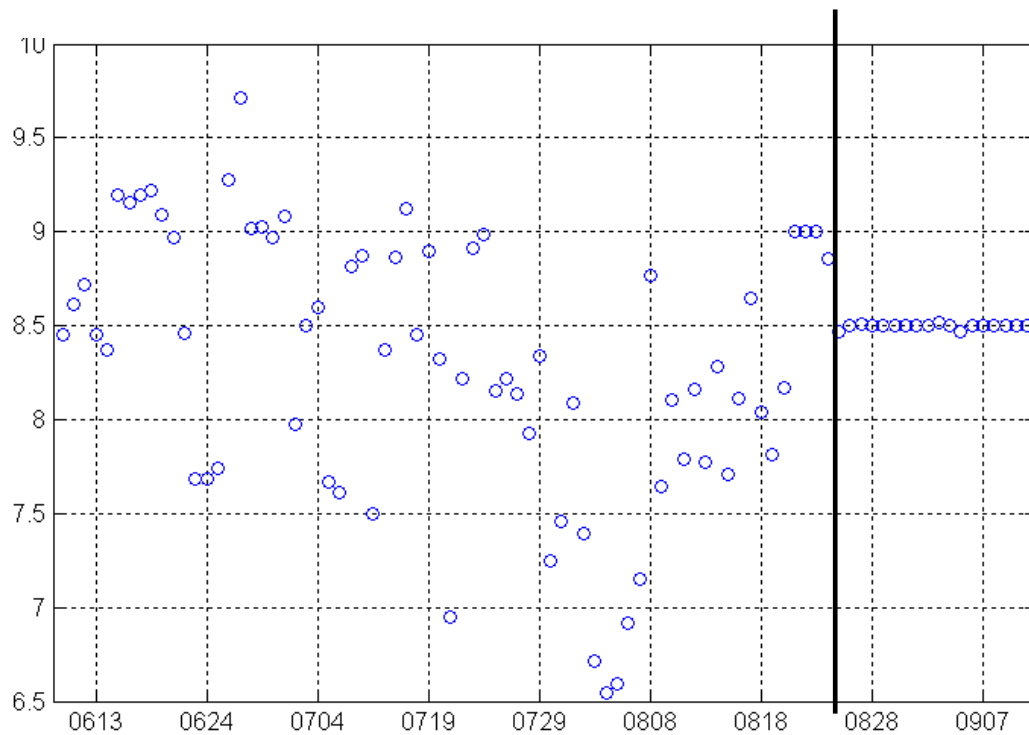
Exercise: How should we control the pH in the pit, using both valves?



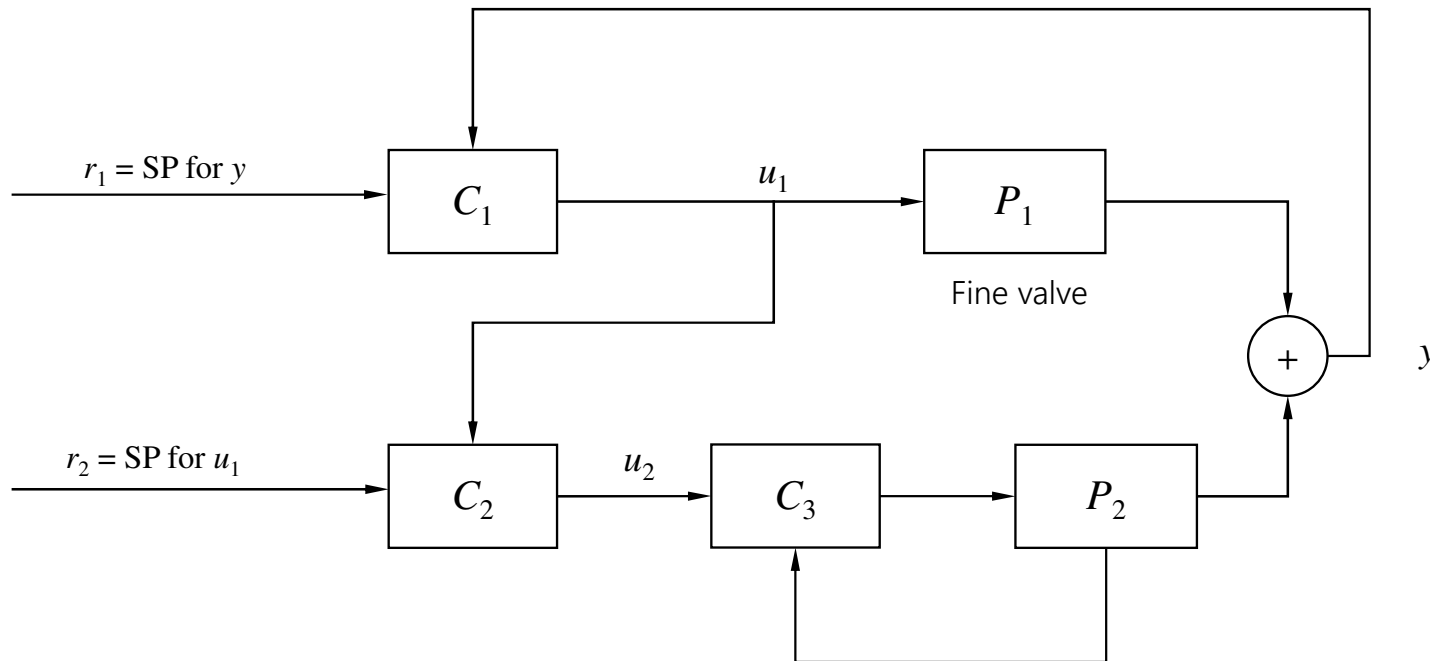
pH control: Results

Before: 10 488 alarms in one month After: 418
 Perstorp
A Subsidiary of PETRONAS Chemicals Group

