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Cascade Control

This chapter summarises the essential principles and practice of cascade control: it is discussed more fully by Murrill (1988). Cascade control is a powerful extension of conventional 3-term feedback control. It is a strategy which compensates for specific disturbances at source and largely prevents them from affecting the process being controlled.

25.1 Master and Slave Loops

A cascade control scheme has two controllers, the output of the master controller being used to adjust the set point of the slave controller. This is best illustrated by means of an example. Consider the boiler level control system of Figure 25.1.

Boiler drum level is notoriously difficult to control. One reason for this is that the steam pressure in the drum can vary significantly. Suppose the pressure in the vapour space above the liquid in the drum suddenly increases. This will cause the pressure drop across the control valve to fall. Thus the water flow will be reduced, irrespective of the level, even if the level is below its set point. The effect of steam pressure disturbances can be compensated for using a slave flow control loop as depicted in Figure 25.2.

The set point of the flow control loop u_M is manipulated by the level controller output. If the level h is too low the set point of the flow loop will be increased, and *vice versa*. The flow loop controls the flow of water F_W against this set point in the normal way. Again, suppose that the pressure in the drum P_S suddenly increases, perhaps due to a transient drop in steam load F_S, causing the flow of water to decrease. The flow loop will respond quickly by opening the valve to maintain the water flow F_W at the rate demanded by the level controller. In effect, the flow loop is insulating the level loop from, or re-

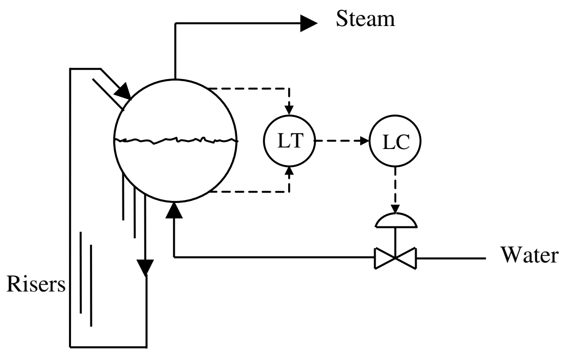

Fig. 25.1 Boiler drum level control with simple feedback

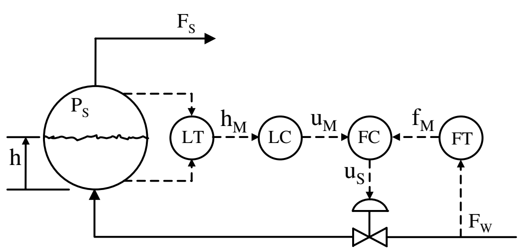

Fig. 25.2 Boiler drum level control with cascade system

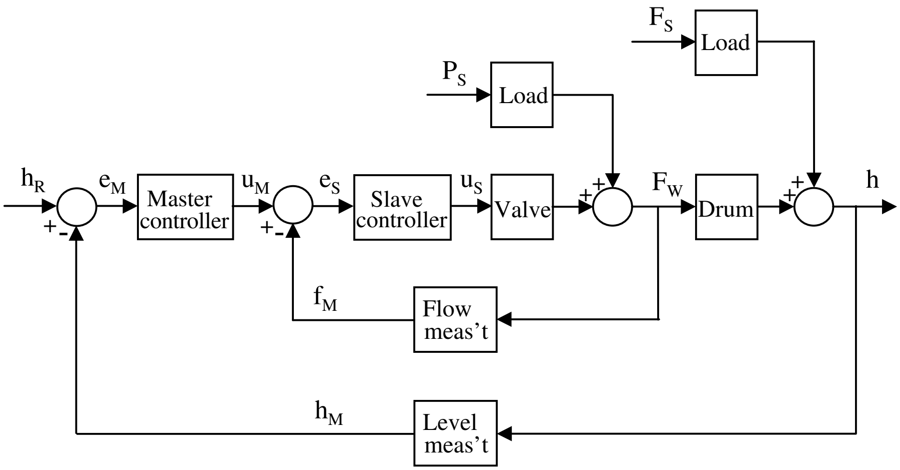

Fig. 25.3 Block diagram of system for cascade control of drum level

jecting disturbances due to, changes in steam pressure. The corresponding block diagram is given in Figure 25.3.

