Comments on paper: Sigurd Skogestad, ''Advanced control using decomposition and simple elements'', *Annual Reviews in Control,* **vol. 56** (2023), Article 100903.

Comment 1 : Correction of reference for hierarchical decomposition

Comment 2. (Des. 2023) Perry (1973) has an early description of predictive control.

Comment 3. (Des. 2023) Perry (1999) gives an example of the use of transformed inputs for linearization and feedforward (E14)

Comment 4. (March 2024, Feb. 2025) On "separate controllers with different setpoints"; E6.

Comment 5. (July 2024) One more split range control (SRC) scheme for MV-MV switching (The 4th alternative)

Comment 6. (December 2024) On the choice of the filter time constant τ_F

Comment 7. (December 2024) Filter on D-action

Comment 1 Correction of reference for hierarchical decomposition

(Des. 2023). Re Figure 4 (decomposition into layers).

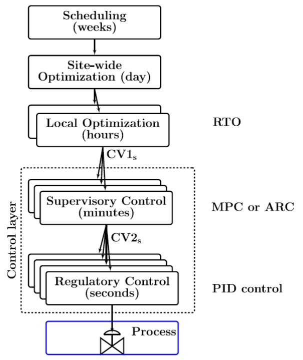

Fig. 4. Decomposition of "overall control system" for optimal operation in typical process plant. This involves a vertical (hierarchical) decomposition (Richalet et al., 1978) into decision layers based on time scale separation, and a horizontal decomposition into decentralized blocks/controllers, often based on physical distance. There is also feedback of measurements (y, w, d, CV1, CV2) (possibly estimates) from the process to the various layers and blocks but this is not shown in the figure. This paper considers the three lowest layers, with focus on the supervisory control layer.
CV1 = Economic controlled variables
CV2 = Regulatory/stabilizing controlled variables
RTO = Real-time optimization
MPC = Model predictive control
ARC = Advanced regulatory control
PID = Proportional–Integral–Derivative.

In the paper it is referred to Richalet et al. (Automatica, 1978) for Figure 4 but in that paper there is actually no such figure, so this must be a misprint. However, Perry's handbook (1973) has the following similar figure:

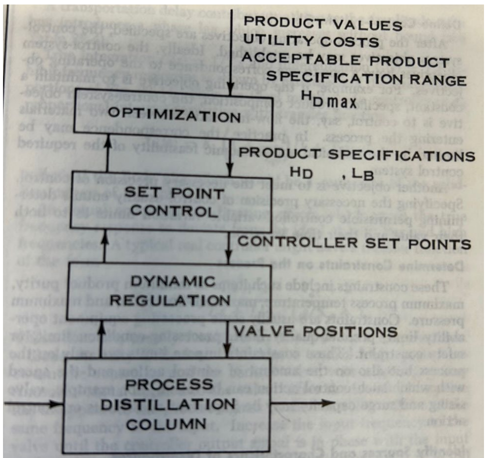

FIG. 22-210. Hierarchical control structure of distillation unit operation.

Comment 2. **(Des. 2023) Perry (1973) also has an early description of predictive control:**

... the jacket-temperature set point ... of the

Predictive Control. An analog computer, functioning as an on-line predictive controller, serves as another example of control techniques available for use on batch reactors. This example is a pilot-plant reactor producing a synthetic rubber by an exothermic reaction. [Adams and Schooley, *Instrumentation Tech.*, **16**, 57–62 (1969).] Heat-transfer characteristics of the reaction mixture deteriorate as reaction proceeds and temperature variations produce large fluctuations in the reaction rate. Product properties are highly dependent on reaction-temperature history. Low or high monomer concentrations yield poor-quality product; additionally high monomer concentration can cause a rapid secondary reaction. The control problem is to make a batch of rubber within a small range of reaction temperature, and stop the reaction at a predetermined conversion.