pH: Daily averages before and after VPC



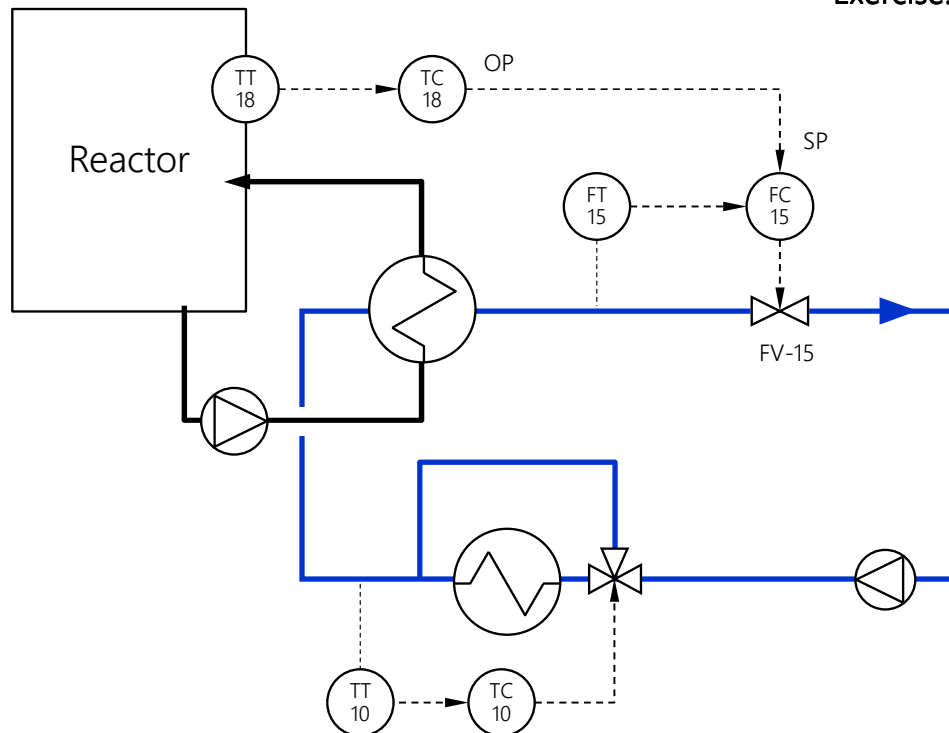
Variant: The VPC gives the SP for a slave controller



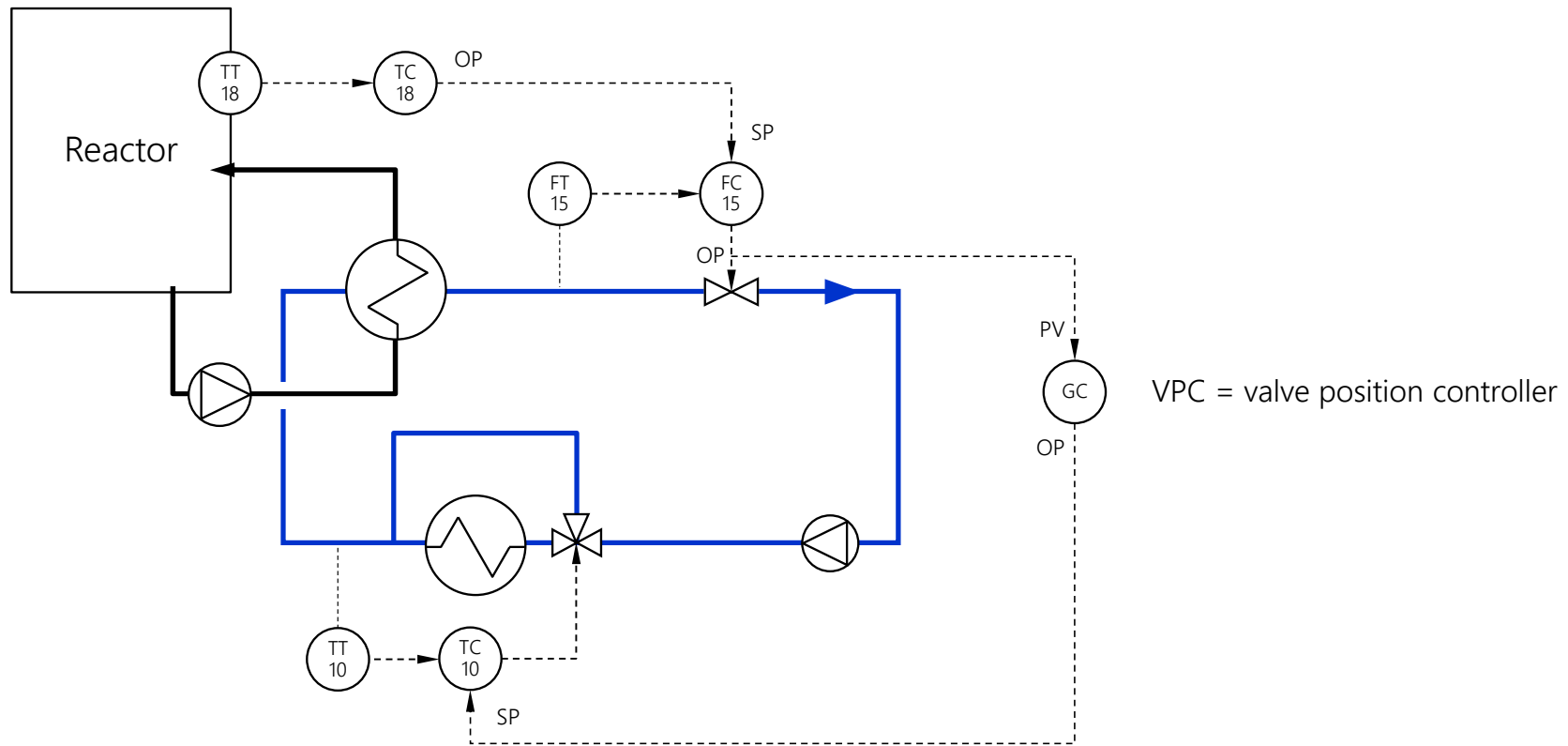
Mid-ranging application: Cooling water system

Problem: Sometimes the flow valve FV-15 saturates. Then the operator has to change the SP for cooling water temperature.

