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AN INTRODUCTION TO PROCESS CONTROL SYSTEM DESIGN

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Ch. 18: Has cascade, parallell and ratio control

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CHAPTER 18

COMPLEX CONTROL SYSTEMS

18.1 INTRODUCTION

It has been seen that the control quality obtainable with a single loop system is determined by the characteristics of the plant and by the nature, magnitude and point of entry of the disturbances. It is common practice to minimise disturbances in as many of the operating conditions as possible by installing independent control systems, as illustrated by Fig. 2.1. Sometimes, however, large disturbances cannot be prevented from entering the main control system which may, therefore, be unable to hold the deviation within the specified limits, particularly when the time lag produced by the plant is large or when the disturbances enter near the detecting element. In these circumstances it is often profitable to employ more complex systems, which contain additional measuring, controlling or regulating units as in the examples discussed in the following sections. Clearly, a complex system is used only if the control quality which can be obtained from a simple single loop system is unsatisfactory.

The purpose of this chapter is to discuss the operation of some types of complex systems. The cascade system is discussed most fully, because it is the most commonly used. A knowledge of the basic principles which govern the operation of the systems described will assist in reading the published descriptions of specific applications, and in selecting and adjusting the system to be used in a given application. The papers to which reference is made contain the results of long experience in the field and should be consulted.

It will be clear that the systems described represent a step towards fully integrated control systems, in which each condition affecting the process is measured and maintained, by a single master controller, at the optimal value relative to that of every other condition. Such systems will soon be installed to control processes which demand more consistent operating conditions than can be obtained with the systems in current use, but for some time the majority of plants will not employ systems more complicated than those described below, namely:

(1) Disturbance-feedback.
(2) Cascade.
(3) Systems which employ one measuring unit for the adjustment of two correcting units.

(4) Systems which employ two independent controlling units for the adjustment of one correcting unit.

A note is appended on Ratio and Averaging Systems, which are both widely used in practice. They are not necessarily complex systems since in their simplest forms neither employs more than one measuring, controlling or correcting unit.

18.2 DISTURBANCE-FEEDBACK

It was explained in Section 3.4.2 that for successful manual operation of a plant exhibiting *distance/velocity lag*, the instruments must be installed in order to indicate to the operator changes in supply conditions *before* they enter the plant. Otherwise, no corrective action can be taken until the effect of the disturbance is indicated by the detecting element at the plant output. It will now be appreciated that it will also be an advantage when the plant exhibits *transfer* lag to provide indications of changes in supply conditions to the operator, if these conditions are not separately controlled. The same considerations apply when automatic control is employed. When measurements of supply conditions (or other operating conditions) are supplied to assist the controller, the control system is said to be a 'disturbance-feedback system'.

This system is most frequently used when a disturbance occurs in a condition which can be measured but for some reason cannot be controlled. For example, the plant may have to accept the total production of a previous plant as one of its raw materials. The alternative to accepting a disturbance in supply rate is to install storage capacity between the two plants.

This involves unnecessary expenditure if, by a suitable arrangement of control equipment, control can be effected satisfactorily in spite of large supply disturbances. The provision of storage capacity between plants may result in considerable additional running costs as well as in additional capital expenditure; for example, the product of the previous plant may cool in the storage vessel and may have to be re-heated in the following plant.

The disturbance-feedback system is arranged as shown in Fig. 18.1. As already explained, its object is to provide corrective action immediately a disturbance appears, instead of waiting for the effect of the disturbance to pass through the plant and to be indicated by the detecting element (D.E.1) at the plant output. The disturbance is measured at X, by the detecting element (D.E.2), before it enters the plant, and an appropriate signal is generated by the unit Y and transmitted to unit Z, in which it is added to the output signal from the controlling unit (C.U.).

The full lines in Fig. 18.1 show the arrangement of the equipment when the disturbance enters at W on the demand-side (or when the disturbance occurs in a subsidiary supply condition). It was seen in Chapter 17 that the D.R.F. given by a single loop system is proportional to the attenuation (A_1) suffered by a signal in passing from its point of entry to the detecting element. When A_1 is too small to permit a satisfactory D.R.F. with a single loop system, disturbance-feedback can be profitably used to assist the controller to keep deviations within the specified limits in spite of the disturbances which must be accepted.