Its structure consists of two feedback loops, one nested inside the other. The outer loop is the master loop, sometimes referred to as the primary loop, and controls the level. The inner loop is the slave loop, sometimes referred to as the secondary loop, and controls the flow. The flow loop has a much faster response than the level loop, the dynamics of the later being dominated by the lags due to the capacity of the drum. It is often the case that the slave loop is a flow control loop. The need for the inner loop to have a faster response than the outer loop is characteristic of all cascade control schemes.

Note in particular the location of the disturbances. Changes in P_S affect the slave loop which compensates for them before they have any significant effect on the drum level. Changes in F_S affect the level and are compensated for by the master loop. Strictly speaking, there is a linkage between changes in F_S and P_S: the details of this are omitted here for simplicity.

25.2 Cascade Control of Heat Exchanger

Another example of the application of cascade control is in the control of heat exchangers. Consider Figure 25.4 in which a process liquor is heated on the tube side of an exchanger by condensing steam on the shell side. A conventional feedback control scheme uses the outlet temperature T_0 to manipulate the flow rate F_S of steam.

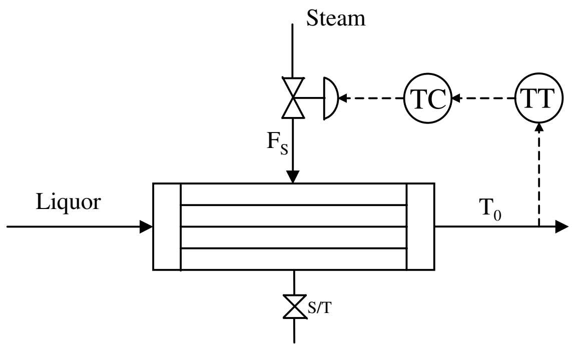

Fig. 25.4 Exchanger outlet temperature control with simple feedback

This scheme works well, but can be improved by cascade control. In particular, changes in the steam supply pressure affect the flow through the control valve which in turn affects the pressure inside the shell. Use of steam flow as a slave control loop would work, but steam temperature inside the shell would be much more effective as a slave loop since this relates directly to the rate of heat transfer: ultimately this is the manipulated variable that matters. Noting that the steam pressure is directly related to its temperature in condensing systems, it is logical to use a pressure loop instead of a temperature loop as the slave, as depicted in Figure 25.5. That is for two reasons: first, the rate of heat transfer is significantly more sensitive to the steam's pressure than its temperature. And second, given any uncertainty over the nature of the steam quality and its measurement, pressure is the more reliable metric. Note that the slave loop rejects disturbances due to steam supply pressure only, disturbances due to changes in the supply pressure of the process liquor and its temperature are handled by the master loop.

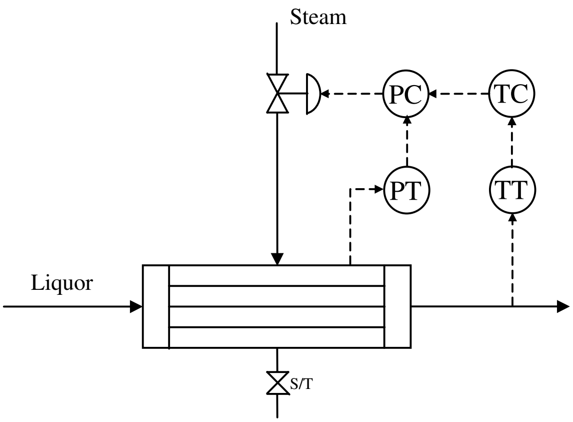

Fig. 25.5 Exchanger outlet temperature control with cascade system

25.3 Cascade Control of Jacketed Reactor

The choice of slave variable is not always straightforward. Consider the jacketed stirred tank reactor of Figure 25.6. Reagents flow into the reactor and displace products through the overflow at the same rate. The reaction is exothermic, heat being removed by cooling water being circulated through the jacket. Reactor temperature is controlled by a conventional feedback loop which manipulates the flow of water.