The control scheme used is illustrated in Fig. 22-209. The model of reaction kinetics and energy balance is operated in parallel with the real reactor, using measured process temperatures as inputs. In real time, the model calculates the current state of the reaction in terms of monomer conversion and catalyst activity. Periodically, the model is switched to fast-time operation to predict the future state of the reaction. Based on this prediction, the reactor-temperature error in the controller is biased by the future temperature, giving additional control action before the reaction moved out of desired limits.

Optimizing Control. A batch hydrogenation reaction has become the classic laboratory example of dynamic optimization of batch-unit operation. [Eckman and Lefkowitz, *Control Eng.*, **4**, 197–204 (1957).] In this example the performance variable is best described by an integral involving quantities which vary with time. Utilizing information about a typical batch, an optimum path for reaction and control parameters can be determined. Using this as a guide, each batch is closely monitored with a computer. As the batch progresses, deviations from the optimum are used to predict and make control adjustments to keep the batch on a reaction path close to optimum.

In the simplest case such performance variables are of the form

$$I = \int_0^T f(\dot{x}, x, t)\, dt$$

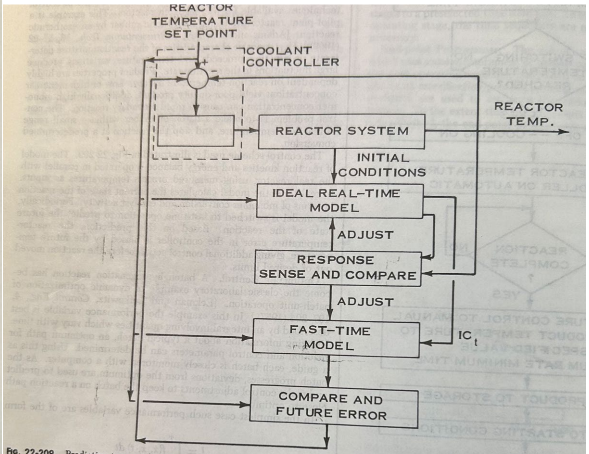

FIG. 22-209. Predictive-type analog-computer control of exothermic-polymerization batch reactor.

Comment 3. (Des. 2023) Perry (1999) gives an example of the use of transformed inputs for linearization and feedforward (E14) for a heat exchanger (the heat exchanger example is discussed in more detail in the paper by Skogestad, Zotica and Alsop (JPC, 2023).

$$Q = WH = FC_L(T_2 - T_1) \qquad (8\text{-}74)$$

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Figure 8-50 shows a temperature controller (TC) setting a heat-flow controller (QC) in cascade. A measurement of the manipulated flow is multiplied by its temperature difference across the heat exchanger to calculate the current heat-transfer rate, using the right side of Eq. (8-74). Variations in supply temperature, then, appear as variations in calculated heat transfer, which the QC can quickly correct by adjusting the manipulated flow. An equal-percentage valve is still required to linearize the secondary loop, but the primary loop of temperature-setting heat flow is linear. Feedforward can be added by multiplying the dynamically compensated flow measurement of the other fluid by the output of the temperature controller.

