Feedforward Control

The inherent limitation of feedback control is that it is retrospective. A feedback controller can only respond to disturbances once they have affected the controlled variable. For many processes this does not matter unduly. However, when the disturbances are large, or where the process dynamics are sluggish, feedback control results in significant and sustained errors. Using cascade control to reject specific disturbances can produce substantial improvements in performance. However, control is still retrospective. Ratio control is different. It responds to changes in one variable by adjusting another to keep them in proportion. In a sense it is anticipating the process needs and is a particular case of feedforward control.

What feedforward control offers is the prospect of control action which anticipates the effect of disturbances on the process and compensates for them in advance. This chapter develops the concept of feedforward control, considers some of its limitations, and introduces its implementation.

27.1 Feedforward Compensation

Consider a process and its load, with steady state gains of K_1 and K_2 respectively, as depicted in Figure 27.1. Suppose that the dynamics are negligible. Let θ_0, θ_1 and θ_2 be the controlled, disturbance and manipulated variables in deviation form.

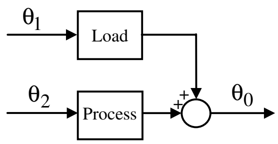

Fig. 27.1 Block diagram of process and its load

To compensate for changes in θ_1 a feedforward element of gain K_F may be introduced, as depicted in Figure 27.2.

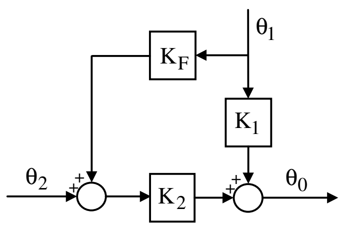

Fig. 27.2 Process and load with feedforward compensation

Steady state analysis yields:

$$\begin{aligned}\theta_0 &= K_1.\theta_1 + K_2.(K_F.\theta_1 + \theta_2)\\ &= (K_1 + K_2.K_F)\,\theta_1 + K_2.\theta_2\end{aligned}$$

For ideal disturbance rejection, changes in θ_1 have no effect on θ_0 such that

$$\theta_0 = K_2.\theta_2$$

To satisfy this criterion it is necessary for:

$$(K_1 + K_2.K_F)\,\theta_1 = 0$$

Since θ_1 cannot be assumed to be zero, it follows that:

$$K_F = -K_1/K_2$$

In effect, the feedforward path is creating an inverse signal which will cancel out the effect of the load operating on the disturbance. Any practical implementation of this feedforward compensation requires a measurement of θ_1 and some means of applying the compensation. Let K_T, K_I and K_V be the steady state gains of the measuring element, I/P converter and control valve as depicted in Figure 27.3. It is now appropriate to consider θ_2 as being the output of a conventional controller and F_2 as the manipulated variable.

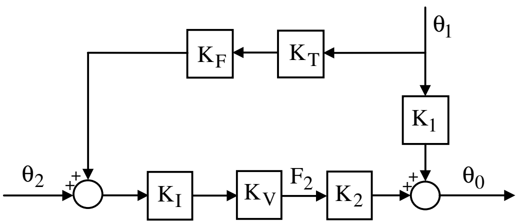

Fig. 27.3 Practical implementation of feedforward compensation

A similar steady state analysis to the above yields:

$$K_F = -K_1/\,(K_T.K_I.K_V.K_2) \qquad (27.1)$$

It is necessary that the values of these steady state gains are known. K_1 and K_2 may be determined either empirically or theoretically. K_T, K_I and K_V are known by specification and/or calibration. However, any inaccuracies will lead to a steady state offset in θ_0 in the event of a change in θ_1. The significance of this should not be underestimated.

27.2 Dynamic Compensation

It is unrealistic to ignore the dynamics of the process and load. Indeed, it is largely because of their dynamics that feedforward compensation is being considered. Whereas the gains of the various elements are relatively easy to establish, their dynamics are not. This is especially true of the process and load. Whilst the structure of their dynamic models may be determined from first principles, it is often difficult to predict the values of the parameters involved with any confidence. Therefore, in practice, it is usual to separate out the steady state gain from the dynamics, and to have two feedforward compensation terms as depicted in Figure 27.4.