The advantage of the disturbance-feedback system increases *up to a point* as the disturbance enters the plant later in the loop, but it clearly cannot result in satisfactory control when A_1 becomes small —as when the point of entry (W) is near to the detecting element or when the time lag in the plant between W and the detecting element is mainly due to distance/velocity lag.

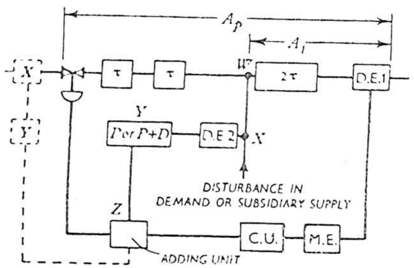

FIG. 18.1. Disturbance-feedback control system

The unit Y is usually arranged to give proportional + derivative action. The proportional action is adjusted so that the potential correction due to the output signal from Y is equal to the potential deviation due to the disturbance. Thus sustained changes in the uncontrolled condition produce an exactly compensating change in the correcting condition. The derivative action time is adjusted so that the lag experienced by a signal passing from X to detecting element D.E.1 through the correcting unit is equal to the lag suffered by the disturbance in passing from X to D.E.1 direct.

The time lag suffered by the disturbance in passing from X to W will assist control, since it will reduce the delay between the arrival of the disturbance and of the corrective signal at the detecting element. If this lag is a transfer lag the attenuation between X and W will also have the advantage of attenuating transitory or cyclic disturbances before they reach the plant.

Fig. 18.2 gives recovery curves calculated for a plant equivalent to the system shown in Fig. 18.1, for a single loop control system (curve (a)) and for a disturbance-feedback system (curves (b) and (c)). In all cases the controlling unit was used with P + I action. The equivalent system of the plant consists of three transfer stages with time constants τ, τ and 2τ; the disturbance enters between the second and

D.R.F. = ? Disturbance response factor?

third transfer stages. Curves (b) and (c) give the recoveries when the unit Y has proportional action and proportional + derivative action respectively ($R = 2\tau$).

Comparison of curves (a), (b) and (c) shows that the effect of *proportional* disturbance-feedback is not very large in this case, but that *proportional + derivative* disturbance-feedback results in a very considerable increase in D.R.F.

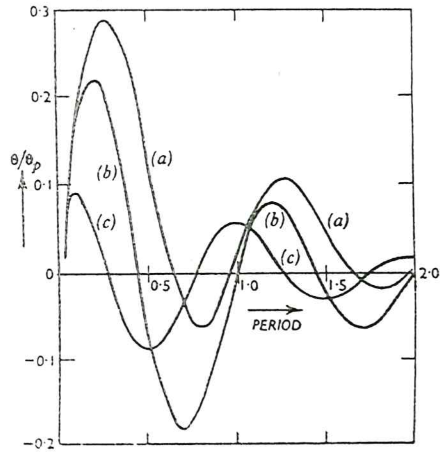

FIG. 18.2. Recovery of system of Fig. 18.1. with
(a) no disturbance-feedback
(b) P disturbance-feedback
(c) P + D disturbance-feedback.

Controller Adjustment—When setting up a disturbance-feedback system the controller should be adjusted to give μ_{max} with e:1 damping. The additional measuring element X, the measuring and transmitting element Y and the adding unit Z do not form another closed loop system, and therefore the response of the complete system to disturbances other than those measured at X remains the same as for the single loop system. The settings of the controller can therefore be found, as explained in Chapter 10, by using the gain/phase diagram of the controller and the phase-lag/attenuation diagram of the plant.

Feedback Adjustment—The proportional action factor of unit Y is adjusted, as explained above, so that its output signal corresponding to a step disturbance (θ) at X produces a potential correction θ_p—where θ_p = potential deviation due to θ.

The derivative action time of unit Y is set equal to $\tau' + \tau'' + \tau''' \ldots + \tau^n$ * where $\tau', \tau'', \tau''' \ldots \tau^n$ are the time constants of exponential transfer stages equivalent to the system composed of D.E.2, the plant between the correcting unit and W, and the transmission lines between D.E.2 and the correcting unit.†

* $\tau' + \tau'' + \tau''' + \ldots + \tau^n = L$; the value of $L - 2\tau$ for the system composed of D.E.2, Y, Z, and the correcting unit motor in Fig. 18.1 can be found by measuring the response of the motor to a ramp signal injected at X (see Sections 4.3 and 14.3.2). For this purpose Y must be adjusted to give proportional action only.