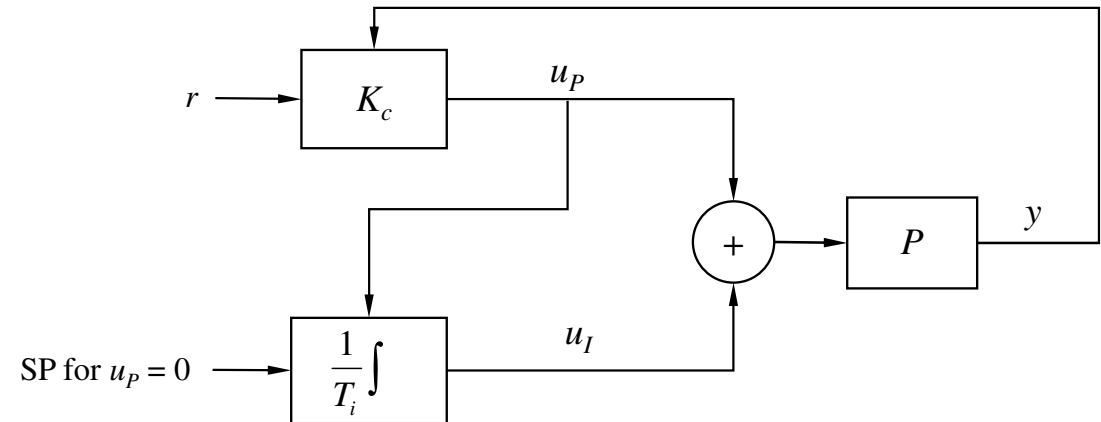
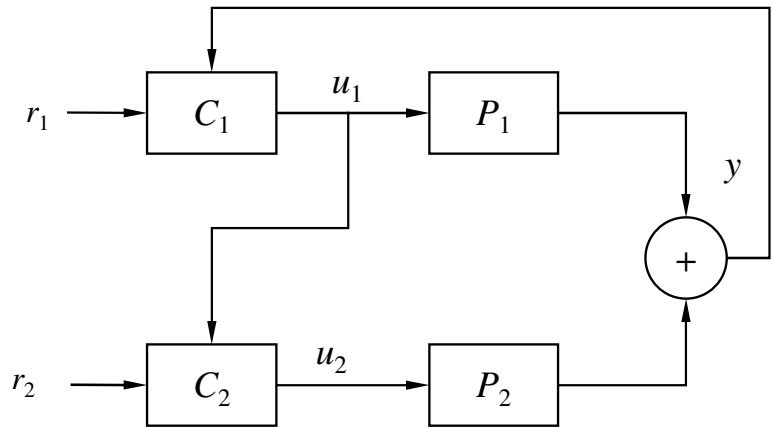
Exercise: Suggest a mid-ranging solution that does this automatically!



Solution: New controller (VPC) giving the SP for TC

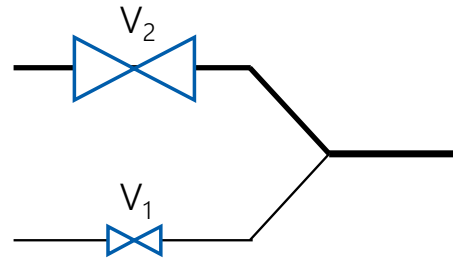


Fun fact: VPC ~ PI-controller



- The PI-controller can be seen as a special case of mid-ranging, internally.
 - The I-part and the P-part can be seen as two controllers. The task of the I-part is to reset the P-part to 0.

Split-range or mid-ranging?



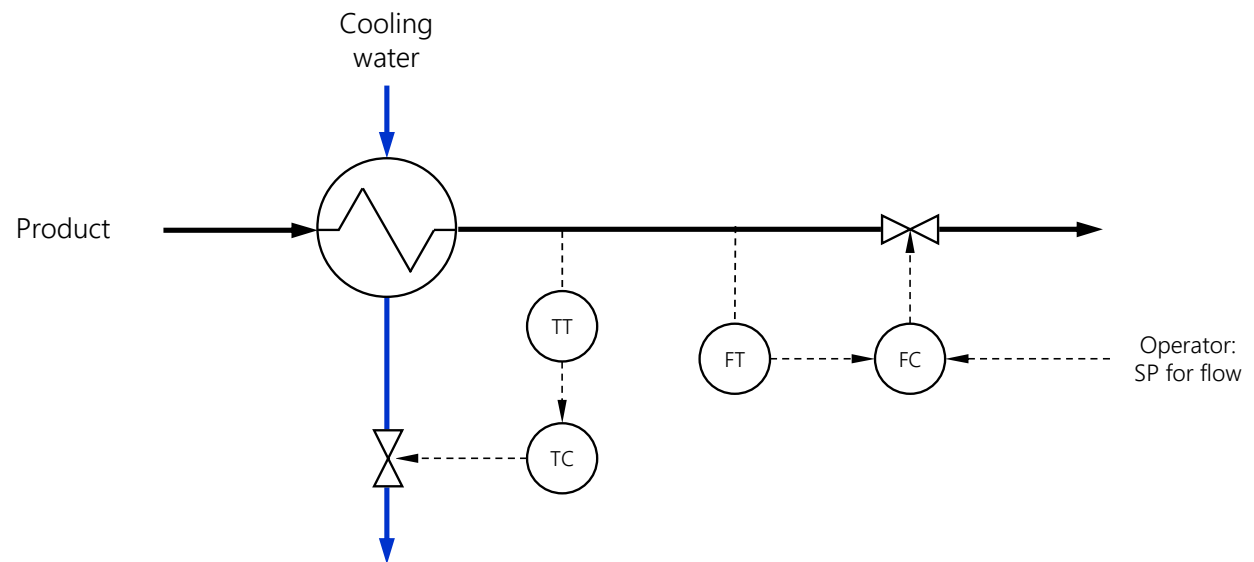
- If we have a process with big difference in valve dynamics and gain: Should we use mid-ranging or split-range?
- Answer: It depends on the specifications (the optimization criterion).
 - Split-range lets us use stream 1 to the max before taking from stream 2. Could be optimal if number 1 is cheaper.
 - Mid-ranging gives maximum control precision.