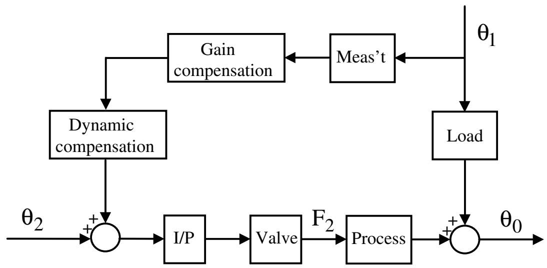

Fig. 27.4 Distinction between gain and dynamic compensation

The steady state feedforward gain is as defined by Equation 27.1. By a similar argument, assuming that the dynamics of the instrumentation is insignificant, the dynamic compensation term C(s) is the ratio of the dynamics of the load L(s) to those of the process P(s).

$$C(s) = \frac{L(s)}{P(s)}$$

The Laplace notation used here, necessary for articulating the dynamics, is explained in detail in Chapters 70 and 71. If the process and load dynamics are the same, which is often the case, they cancel and there is no need for dynamic compensation. Otherwise, the dynamic compensation term consists of time lags, leads and delay terms, such that its structure corresponds to the required dynamic ratio C(s). The parameters of the dynamic compensation term are then tuned empirically.

Because of the scope for offset due to errors in the values of the various steady state gains, and the approximate nature of the dynamic term, feedforward compensation is seldom used in isolation.

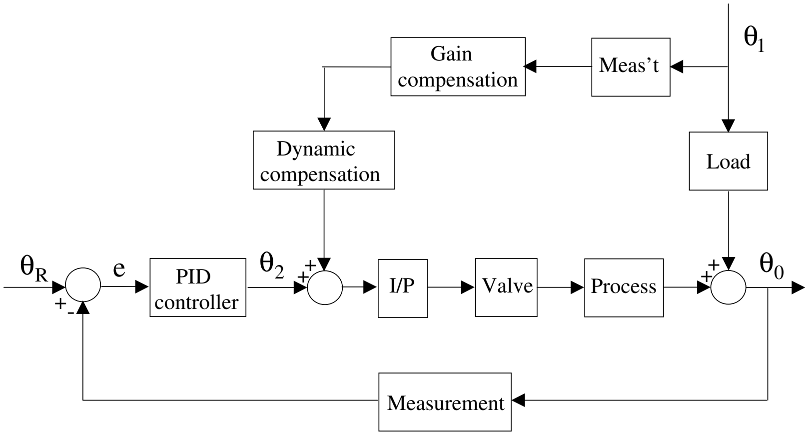

Fig. 27.5 Feedforward compensation grafted onto a PID loop

The most common strategy is to use it in conjunction with a conventional 3-term feedback control loop as depicted in Figure 27.5. The feedback loop will eliminate offset due to inaccuracies in the feedforward compensation, handle residual dynamic errors, and correct for other disturbances.

A practical example of the use of feedforward compensation in the control of a distillation column is depicted in Figure 27.6. It is used in conjunction with a cascade system which controls the composition in the bottom of the column by manipulating the flow of steam into the reboiler, steam pressure