† This is a practical compromise, since ideally n (P + D) units with derivative action times $\tau', \tau'', \tau''' \ldots \tau^n$ should be used to 'compensate' for the lag due to these n exponential stages with the time constants $\tau', \tau'', \tau''', , , \tau^n$.

In the example taken the measuring and transmission lag are assumed to be negligible and therefore the derivative action time was set equal to $(\tau + \tau) = 2\tau$ (see Fig. 18.1).

The disturbance-feedback system has been discussed fully by Porter (62).

Practical examples, selected to show the advantages and disadvantages of the disturbance-feedback system compared with the cascade system are given in detail by Toop (5).

Note: (i) The disturbance-feedback system is likely to be most useful when major disturbances enter the plant at a point such that A_1 is about $\sqrt{A_P}$.

(ii) The system could be used to assist the controller when major disturbances occur in the main supply, using the arrangement shown by the broken lines in Fig. 18.1, but in general a cascade system is used for this purpose.

18.3 CASCADE SYSTEM

This system generally has the purpose of improving upon the control quality obtainable with a single loop system when the major disturbances enter through or near the correcting element. The controller (C) (Fig. 18.3) resets the desired value of the subsidiary controller (C′), instead of positioning the correcting unit directly as in the single loop system. The function of controller C′ is to reduce as far as possible the effect of supply disturbances.

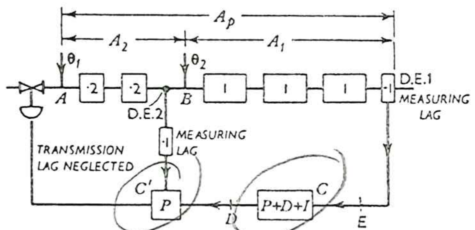

FIG. 18.3. Cascade control system.

The subsidiary controller decreases the operating period of the whole system and therefore has the effect of producing phase advance in the control loop; this is equivalent to increasing the derivative action time of the main controller.*

In adjusting the system it is essential to make $T > 3T'$ in order to avoid resonance between the main and subsidiary loops, where T, T′ are the operating periods of the two loops respectively. This usually presents no difficulty because the subsidiary loop is normally arranged to contain only a small part of the plant which is in the main loop. This implies that T′ must be small in order that the phase lag produced by the transfer stages in the subsidiary loop is equal to $(180-\varphi_c)$ degrees. At the short operating period the attenuation (A_2) produced by the plant in the subsidiary loop at period T′ will

* For an explanation see Ref. (54).

normally be large, compared with the attenuation (A_P) produced by the whole of the plant in the main system at its lower operating period (T); therefore the proportional action factor (K_1') of C will generally be large compared with the factor (K_1) of C.

Care must be taken to ensure that the overall gain of C and C′ is not so large that anticipated deviations will over-range the correcting unit.* This is particularly important when the subsidiary loop forms a flow control system.

Controller Adjustments—The settings of the main and subsidiary controllers can be determined as follows:

(1) The subsidiary controller is first adjusted to give $\mu_{max.}$ and e:1 damping using the phase/gain diagram of the controller and the phase-lag/attenuation diagram of the part of the plant in the subsidiary loop.

Normally, however, the subsidiary controller has proportional action only, and hence it is only necessary to find the plant attenuation (A_2) at the period (T′) which gives 180° phase lag (for the required subsidence ratio). This can be found directly from the frequency response diagram.†

(2) The frequency response of the plant with the subsidiary control loop is then found. If this is done experimentally, by the method of Chapter 7, the loop is opened at D. The output of the analyser is connected to the subsidiary controller (so as to vary the desired valve setting sinusoidally) and the main controller is used (with proportional action only) as a transmitter, to provide a pressure signal to the analyser proportional to the signal from the main detecting element. From the frequency response data obtained the phase-lag/attenuation diagram can be plotted for the plant plus subsidiary control loop. The main controller is then adjusted to give $\mu_{max.}$ and e:1 damping, using this diagram and the controller phase/gain diagram.