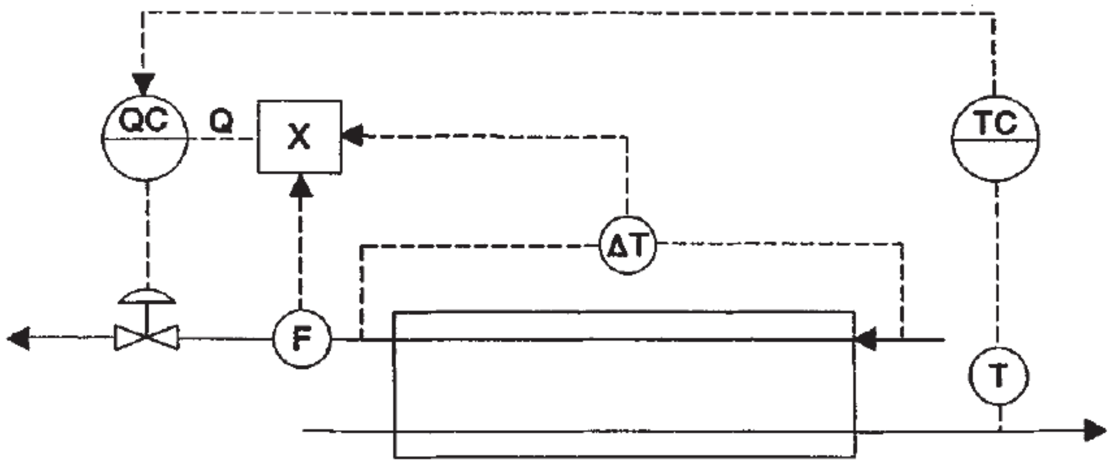


FIG. 8-50 Manipulating heat flow linearizes the loop and protects against variations in supply temperature.

Comment 4. On "separate controllers with different setpoints"; E6, (March 2024, Feb. 2025).

Feb. 2025: As a better name for "separate controllers with different setpoints" (E6) I now propose **"split parallel control".** This is a good name because it is a parallel structure which is an alternative to "split range control" (E5). There is also a structure known as "parallel control" (Fig. 13, see below) but this has only one setpoint and both controllers (C1 and C2) are active all the time. In "Split parallel control" (Fig, 22, see below) the two controllers are normally not actice at the same time (except at change-over).

March 2024: In my paper, the oldest reference I give for using "separate controllers with different setpoints"; E6, Fig. 22) for MV-MV switching is the book by Smith (2010) (page 86) (see below). The name "separate controllers" is used by Smith (2010). However, this scheme has obviously used in industry long before this. For example, an older reference is the book by **Forsman (2005) (in Swedish)** (page 152-153).

In the section title (and also in the flowsheet, see his Figure 6.28) Forsman calls it "Many controllers with the same CV" (similar to what I call it based on Smith (2000), but in the corresponding block diagram (Figure 6.29) he calls it "Parallel control". However, I have used the term "parallel control" for the case where both controllers have the same setpoint and are used all the same time. On the other hand, in Figure 22 ("separate controllers") they are (ideally) used sequentially (one at a time), that is, only when u1 is saturated do we start using u2. So in some sense the control action is "split" similar to split range control. So I propose (Feb. 2025) to call "separate controllers with different setpoints" (E6, Fig. 22) for "Sequential parallel control" or even better "**Split parallel control".** This shows clearly that it (E6) is an alternative to split range control (E5).

This would also make it possible to distinguish between the two similar schemes for VPC. We could call "VPC on extra dynamic input" (E3, Fig. 12) for simply "VPC" and "VPC on main steady-state input" (for MV-MV switching) (E7, Fig. 24) for "Sequential VPC" or "Split VPC" (for MV-MV switching).

Some more details on Comment 4:

This is from my paper (just a reminder):

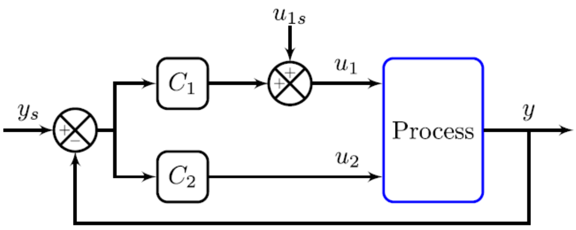


Fig. 13. Parallel control to improve dynamic response – as an alternative to the VPC solution in Fig. 12.
The "extra" MV (u_1) is used to improve the dynamic response, but at steady-state it is reset to u_{1_s}. The loop with C_2 has more integral action and wins a steady state.