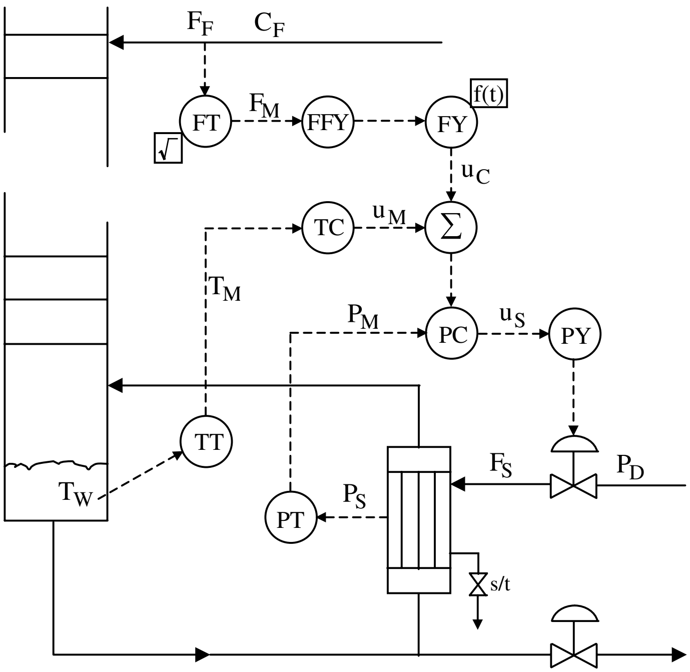

Fig. 27.6 P&I diagram of feedforward compensation for distillation column

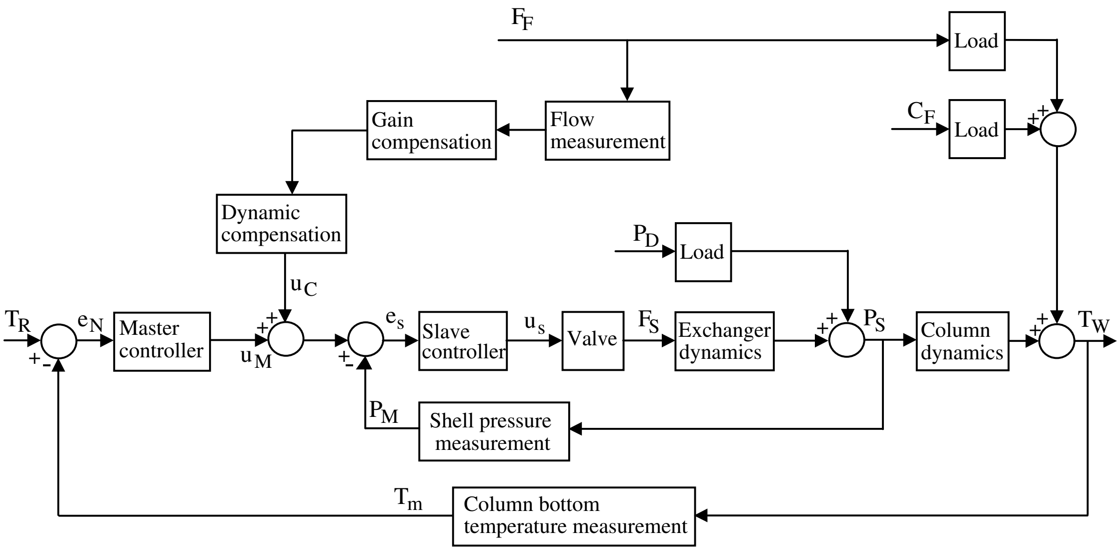

Fig. 27.7 Block diagram of feedforward compensation for distillation column

being the slave variable. The feedforward compensation varies the boil-up rate in anticipation of the effects of changes in the column feed rate by applying a bias to the set point of the slave loop. The corresponding block diagram is given in Figure 27.7.

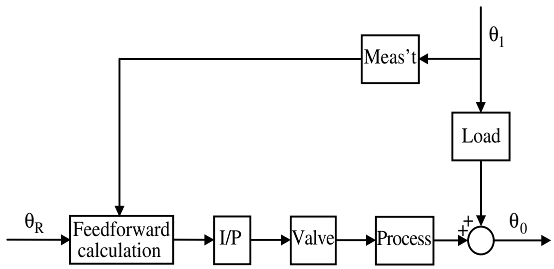

Fig. 27.8 Process and load with feedforward calculation

27.3 Feedforward Control

Alternatively, the feedforward control can incorporate the set point function, as depicted in Figure 27.8. The use of the set point as an input to the feedforward calculation is what distinguishes feedforward control from feedforward compensation. Indeed, it is the litmus test.

In practice, feedforward control as depicted in Figure 27.8 is seldom used in isolation because of the problems of offset and the need to handle other disturbances. The most common strategy is to use it in conjunction with a conventional 3-term feedback control loop as depicted in Figure 27.9. This feedback loop is analogous to the slave loop used in cascade control. Note that the steady state and dynamic compensation have again been separated out.