The main controller often has proportional + integral action and hence again a phase/gain diagram need not be used. The settings will be $S = T$, where T = period at which the phase lag of the plant plus subsidiary control system is 171°, and the proportional action factor K_1 can be found at once from the attenuation (A) at the operating period (T).

If the settings determined in this way are such that either (a) the overall gain of the two controllers is too large, or (b) the ratio $\frac{T}{T'}$ is likely to lead to resonance, then K_1' and T′ must be adjusted to more

* The total gain of C and C′ is K_1'.(I.F.K_1)—(see Section 11.6).

† In practice, when the frequency response diagrams are not available the controllers can be set up by the usual trial and error methods (see Appendix IV).

suitable values in relation to K_1 and T.* This will necessitate departing from e:1 damping in the subsidiary loop; this is not important when the subsidence ratio in the subsidiary loop has a negligible effect on the form of recovery given by the complete system. The main loop will be adjusted to give a subsidence ratio e:1 as before.

The improvement which is effected by using a cascade system instead of a single loop system depends on the characteristics of the plant and the point of entry of the disturbance, as illustrated by the following example.

18.3.1 *Comparison of Cascade and Single Loop Performances*

The plant to be controlled is represented by the equivalent system shown in Fig. 18.3. Disturbances (θ_1 and θ_2) enter the plant at A (through or near the correcting element) and at B (after two transfer stages). The performances of a single loop system and a cascade system in controlling the given plant will be compared in terms of the D.R.F. and the period of operation which each system gives.

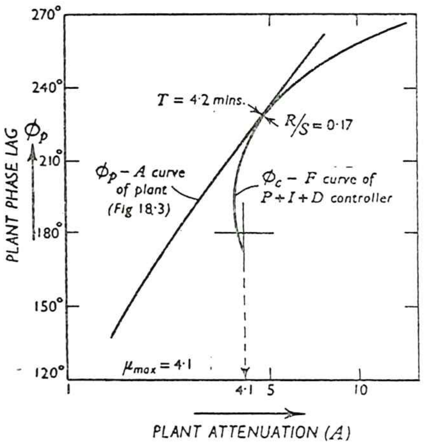

FIG. 18.4. Determination of controller action settings for single loop control of plant of Fig. 18.3.

The disturbances θ_1 and θ_2 are assumed to be of step form and to give equal potential deviation of 1,000 units. In this example the controller in the single loop system and the main controller in the cascade system both have proportional + integral + derivative action. The subsidiary controller in the cascade system has proportional action.

Determination of Controller Adjustments—In order to clarify the method described above for determining the controller action settings, the application of the method to this example is indicated very briefly. The method of finding the frequency response of an open loop system experimentally has been described in detail in Chapter 7 and the determination of optimal controller settings

* Alternatively T′ can perhaps be reduced by repositioning the detecting element D.E.2, so that fewer transfer stages are included in the subsidiary loop, or by reducing the measuring and transmission lag in the loop.

from the phase-lag/attenuation diagram of a plant and the phase/gain diagram of the controller have been given in detail in Chapter 10.

Single Loop System—The phase/gain diagram of the controller is shown in Fig. 18.4 in the tangential position relative to the phase-lag/attenuation diagram of the plant contained within the single loop system. The value of $\frac{R}{T}\left(=\frac{R}{S}\right)$ is 0·17 and the period of

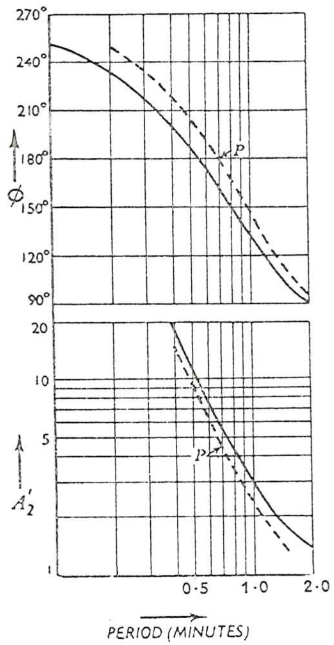

FIG. 18.5.1. Determination of μ for subsidiary system of Fig. 18.3.