Maximizing
control

Ex: Maximize flow in a heat exchanger (cooler)

Temperature must be kept at a given setpoint, for example 45 degrees.

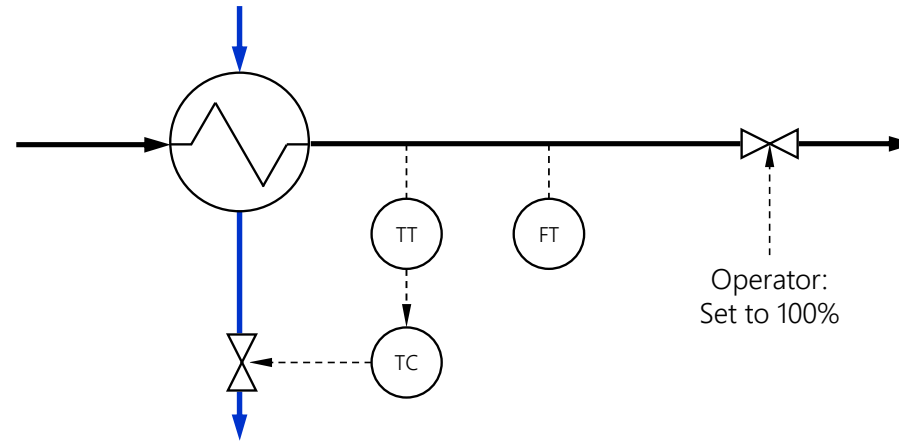
“Traditional” structure:



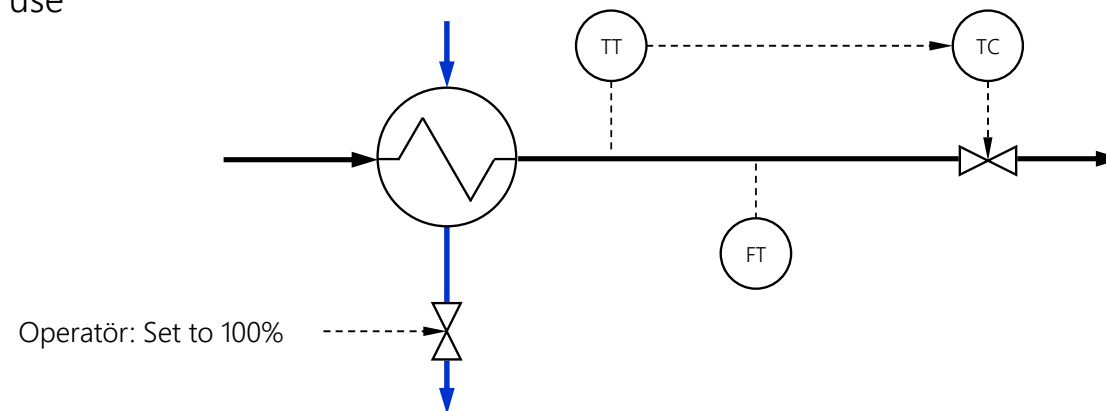
In this structure there is no automatic mechanism that guarantees maximal flow.

Cont; Maximize flow in a heat exchanger

If the flow valve is the limiting factor, this structure should be used.



If the cooling water valve is limiting, use this structure.

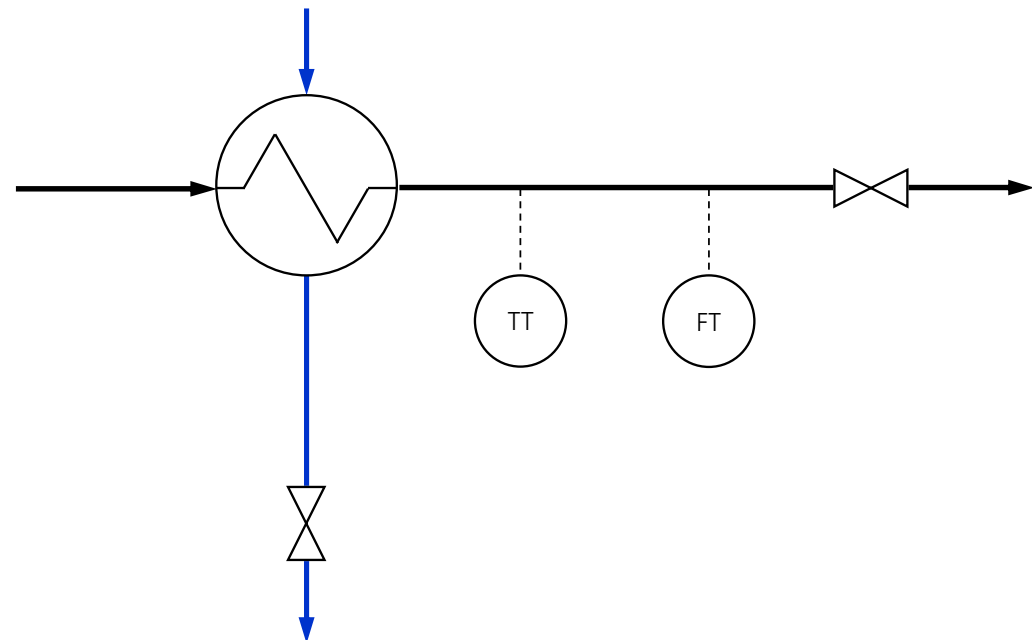


Maximizing control

- Main idea: Use the throughput manipulator (TPM) as manipulated variable
 - Lock the bottleneck variable and manipulate the master flow to control what the bottleneck variable used to control.
- Example: Drying limited paper machine. Open steam valves 100% and manipulate machine speed to control the moisture content in the sheet!