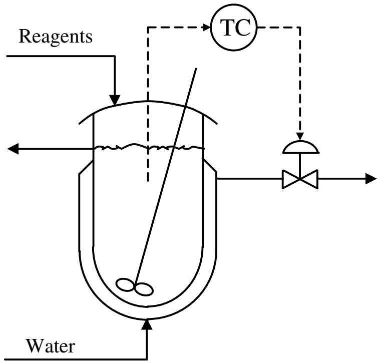

Fig. 25.6 Simple feedback of temperature in jacketed reactor

This scheme has a poor response because of the sluggish dynamics of both the reactor and its jacket. In particular, any disturbance in the cooling water supply pressure will cause a change in jacket temperature which, eventually, will affect the reactor temperature. Only when the reactor temperature moves away from its set point can the controller start to compensate for the disturbance. Significant errors occur before compensation is complete. This is a classical application for cascade control. There is a choice of three slave loops using either the water flowrate, pressure or temperature as the controlled variable, as depicted in Figures 25.7, 25.8 and 25.9 respectively.

- Flowrate: use of water flowrate as the slave variable is essentially the same as in the example of the boiler drum level control scheme of Figure 25.2. The slave loop specifically rejects disturbances in water flow rate due to changes in its supply pressure. The slave loop has a fast response since it is dependant only upon the hydrodynamics of the water system and the dynamics of the instrumentation and valve.
- Pressure: use of the jacket pressure as the slave variable also specifically rejects disturbances in water flow rate and has a fast response. How-

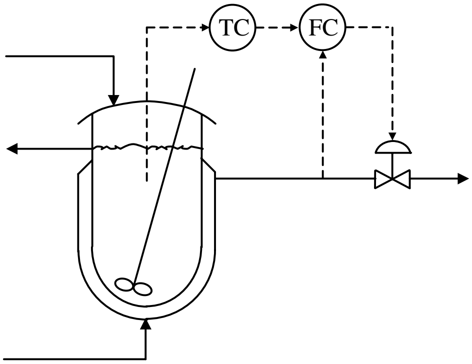

Fig. 25.7 Cascade control of reactor: jacket water flowrate as slave loop

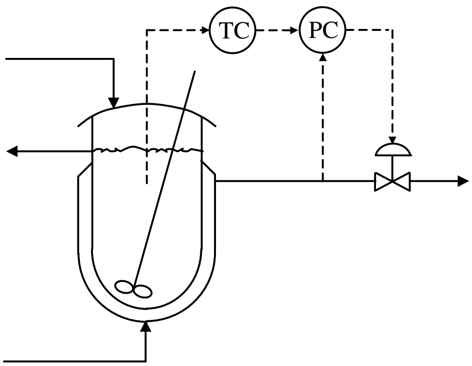

Fig. 25.8 Cascade control of reactor: jacket water pressure as slave loop

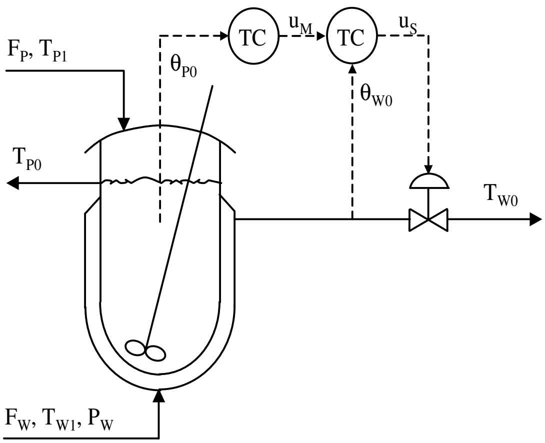

Fig. 25.9 Cascade control of reactor: jacket water temperature as slave loop

ever, the scheme is counter-intuitive and could confuse the operators:

i. The slave controller must be reverse acting. For example, following an increase in supply pressure the jacket pressure rises and the controller opens the valve. Assuming most of the resistance to flow is due to the jacket and the pipework upstream of the valve, opening the valve increases the flow and the pressure drop across the fixed resistances. Hence the jacket pressure falls. One would intuitively expect to close the valve to counter the effect of an increase in supply pressure.
ii. The master loop must be forward acting. Thus, following an increase in reactor temperature, the master controller reduces the set point of the slave controller. This results in the valve opening thereby thereby giving the increase in cooling water necessary to counter the temperature.