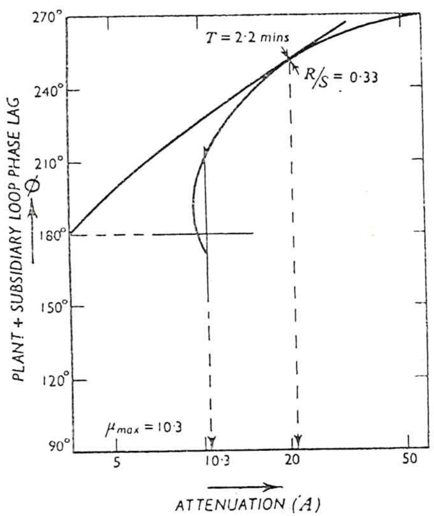

FIG. 18.5.2. Determination of main controller action settings for cascade system of Fig. 18.3.

operation is 4·2 minutes, so that the derivative action time (R) must be set at 0·71 minutes. K_1 and the proportional band width can be found from the value of $\mu_{max.}$ (= 4·1) as described in Section 10.3.

Cascade System—The proportional band width of the *subsidiary* (proportional) controller is found from the value of μ given by the frequency response diagram in Fig. 18.5.1. This diagram gives the frequency response of the part of the plant in the subsidiary loop + the transfer stage (τ_m = 0·1 min.) corresponding to the detecting element (D.E.2) shown in Fig. 18.3. The attenuation (A_2')* at the period of operation (0·7 minute) which gives a phase lag of 180° is 4·7, for e:1 damped oscillations, so that $\mu = 4·7$.

The action settings for the *main* controller (P + I + D) are found from the phase-lag/attenuation diagram of the part of the system between D and E in Fig. 18.3, and the phase/gain diagram of the controller (C). The point of tangency (Fig. 18.5.2) gives the value of $\frac{R}{T}\left(=\frac{R}{S}\right)$ as 0·33, and the period of operation as 2·2 minutes. Therefore the derivative action time must be set at 0·73 min. The proportional band width and K_1 can be found from the value of $\mu_{max.}$ (= 10·3).

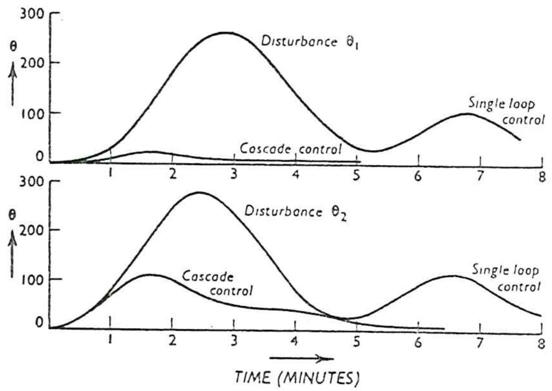

FIG. 18.6. Comparison of recoveries of single loop and cascade systems.

Value of $\frac{T}{T'}$ *and Gain*—$\frac{T}{T'} = \frac{2·2}{0·7} > 3$, so that resonance effects will not be appreciable.

The total gain (G_T) of the two controllers is given by the products of the attenuation (α) due to the subsidiary and main loops at their respective periods of operation (T′ and T); i.e. $G_T = \left(\frac{A_p}{K} \times \frac{A_2}{K'}\right)$.

The values of A_p and A_2 given by Figs. 18.5.1 and 2 are 21 and 4·7. The value of G_T is thus $\frac{98·7}{K.K'}$. This value was found to give no risk of over-ranging the control valve in the given operating conditions.

* $A_2' = A_2 \times$ (attenuation due to exponential stage with time constant τ_m at period of operation (T′) of the subsidiary loop)

$$= A_2 \times \left[1 + \left(\frac{2\pi\tau_m}{T'}\right)^2\right]^{\frac{1}{2}} \quad \text{(See Section 6.2.2.)}$$

Therefore the controller settings given require no modification to avoid resonance or over-ranging of the correcting unit.

Comparison of Performance—The recovery curves for the single loop and cascade systems are given in Figs. 18.6 for θ_1 entering at A and θ_2 entering at B respectively.

The values of the (true) D.R.F. and period are summarised in Table 18.1.