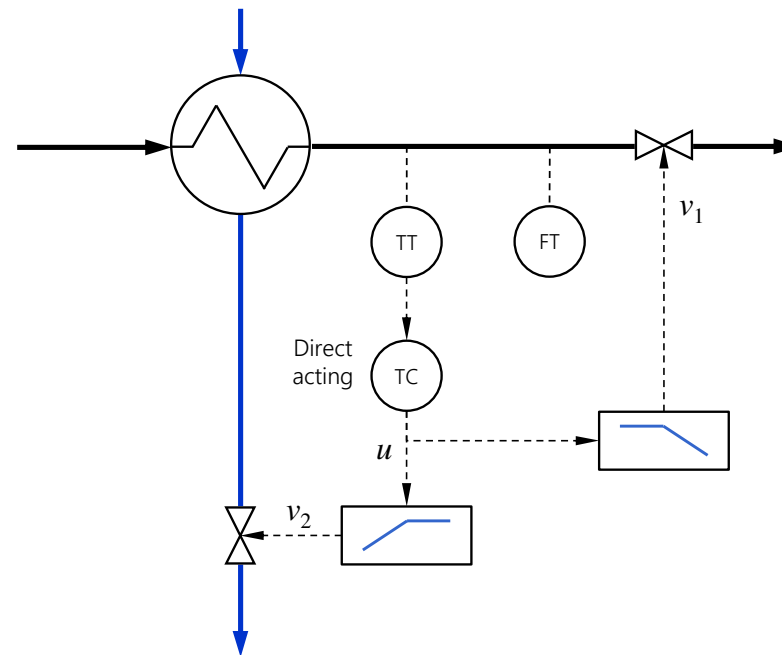
Exercise: Handle both scenarios

- It's not uncommon that sometimes the flow valve and sometimes the cooling water valve is the limiting variable, e.g. due to variations in cooling water temperature.
- **Exercise:** Find a structure that handles both cases.
Clue: The structure relies on split-range.



Solution: Globally maximizing structure

- Temperature controller does the following:
 - Primarily: keep the production valve fully open and manipulate cooling water valve to control the temperature
 - Secondly: keep cooling water valve fully open and manipulate the production valve to control the temperature

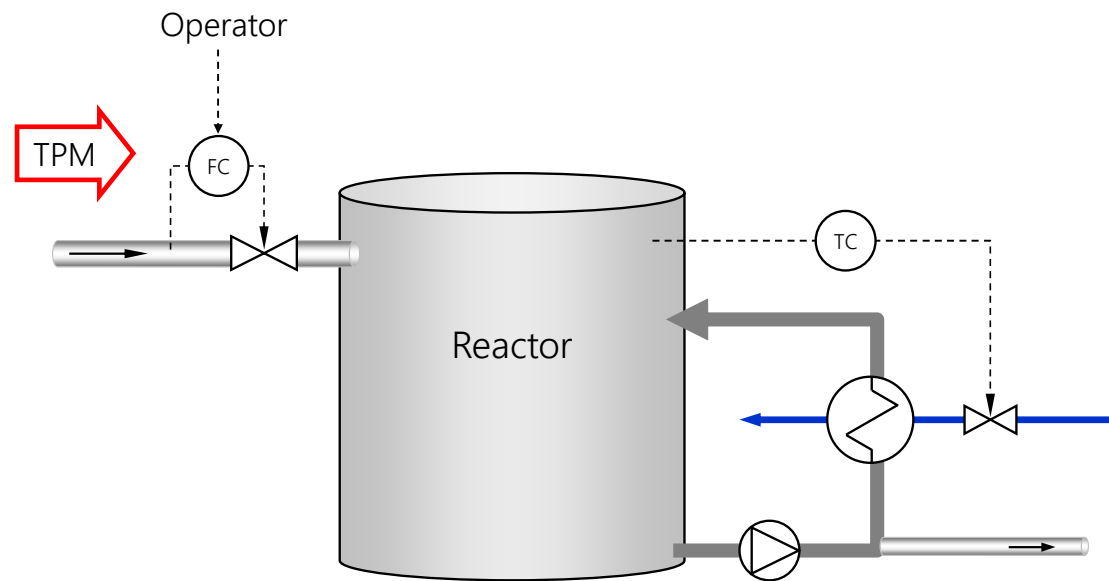


Variant of maximizing control

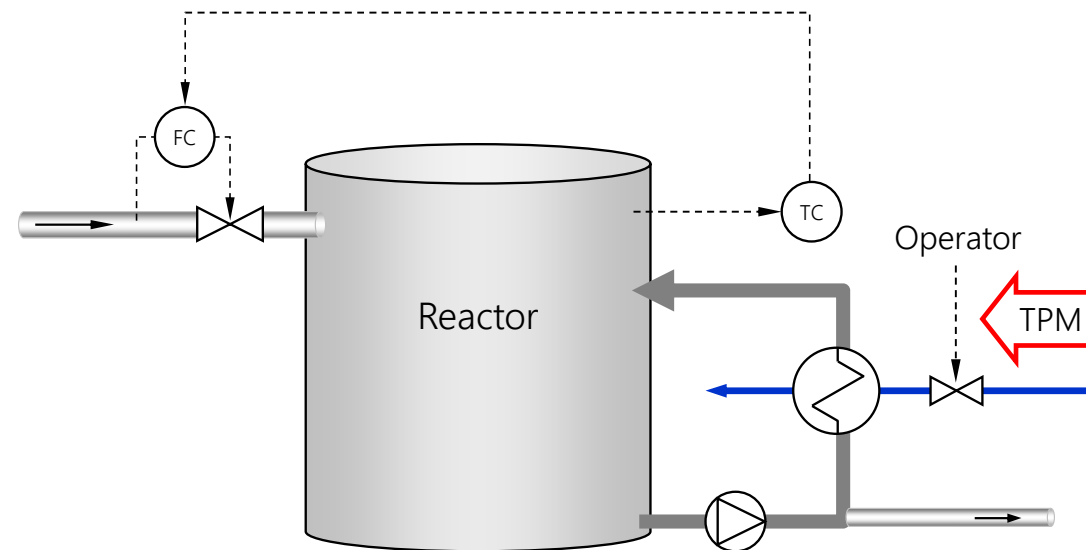
- Don't switch manipulated variable, but slowly adjust SP for master flow so that controller output for the limiting process is at a suitable distance from its limit.
 - Example: Dryer limited paper machine. Machine speed automatically increased until the steam valves are 90% open, on average.
 - A **mid-ranging controller** can be used for this.