It would be bad practice to use this scheme given that there are viable alternatives.

- Temperature: as can be seen from the corresponding block diagram of Figure 25.10, use of jacket temperature T_{W0} as the slave variable rejects disturbances in water temperature T_{W1} as well as disturbances in water flowrate F_W due to changes in supply pressure P_W. Disturbances in the temperature T_{P1} and flowrate F_P of the reagents feed stream are handled by the master loop.

It is evident that the dynamics of the plant have been split, the reactor is still in the master loop but the jacket has been shifted into the slave loop. The response of this slave loop is relatively slow since it is dominated by the thermodynamics of the jacket. However, this is not necessarily a disadvantage, provided the slave loop's response is still faster than that of the master. Indeed, shifting the jacket's dynamics into the slave loop makes for better control of the master loop. Note that there are important interactions between the reactor and jacket so, strictly speaking, they cannot be split as described. The details of this are omitted here for simplicity, but are considered quantitatively in Chapter 85.

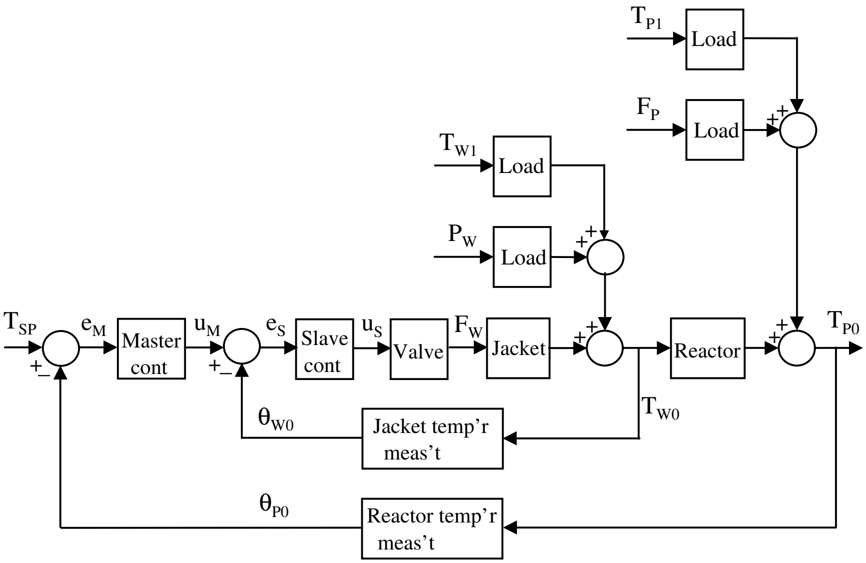

Fig. 25.10 Block diagram of cascade system with jacket temperature as slave loop

25.4 Implementation

Without doubt cascade control can bring about substantial improvements in the quality of control. However, the benefits are critically dependant upon proper implementation. The control scheme should be designed to target specific disturbances. If possible choose slave loops that reject disturbances from more than one source. It is essential that the slave loop's dynamics are significantly faster the master loop's to minimise the effects of interactions between the loops. As a rule of thumb, the dominant time constant in the slave loop must be less than one third of the dominant time constant in the master loop.

Because of the interactions between the loops, any non-linearity introduced by the slave loop will have an adverse effect on the behaviour of the master loop. The slave loop should not introduce non-linearity. It is particularly important when the slave loop is controlling flow that square root extraction is used.