TABLE 18.1

System	D.R.F.		Period (minutes)
	for θ_1 at A	for θ_2 at B	
Single loop	3·7	3·6	4·2
Cascade	50	8·8	2·2 (main loop)

D.R.F.—It will be seen that for a disturbance entering at A (e.g. for a supply disturbance) the ratio of the D.R.F.'s given by the cascade and single loop systems is about 12, but for a disturbance entering at B (outside the subsidiary loop) the ratio of the D.R.F.'s is only about 2·5. Thus, the cascade system is much more effective in reducing peak disturbances in supply than the single loop system. When the disturbance enters later in the plant, outside the subsidiary loop (e.g. for a disturbance in a subsidiary supply condition or for a demand disturbance), the cascade system does not produce as large an increase in D.R.F., although in some plants the increase may be considerable (e.g. 2·5 in the example taken) due to decrease of the period of operation, as explained below.

Period—It is important to note that the cascade system decreases the period of operation of this plant from 4·2 to 2·2. This is a considerable advantage when the process demands a fast return of controlled condition to desired value after a disturbance. It also explains why the cascade system effects an increase in D.R.F. for a disturbance which falls outside the subsidiary loop, i.e. by increasing A_1 (Fig. 18.3) as a result of the shorter operating period.

18.3.2 *Conditions for use of Cascade System*

The following notes on the properties of the cascade system may assist designers in deciding in what conditions it is profitable to use it.

(1) *Plant exhibiting distance/velocity lag.* It is clear that if a plant exhibits mainly distance/velocity lag and the disturbance enters after the detecting element of the subsidiary loop of a cascade system, cascade control will improve control quality by decreasing the period of operation, but there will be little increase in D.R.F. If, however, the disturbance enters within the subsidiary loop the increase in D.R.F., as well as the decrease in period, will be considerable. (See description of Fig. 18.7 below.)

(2) *Change of load.* The subsidiary controller of a cascade system ensures that the correcting condition is maintained very close to the value called for by the main controller, in spite of changes in plant characteristics with change of load or of supply pressure. It will be remembered, from Chapter 16, that in a single loop system such changes may demand a readjustment of proportional band width in order to obtain the same stability of control, i.e. the same subsidence ratio. In a cascade system, variation in the subsidence ratio of the subsidiary control system generally has only a secondary effect on the stability of the whole system; therefore it is not necessary to characterise the correcting element in order to maintain $e:1$ damping in spite of load changes.

It should be noted however that when the controlled condition in the subsidiary loop is flow, measured by an orifice plate, the square root scale of the subsidiary controller will cause the overall gain of the controllers to increase very considerably at flow rates say 50% below the normal value for which the system is set up. Consequently instability may result. Conversely at flow rates higher than normal the gain will be too low. This difficulty can be overcome by using a flow controller designed to give a linear scale, or by adding a linear to square root conversion device to the unit which receives the signal from the main controller and adjusts the desired value of the subsidiary controller.

An example of the effect of the square root relationship in a cascade system is given by Hoyt and Stanton (71).

(3) *Long operating period.* The cascade system is particularly useful for decreasing the effect of disturbances entering early in the plant, when the period of operation of the system is long under single loop control. Cascade systems are frequently used therefore when the final controlled condition is the measured composition of the product, because most analytical instruments exhibit a large time lag and so tend to give a long operating period.

(4) *Occasional large or rapid disturbances.* An incidental advantage of the cascade system is that the subsidiary controller can be arranged to limit the maximum corrective signal which can be applied to the control valve, without decreasing the potential correction available to the main controller as long as the signal to the correcting unit

does not exceed the permissible maximum. Hence the overall gain of the two controllers can be high and the main controller can safely be adjusted to give a long derivative action time, without incurring the risk of over-ranging the control valve when unusually rapid disturbances or unusually large transitory disturbances occur.