Example: Cooling capacity limited reactor

Normal operation

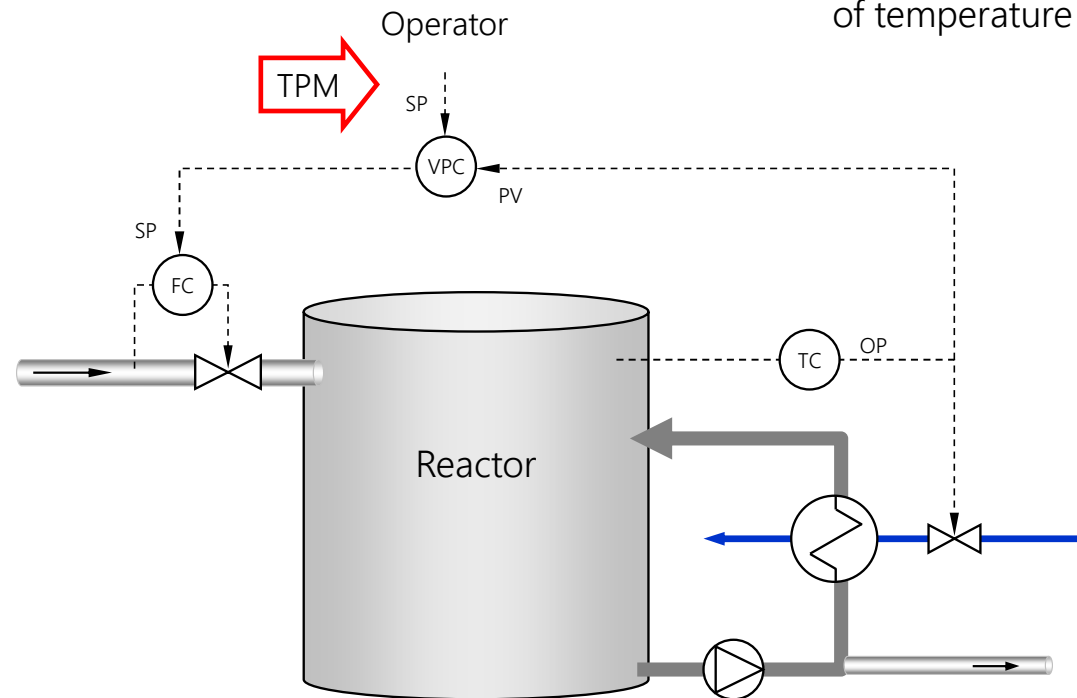


Cooling limited Rx: Maximizing control 1



Maximizing control 2: Use VPC

Not as high production rate as with solution 1, but better performance of temperature controller.