3.7. Separate controllers (with different setpoints) for MV-MV switching (E6)

Consider again MV-MV switching where we want to use one MV at a time in a specific order (first u_1, then u_2, etc.). An alternative to split range control is to use separate controllers for each MV with different setpoints (Fig. 22) (Smith, 2010) (Reyes-Lúa & Skogestad, 2019).

The setpoints ($y_{s1}, y_{s2}, \dots$) should in the same order as we want to use the MVs. The setpoint differences (e.g., $\Delta y_s = y_{s2} - y_{s1}$ in Fig. 22) should be large enough so that, in spite of disturbances and measurement noise for y, only one controller (and its associated MV) is active at a given time (with the other MVs at their relevant limits).

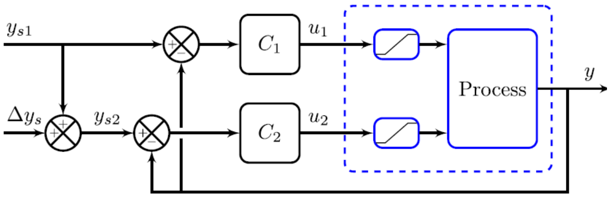
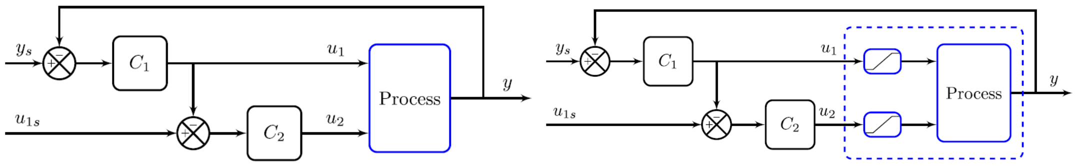


Fig. 22. Separate controllers with different setpoints for MV-MV switching.



This shows the two VPC schemes. Left is "standard VPC" (E3, Fig.12) and right is "sequential VPC" (E7, Fig.24)

This is what Smith (2010) writes on page 86:

Separate controllers for each operating mode. This normally requires that the set points for the individual controllers be separated sufficiently so that only one controller is active at a given time, the other having driven its final control element to a limit.

Split range. A single controller is provided, but its output range is "split" such that one mode of operation is active from 0 to 50% and the other is active from 50 to 100%.

Smith, C. L. (2010). *Advanced process control - beyond single-loop control*. New York: Wiley.

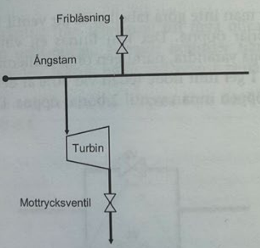
Here is from the book by Forsman (in Swedish), pages 152-153.

Krister Forsman, «Reglerteknik för processindustrin», Studentlitetratur, 2005

Flera regulatorer med samma ärvärde

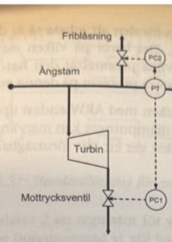
Som alternativ till split-range-reglering kan man använda flera regulatorer som alla arbetar med samma mätvärde, men med olika börvärden och olika styrsignaler, d.v.s. olika ventiler, typiskt. Detta är ganska vanligt vid processer där man har en extra möjlighet att förhindra att den reglerade variabeln inte går för högt, men det finns en kostnad för detta. Det kan handla om att vid höga tryck friblåsa ånga som annars kunnat köras genom en ångturbin, eller att fackla en gas som annars hade kunnat gå till produktion eller eldas i en panna.

Antag t.ex. att vi vill reglera trycket i en ångstam till 50 bar och för det i första hand använda ångturbinens mottrycksventil och i andra hand en friblåsningsventil. Processen ser då ut som i Figur 6.27.



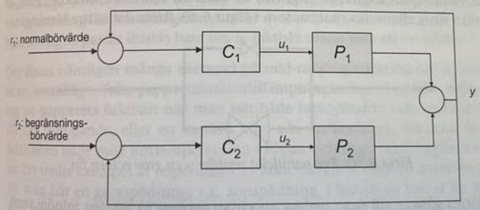
Figur 6.27: Ångstam med turbin och friblåsningsventil.