18.3.3 *Example*

A good example of the use of a cascade system for reducing the effect of disturbances which enter near the correcting element is provided by the steam temperature control system as applied to boiler superheaters, shown in Fig. 18.7. This system, which is discussed in detail by Toop (5), employs a desuperheater between the primary and secondary sections of the superheater to correct steam temperature disturbances. The steam temperature is corrected by spraying water

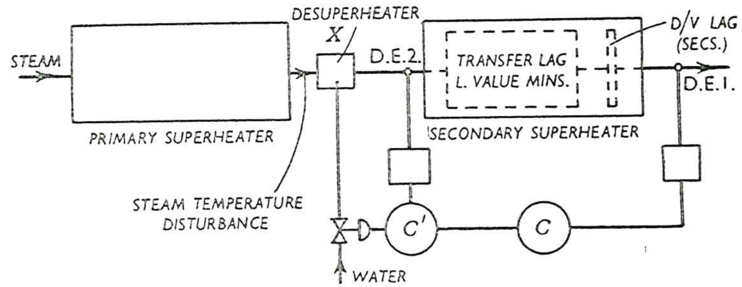

FIG. 18.7. Example of cascade control system.

into the steam in the desuperheater before it enters the secondary section.

Major disturbances are due to changes in gas flow or temperature, and to changes in the steam rate. Since the two sections of the superheater are in the same gas stream, disturbances both in steam rate and in gas conditions will be reflected in steam temperature disturbances at the exit from the primary section. These will clearly be corrected much more rapidly by the subsidiary control loop than by the main loop, which contains the large transfer stage formed by the mass of the superheater tubes and the fluid-film resistances to heat transfer (see Section 15.10). The only appreciable transfer stage in the subsidiary loop is associated with the temperature detecting installation. Therefore the period of the subsidiary loop will be short and the proportional band of the subsidiary controller C′ will be narrow. Disturbances entering the loop will therefore be very quickly corrected.

It should be noted that the distance/velocity lag, due to the time taken for the steam to pass through the secondary superheater tubes, occurs after the subsidiary loop. Normally this lag is small in superheaters (of the order of seconds) and therefore has a negligible effect compared with the transfer stage (whose L value is of the order of minutes). If, however, in a similar type of installation the distance/velocity lag was large, then the system would illustrate Note (i) of Section 18.3.2 concerning distance/velocity lag. The cascade system would prove increasingly more advantageous as the distance/velocity lag increased for disturbances entering at X. For disturbances entering after D.E.2, a cascade system would only improve control in so far as it would decrease the period of operation of the system as a whole, as previously explained.

For other examples taken from practice a paper by Ziegler (60) should be consulted.

Parallell control

18.4 System of One Measuring Unit and Two Correcting Units

A system which proves valuable in practice when disturbances enter the plant in a subsidiary supply condition or within the plant itself is shown in Fig. 18.8. Very rapid corrective action is required in these conditions in some cases, to prevent damage to the plant or product quality. For example, the part Z of the plant in Fig. 18.8 may be a converter in which a violently exothermic reaction takes place. If a hot spot develops, the process may commence to 'run away' and very rapid corrective action is required. Similarly, if a liquid/vapour phase reaction takes place in the converter and a sudden increase occurs in the liquid supply rate, again the process may commence to run away and immediate corrective action is essential.

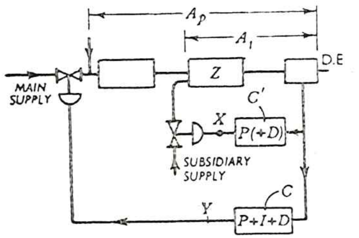

Fig. 18.8. Control system with two controllers operating from one detecting element.

The signal from the detecting element which measures the value of the controlled condition is transmitted to the main controller, which operates a control valve in the main supply line, and also to a subsidiary controller, which operates another control valve in a second supply line. If the attenuation (A_1) due to the part Z of the plant is small compared with the attenuation (A_p) due to the whole plant at the operating period, the period of the subsidiary control loop will be short compared with that of the main loop; the correction applied by the subsidiary controller will become effective much earlier than that applied by the main controller.

The function of the subsidiary controller is to correct for the

disturbances entering late in the loop, while the main controller is making the necessary adjustments to the main supply condition. When this correction has been made the subsidiary controller should return the subsidiary supply to its normal value. Hence the subsidiary controller should have proportional or proportional + derivative action. It should not have integral action, which would give a permanent change in subsidiary supply. The main controller can have proportional + integral + derivative action, as called for by the process.