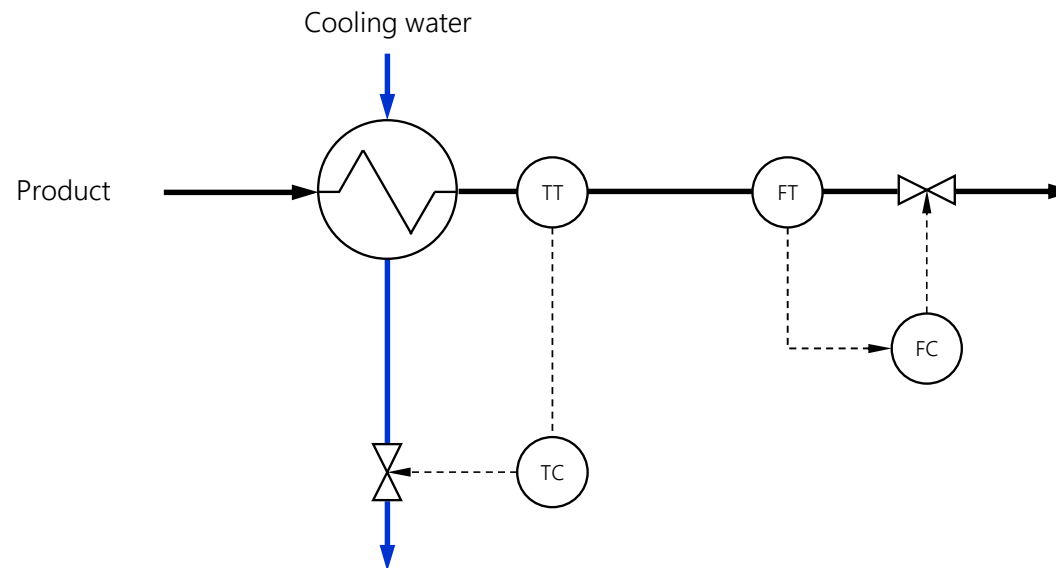


Exercise: HEX flow maximization with exact temp control

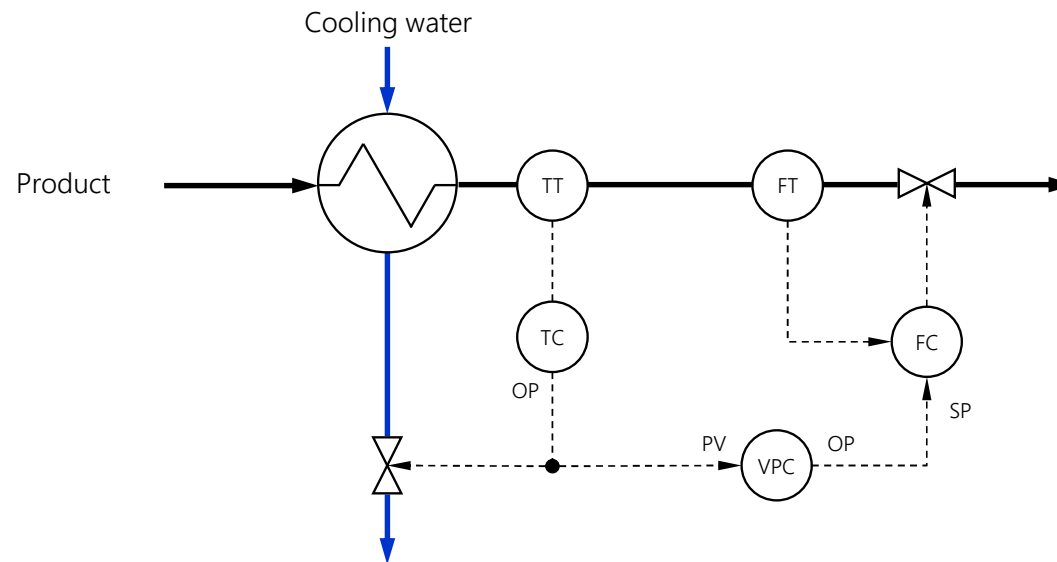
Cooler example.

Design a control structure where the flow can be set close to its max limit, but only the cooling water valve is manipulated for controlling temperature.

Start from this structure:

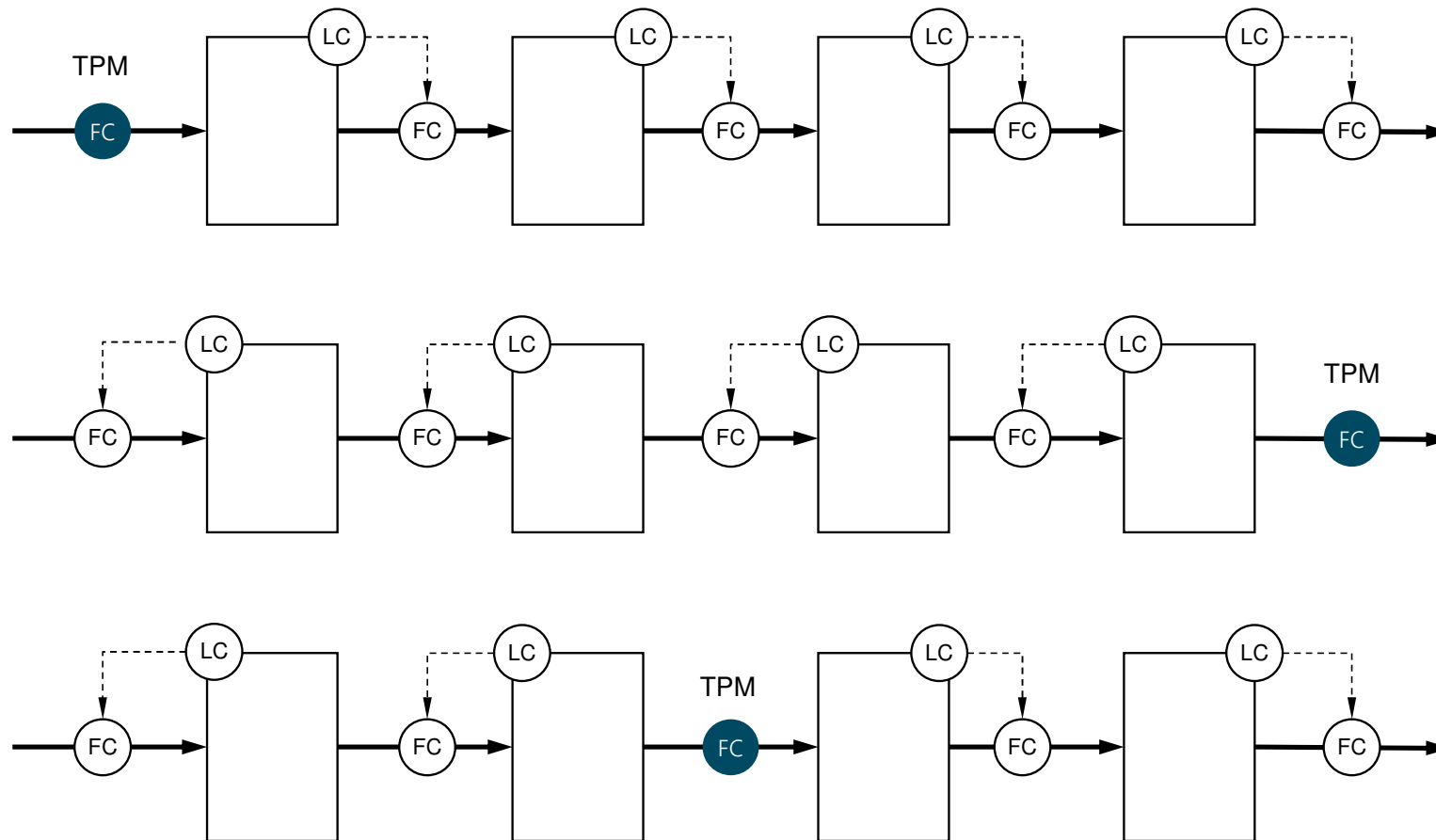


Solution: VPC for maximizing control of HEX



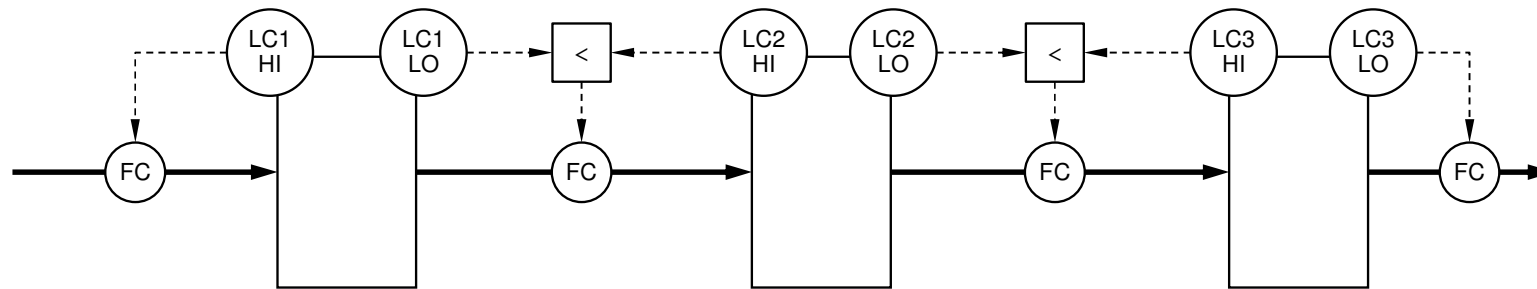
Bidirectional control

Common scenario: Bottleneck/TPM moves over time



“Bidirectional control” automates when TPM moves

- Often, there is a different bottleneck in different operating points
- Sometimes you want to handle this automatically. Solution is **bidirectional control**.
- Can be implemented in different ways. Simplest: dual level controllers with different setpoints and min-selector.



HI-controllers: reverse acting
LO-controllers: direct acting

A simple bidirectional inventory control structure with optimal buffer management

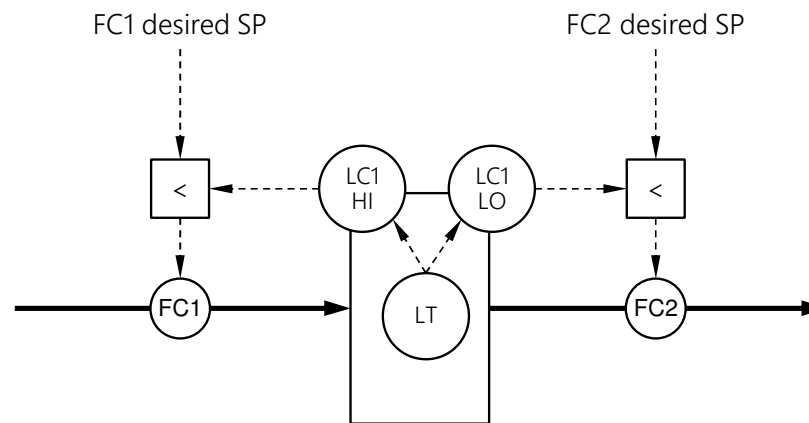
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^aDepartment of Chemical Engineering, Norwegian University of Science and Technology (NTNU), 7491, Trondheim, Norway

^bPerstorp Specialty Chemicals, 284 80, Perstorp, Sweden

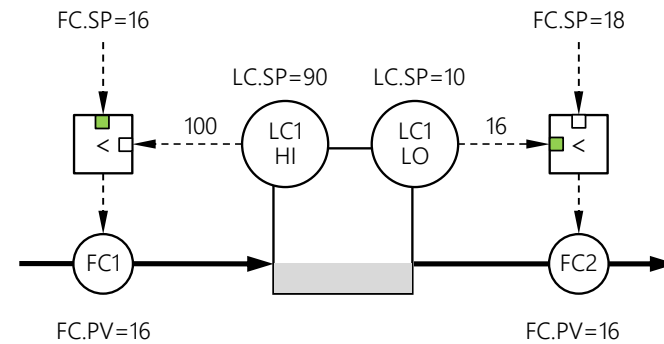
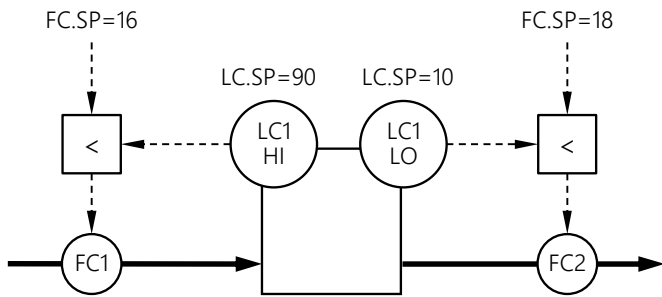
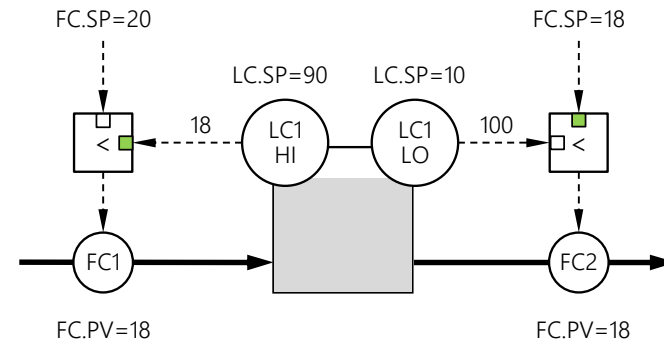
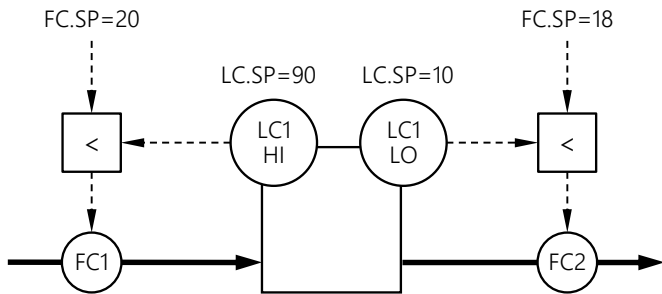
Basic element of bidirectional control

- A system of bidirectional controllers is made up from the following basic building block.



HI-controller: reverse acting
LO-controller: direct acting

Examples



Example: New raw water system in Stenungsund

Combination of selector control and parallel control