When tuning the controllers, the basic strategy is to tune the inner loop first. Switching the master controller into its manual mode effectively disconnects the two loops. By applying step changes to the master controller output, the slave loop can be tuned using the continuous cycling method. Alternatively, by switching the slave loop into its manual mode, it can be tuned by the reaction curve method. Once the slave loop has been tuned, it can be switched into its automatic mode and the master loop tuned in the conventional way. In effect, the slave loop is treated as if it were any other element whilst tuning the master loop. It may be necessary to slightly detune the slave loop if its dynamics propagate into the master loop to the extent that the interactions adversely affect the master loop's performance.

If analogue single loop controllers are being used, remember that the slave controller is physically different to the master controller. The slave controller requires an input for a remote, or external, set point signal whereas the master controller's set point is local and adjusted manually.

It would appear that cascade control is more expensive than conventional feedback. It certainly requires an additional slave measurement, although it is often the case that the measurement is probably being made anyway, or ought to be if the disturbances are significant. Otherwise, the only additional costs are for an extra analogue input channel and additional function blocks. If a DCS, PLC or modern digital SLC is being used, it will support all of the functionality necessary for handling both the master and slave controllers. As far as the output channel is concerned, the same I/P converter, valve and actuator are used for cascade control as would be otherwise. So there truly is relatively little extra cost.

Ratio Control

Chapter 26

Ratio control is another control strategy commonly used in the process industries. It is used when the flow rates of two or more streams must be held in proportion to each other. Typical applications are in blending , combustion and reactor feed control systems. There are essentially three approaches, one is based upon the simple scaling of signals and the other two are based upon the PID controller. In the latter cases, the ratio may be controlled either directly using a proprietary ratio controller or else indirectly by means of a ratio station with a conventional controller. All three approaches are discussed. A good treatment of ratio control is given by Shinsky (1996).

26.1 Scaling Approach

Consider Figure 26.1 in which stream A is wild and stream B is manipulated to keep it in proportion to stream A. Assume that the flow transmitter is calibrated for the full range of the manipulated flow. Also assume that the valve is carefully sized such that its full range of flow corresponds to that of the manipulated flow. The flow transmitter output may be applied *via* an I/P converter directly to the valve to achieve the desired ratio control. This is a simple and effective means of ratio control, but is critically dependant upon the linearity of the elements. The flow transmitter may require square root extraction, and the control valve must have a linear installed characteristic.

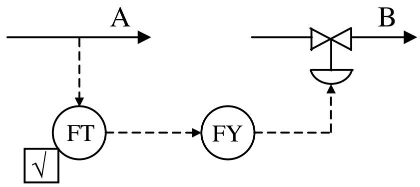

Fig. 26.1 Simple scaling approach to ratio control

In practice it is unlikely that the valve could be sized such that the ranges of the wild and manipulated flows are exactly in the desired ratio. A scaling factor is therefore necessary. This can be realised by changing the calibration of the flow transmitter, or by fitting a positioner to the valve and adjusting its range. An alternative is to use a so-called ratio station, as depicted in Figure 26.2.

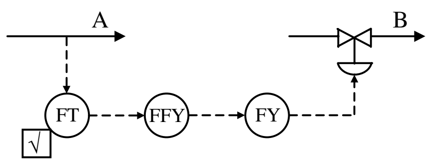

Fig. 26.2 Scaling approach with a ratio station

The ratio station, denoted by the code FFY, is simply a device for applying a user definable scaling factor K to a signal. For example, if the station's I/O are 4–20 mA signals, then its operation is described by:

$$\theta_0 = 4 + K\,(\theta_1 - 4)$$

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Care has to be taken in deciding what scaling factor to apply. It depends on the calibrations of the transmitter, valve and/or positioner. Also, if the signals are within a digital system, they may well have been scaled into engineering units too.

26.2 Direct Approach

An example of the direct approach is given in Figure 26.3. Again stream A is assumed to be wild. Both flow rates are measured. The ratio controller FFC manipulates the flowrate of stream B to produce the desired ratio of B to A. Note that the ratio control loop rejects disturbances in stream B due to changes in supply pressure P_S.