Friblåsningsventilen ska bara gå in om trycket överstiger 54 bar. Då skapar vi först en regulator som har mottrycksventilen som styrvariabel och ångtrycket som mätvärde. Låt oss kalla detta **huvudregulatorn**. Den har börvärde 50 bar. Sedan skapar vi en **stödregulator** som har friblåsningsventilen som styrvariabel, och ångtrycket som mätvärde, även denna. Börvärdet för stödregulatorn kan vara 54 bar. Slutresultatet blir som i Figur 6.28, där PC1 är huvudregulator, och PC2 stöder med friblåsning.



Figur 6.28: Två regulatorer med samma ärvärde och olika börvärden.

Ett blockschema för en sådan reglering ser vi i Figur 6.29. För exemplet ovan skulle C_1 kunna vara turbinregulatorn och C_2 den regulator som använder friblåsningsventilen. Hur väl det hela fungerar beror på trimningsparametrar och processdynamik m.m. Värt att notera är att ovanstående lösning inte är någon garanti för att börvärde 2 inte överstigs.



Figur 6.29: Parallella regulatorer.

En möjlig fördel med ovannämnda lösning, jämfört med split-range, är att man inte behöver tänka igenom hur split-gränsen ska ligga. Å andra sidan måste man trimma två regulatorer...

En nackdel med parallellstrukturen är att om operatören vill köra regleringen i manuell så är det två regulatorer, eller ännu fler, som ska läggas i manuell. Det är lätt hänt att glömma någon, så det kan vara motiverat att bygga in logik som hanterar detta i styrsystemet.

En annan nackdel är att man är beroende av anti-windup-funktionen i regulatorerna. Stödregulatorn ligger för det mesta med utsignal u=0% eller

End Comment 4

Comment 5 (July 2024). One more split range control (SRC) scheme (Alternative 4) is shown in Figure 3 below:

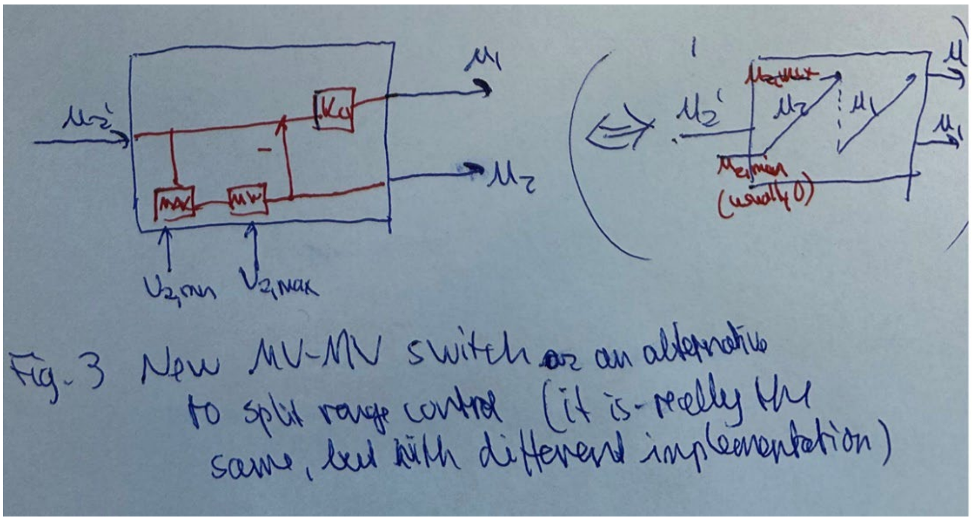


Fig. 3 New MV-MV switch as an alternative to split range control (it is really the same, but with different implementation)