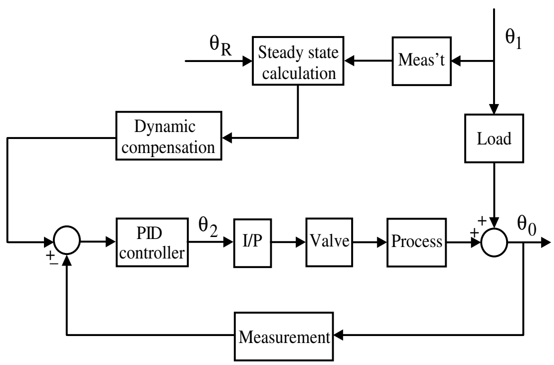

Fig. 27.9 Feedforward control grafted onto a PID controller

27.4 Feedforward Control of a Heat Exchanger

The practicalities of feedforward control are perhaps best illustrated by means of an example. Consider the heating up of a process stream on the tube side of an exchanger by the condensation of steam on its shell side, as depicted in Figure 27.10.

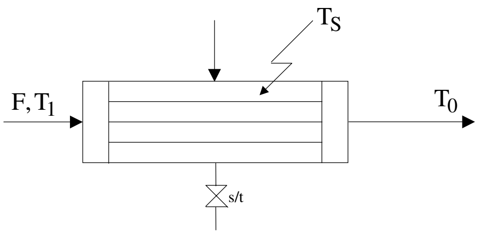

Fig. 27.10 Steam heated shell and tube exchanger

A steady state heat balance across the exchanger gives

$$Q = U.A.T_m = F.\rho.c_p\,(T_0 - T_1)$$

However, the log mean temperature difference is given by

$$T_m = \frac{T_0 - T_1}{\ln\left(\dfrac{T_S - T_1}{T_S - T_0}\right)}$$

Hence

$$\frac{T_S - T_1}{T_S - T_0} = \exp\left(\frac{UA}{F\rho c_p}\right)$$

Assume that all the resistance to heat transfer is due to the tube side film coefficient. From the Dittus Boelter correlation, the overall coefficient may be approximated by

$$U \approx kF^{0.8}$$

Also, for saturated steam and water,

$$T_S \approx mP_S$$

Hence

$$P_S \approx \frac{T_1 - T_0.\exp\left(\dfrac{kA}{\rho c_p F^{0.2}}\right)}{m\left[1 - \exp\left(\dfrac{kA}{\rho c_p F^{0.2}}\right)\right]}$$

This equation is, in effect, the steady state model of the process. For implementation as a feedforward controller, the outlet temperature T_0, which is arbitrary, may be replaced by its desired value T_R. The inlet temperature T_1 and flow rate F may be replaced by their measured values T_M and F_M respectively. The equation explicitly calculates the

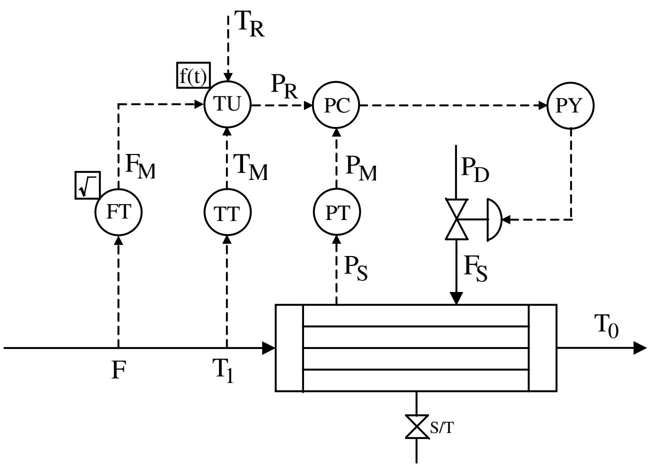

Fig. 27.11 P&I diagram of feedforward control of heat exchanger

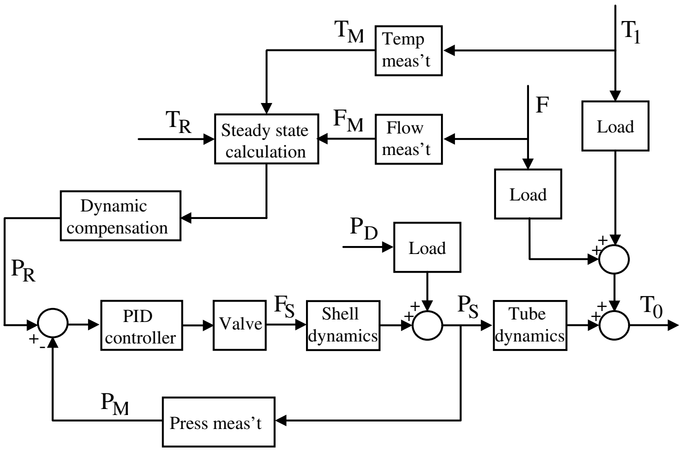

Fig. 27.12 Block diagram of feedforward control of heat exchanger

necessary steam pressure, which becomes the set point P_R for a conventional feedback control loop:

$$P_R \approx \frac{T_M - T_R \exp\left(\dfrac{kA}{\rho c_p F_M{}^{0.2}}\right)}{m\left[1 - \exp\left(\dfrac{kA}{\rho c_p F_M{}^{0.2}}\right)\right]} \tag{27.2}$$

This control scheme is shown in P&I diagram form in Figure 27.11 and the corresponding block diagram is shown in Figure 27.12.

27.5 Implementation Issues

Inaccuracy is the prime source of difficulty in implementing feedforward control. Any errors in the temperature and flow measurements will be propagated into the derived steam pressure set point through Equation 27.2. More fundamental though is the accuracy of Equation 27.2 itself. A model of the process has been developed. Various assumptions and approximations have been made. Even if

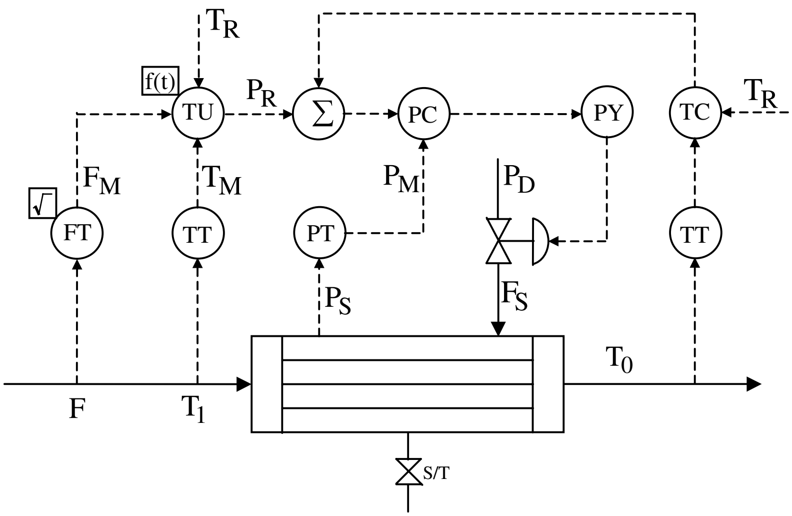

Fig. 27.13 P&I diagram of feedforward control with set point trimming

these are all correct, it is unlikely that accurate values are available for the parameters of the model. Furthermore, the model is steady state and ignores the dynamics of the process. Dynamic compensation by means of time lags, leads and delay terms is, at best, approximate.

There is likely, therefore, to be significant offset in the outlet temperature. This is best handled by another controller which trims the system. Trimming can be achieved by various means: in this case a bias is applied to the set point of the steam pressure control loop, as shown in Figure 27.13.

27.6 Comments

Feedforward control is not an easy option. Developing the model requires both experience and understanding of the process. There are major problems due to inaccuracy. Only specific disturbances are rejected. This results in feedforward control having to be used in conjunction with other loops. The outcome is that the control schemes are complex, some would say unnecessarily so. Finding the optimum form of dynamic compensation and tuning the loops is not easy. Nevertheless, it does work and does produce benefits. There are many feedforward control schemes in operation throughout the process industries. Modern control systems support all the functionality necessary for their implementation.

27.7 Nomenclature

A	mean surface area of tubes	m^2
c_p	specific heat	$kJ\ kg^{-1}\ K^{-1}$
F	flow rate	$m^3\ s^{-1}$
k	coefficient	$kJ\ s^{-0.2}\ m^{-4.4}\ K^{-1}$
m	coefficient	$^\circ C\ bar^{-1}$
ρ	density	$kg\ m^{-3}$
P	pressure	bar
Q	rate of heat transfer	kW
T	temperature	$^\circ C$
U	overall heat transfer coefficient	$kW\ m^{-2}\ K^{-1}$

Subscripts

M	measured
m	logarithmic mean
R	reference
S	shell side steam
1	tube side inlet
0	tube side outlet