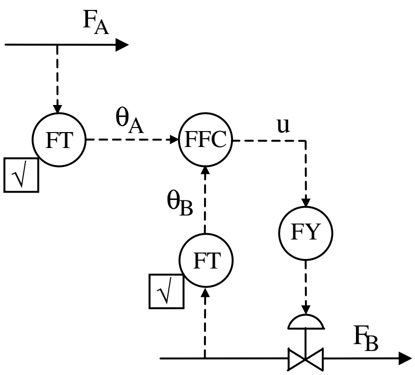

Fig. 26.3 Direct approach to ratio control

Assuming 4–20 mA signals, the ratio is calculated according to:

$$R_M = \frac{(\theta_B - 4)}{(\theta_A - 4)}$$

This measured ratio is then compared with the desired (reference) ratio R_R and an error signal e is generated. The ratio controller operates on the error to produce an output signal u. The block diagram is given in Figure 26.4.

Most proprietary ratio controllers physically combine the division, comparison and control functions into a single unit. Thus a ratio controller has two measured value inputs and one output signal. Typically the ratio of the measured values is displayed on the faceplate alongside the manually set desired ratio. In most other respects a ratio controller is much the same as a conventional 3-term controller. For example, it provides P, I and D actions, has a forward/reverse action switch and supports both automatic and manual modes of operation.

To the extent that each involves two measured values and one output signal, it is easy to confuse the P&I diagrams of ratio and cascade control schemes. For example, notwithstanding the fact that the processes are different, the ratio scheme of Figure 26.3 looks similar to the cascade scheme of Figure 25.2. However, comparison of the corresponding block diagrams reveals that their structures are fundamentally different.

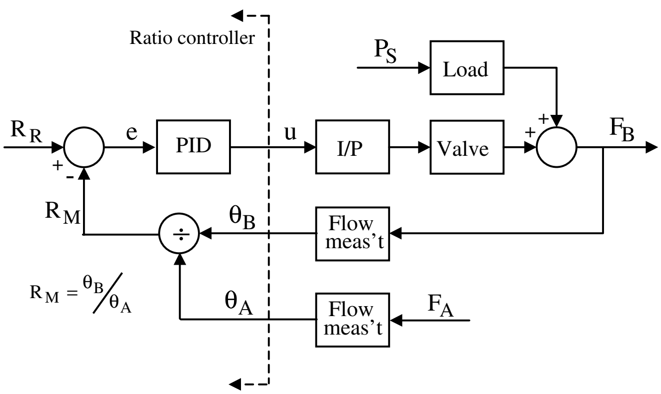

Fig. 26.4 Block diagram of direct approach to ratio control

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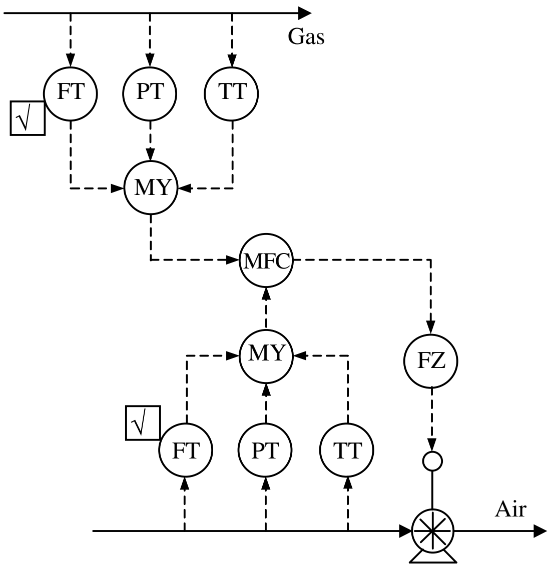

Fig. 26.5 Direct approach with temperature and pressure compensation

Ratio control of gas flows subject to changes in operating conditions may well require pressure and/or temperature correction, as depicted in Figure 26.5. In these circumstances it is best to calculate the mass flow rate of each stream, and to control the ratio of the mass flow rates. However, bearing in mind the scope for calibration and measurement errors, and their potential impact on the ratio calculated, the flow changes must be fairly significant to justify the increased complexity.