This is not a really a new scheme, as it is really just another implementation of conventional SRC (see Fig. 3) and Shinskey has used it before (see below), and Evren Turan has rediscovered it (see Figure 2) and Sigurd added a little (to get Figure 3)

It requires a selector (to subtract the actual value of u2 from u2’ to get u1=u2’-u2) and thus it is very nice to combine it with cases where we anyway need a min-selector (see Shinskey and see Fig. 1/2 below)

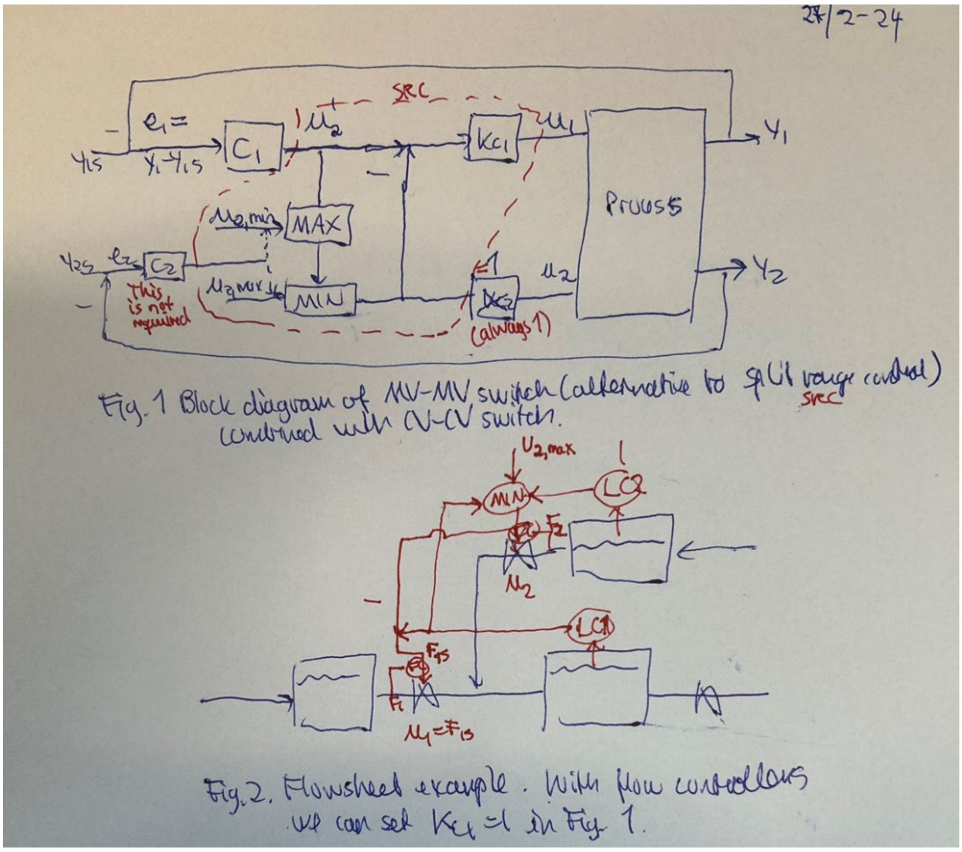


Fig. 1 Block diagram of MV-MV switch (alternative to split range control) combined with CV-CV switch.

Fig. 2. Flowsheet example. With flow controllers we can set $K_{c1}=1$ in Fig. 1.

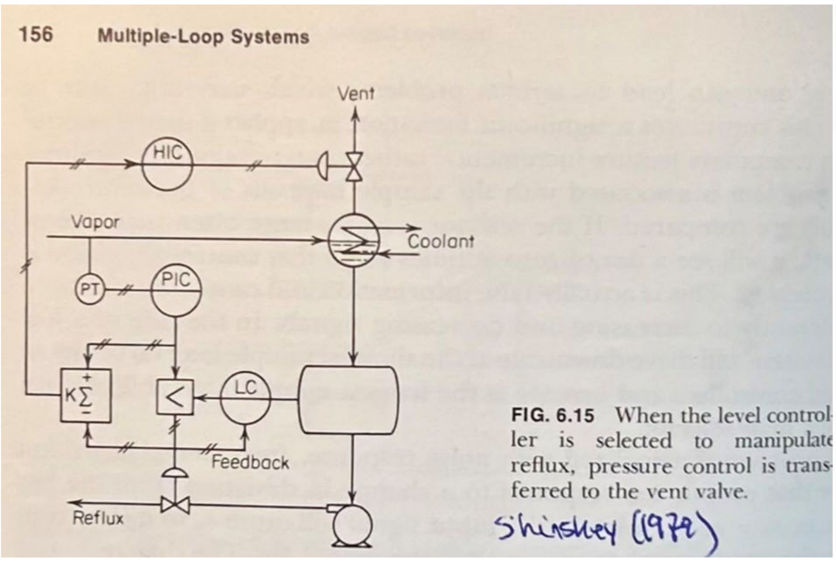


FIG. 6.15 When the level controller is selected to manipulate reflux, pressure control is transferred to the vent valve.

Here is from Reyes-Luas and Skogestad (2020) where we refer to Shinskey.

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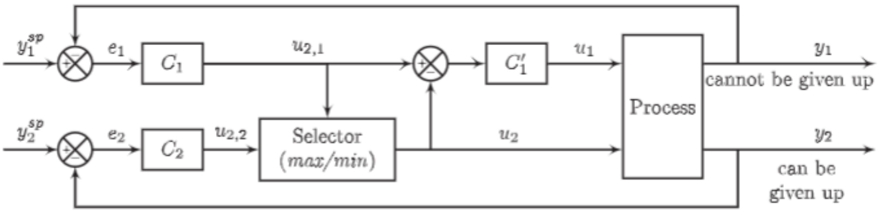


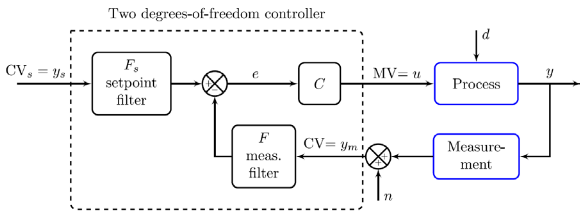
Figure 9. Alternative scheme for MV to CV switching when the input saturation rule is not followed.

An alternative solution from Shinskey[37] is shown in Figure 9. Here, controllers C_1 and C_2, for y_1 and y_2, are both designed for using u_2 as the input. We then have a selector for u_2, followed by a subtraction block that effectively does the split range control. Controller C_2 is used for controlling y_2 using u_2 as the input. C_2 needs antiwindup because u_2 is reassigned to controlling y_1 when u_1 saturates. Controller C_1, which controls y_1, is always active. It uses u_1 to control y_1 when u_1 is not saturated and switches to using u_2 when u_1 saturates. The "extra" control element for input u_1 (C_1' in Figure 9) can be just a gain, but it can also contain lead−lag dynamics. Note that the subtraction block in Figure 9 provides some built-in decoupling, which may be advantageous dynamically in the unconstrained case when both y_1 and y_2 are controlled.

End Comment 5.

Comment 6. (December 2024) On the choice of the filter time constant τ_F

Here is from the Appendix in the paper:



Here C is the feedback controller (e.g., PID), whereas F_s and F typically are lead–lag transfer functions, with a steady-state gain of 1. In process control, we often use $F = 1$ (no measurement filter) or a first-order filter,

$$F(s) = \frac{1}{\tau_F s + 1} \tag{A.3}$$

Here τ_F is the measurement filter time constant, and the inverse ($\omega_F = 1/\tau_F$) is known as the cutoff frequency. However, one should be careful about selecting a too large filter time constant τ_F as it acts as a effective delay as seen from the controller C; see also the recommendation $\tau_F \leq \tau_c/2$ in (C.17).