26.3 Indirect Approach

The indirect approach is both simple and effective, an example is given in Figure 26.6.

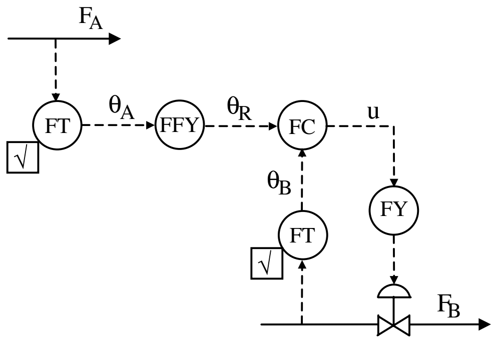

Fig. 26.6 Indirect approach to ratio control

The measured value θ_A of the flowrate of wild stream A is operated on by the ratio station. Assuming 4–20 mA signals it calculates the desired value θ_R for the flowrate of stream B according to the equation:

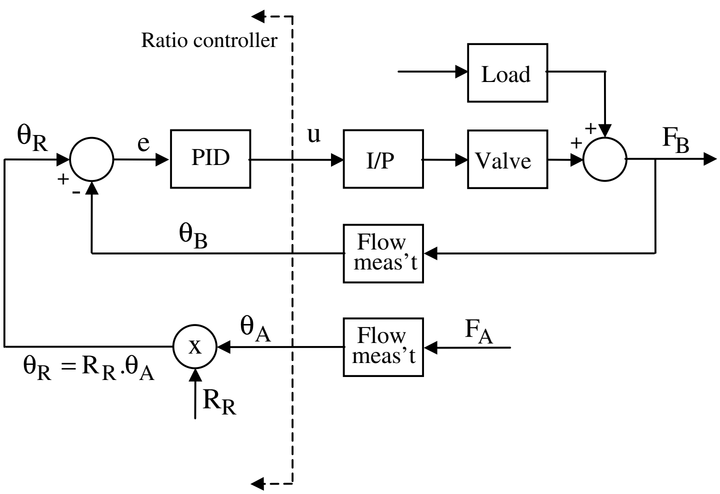

Fig. 26.7 Block diagram of indirect approach to ratio control

$$\theta_R = 4 + R_R\,(\theta_A - 4)$$

where R_R is the desired (reference) ratio. A conventional feedback loop is then used to control the flowrate of stream B against this set point, as depicted in the block diagram in Figure 26.7. Note again the rejection of disturbances in stream B.

26.4 Comments

The direct approach is commonly used throughout industry although, on two different counts, the indirect method is superior. First, with the direct ratio scheme there is the potential for zero division by θ_A giving rise to indeterminate ratios. This is particularly problematic with low flows: errors in the flow measurement become disproportionate and lead to very poor quality ratio control. This can also be a problem if ratio control is used as part of a model based control strategy involving deviation variables. The problem does not arise with the indirect ratio scheme because θ_A is multiplied.

Second, there is the question of sensitivity of the error to changes in the wild flow. Inspection of Figure 26.4 reveals that the sensitivity of the direct scheme is:

$$\left.\frac{de}{d\theta_A}\right|_{\overline{\theta}_B} = \left.\frac{d}{d\theta_A}\left(R_R - \frac{\theta_B}{\theta_A}\right)\right|_{\overline{\theta}_B} = \frac{\theta_B}{\theta_A^2} = \frac{R_m}{\theta_A}$$

whereas inspection of Figure 26.7 reveals that the sensitivity of the indirect scheme is:

$$\left.\frac{de}{d\theta_A}\right|_{\overline{\theta}_B} = \left.\frac{d}{d\theta_A}\left(R_R.\theta_A - \theta_B\right)\right|_{\overline{\theta}_B} = R_R$$

Clearly there is a linearity issue. The lower the flow the greater the sensitivity of the direct ratio scheme, whereas the indirect scheme's sensitivity is constant.