C.3.6. Measurement filter

For noisy processes, one may add a filter F on the measurement of y (Fig. A.41), for example, a first-order filter (A.3) (discrete (C.5)) or a Butterworth filter (A.4) with a tuneable time constant τ_F. To avoid that the filter adds too much lag to the control loop, one should choose

$$\tau_F \leq \tau_c/2 \tag{C.17}$$

Preferably, an even smaller value should be chosen.

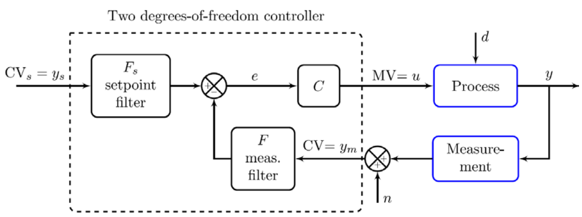
Simple analysis of choice of filter time constant τ_F.

Assume the loop transfer function is $L = C\ G\ G_m F = \frac{1}{\tau_C s}\frac{1}{\tau_F s+1}$

The closed-loop polynomial (set 1+L(s)=0) becomes $\tau_C\tau_F s^2 + \tau_C s + 1$.

- $\tau_F \leq 0.25\,\tau_C$: Real poles. $\tau_C\tau_F s^2 + \tau_C s + 1 = (\tau_1 s + 1)(\tau_2 s + 1)$
  - $\tau_F \ll \tau_C$: $\tau_1 \approx \tau_c$, $\tau_2 \approx \tau_F$
  - $\tau_F = 0.1\,\tau_C$: $\tau_1 = 0.89\,\tau_c$, $\tau_2 = 0.113\,\tau_C = 1.13\,\tau_F$
  - $\tau_F = 0.25\,\tau_C : \tau_1 = \tau_2 = 0.5\,\tau_C = 2\,\tau_F$
- $\tau_F > 0.25\,\tau_C$: Complex poles. $\tau_C\tau_F s^2 + \tau_C s + 1 = \tau^2 s^2 + 2\zeta\tau s + 1$
  - $\tau_F = 0.5\,\tau_C$: $\tau = 0.71\,\tau_c$, $\zeta = 0.71$
  - Note: this is the maximum recommended value for τ_F in (C.17), $\tau_F \leq \frac{\tau_C}{2}$, and the above derivation provides justification for (C.17).

Comment 7. (December 2024) Filter on D-action



Here C is the feedback controller (e.g., PID), whereas F_s and F typically are lead–lag transfer functions, with a steady-state gain of 1. In process control, we often use $F = 1$ (no measurement filter) or a first-order filter,

$$F(s) = \frac{1}{\tau_F s + 1} \quad \text{(A.3)}$$

Here τ_F is the measurement filter time constant, and the inverse ($\omega_F = 1/\tau_F$) is known as the cutoff frequency. However, one should be careful about selecting a too large filter time constant τ_F as it acts as a effective delay as seen from the controller C; see also the recommendation $\tau_F \leq \tau_c/2$ in (C.17).

Traditionally (in most older books), a first-order filter is put on the D-action (often with $\tau_F = \epsilon\tau_D$ with $\epsilon = 0.1$), but I recommend to use instead the above approach where the filter is on y (so also on the PI-part)

- For cascade PID it is equivalent (because here the filter multiplies the whole controller)
- But for ideal PID the filtering of only the D-action is difficult to interpret and not recommended (see papers by Hägglund).
  - So don't use $C(s) = K_c\left(1 + \frac{1}{\tau_I s} + \frac{\tau_D s}{\epsilon\tau_D s+1}\right)$
  - use instead $C(s) = \frac{K_c}{\tau_F s+1}\left(1 + \frac{1}{\tau_I s} + \tau_D s\right)$