Consideration of the operation of the system in frequency response terms will assist in determining the controller adjustments. The subsidiary system must be adjusted first to give a sufficiently large D.R.F. to reduce deviations to within the prescribed limits. The proportional band width (and derivative action time) required can be found by carrying out a frequency response analysis of the subsidiary loop opened at X. Care must be taken to ensure that the gain of the subsidiary controller is not so high that it causes fluctuations in the subsidiary supply in excess of those which can be tolerated from an operational point of view.

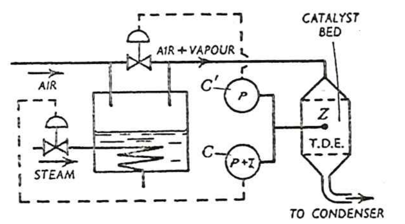

Fig. 18.9. Example of system with two control loops operating from one detecting element.

Parallell control (same cvz. dynea)

The action settings of the main controller are then determined from the frequency response of the whole system, determined by opening the main loop at Y. To avoid resonance the period (T) of the whole system must be such that T is larger than 3T'—where T' is the period of the subsidiary system. It should be noted that if it is important to return the subsidiary supply to its normal value as soon as possible, then the operating period and subsidence ratio of the main system should be adjusted to give the shortest time of return to desired value, if necessary at the expense of departing from the $\mu_{max.}$ criterion.

It will be noticed that the subsidiary loop assists in decreasing the period of operation of the whole system because it introduces phase advance in the same way as the subsidiary loop of the cascade system.

An example of the use of this type of system is shown in Fig. 18.9. A liquid is vaporised and mixed with an air stream. The mixture reacts in the presence of a catalyst in the converter. The controlled condition is converter temperature and normally the steam rate is the correcting condition. If the converter temperature commences to increase rapidly, the subsidiary loop will reduce the vapour concentration long before the main controller action becomes effective.

Similar to cascade (more like state feedback)

18.5 System With Two Independent Controllers and One Correcting Unit

In this system, shown in Fig. 18.10, the object of the subsidiary loop is to counteract the effect of disturbances arising near the correcting element. It will be seen that the system is very similar to a cascade system in both purpose and method of operation. The only difference is that, as in the system described in the previous section, the signal from the main controller does not influence the signal from the subsidiary controller. In this system the output of each is, however, combined in an adding unit to operate the single correcting unit.

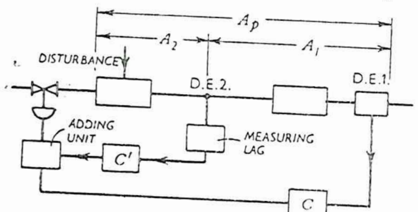

Fig. 18.10. Control system with two independent controllers and one correcting unit.

Applications of this type of system are most numerous in boiler control installations (5, 63).

Controller Adjustment—The action settings of the two controllers can be determined by the method employed for the cascade system. The subsidiary controller (C') is first adjusted, using the frequency response diagram of the subsidiary loop. The period of the subsidiary loop should normally be short and the proportional band will be sufficiently narrow to make integral or derivative action unnecessary in the subsidiary controller.

The frequency response of the plant and subsidiary loop will give the phase-lag/attenuation diagram required for determining the action settings of the main controller (C).

Note: This system differs from the cascade system in that the two controllers are arranged in parallel with the result that:

(a) the subsidiary controller cannot be used to prevent over-ranging of the main control valve by the main controller, and
(b) the main control valve must be characterised to compensate for changes in plant characteristics due to load changes, or for supply pressure changes, as in the single loop system.

18.6 Ratio Control System

In many processes the values of two operating conditions are maintained in a constant ratio by the system of Fig. 2.5 (b), which is a flow ratio control system. A signal proportional to flow (1) is transmitted by the proportional action unit F.C.1 to the ratio controller F.C.2, whose desired value is continuously adjusted to a flow of the required value compared with flow (1).

See last page for Fig. 2.5

Flows (1) and (2) may be, for example, the flows of two fluids supplied to react in a plant or the flows of waste acid and alkali in an effluent neutralisation system. Numerous examples have been published of the use of flow ratio systems in the control of distillation and absorption columns (64), (65), (66). One such use is illustrated in Fig. 18.11, which shows how some of the control systems discussed in this chapter could be used, in place of the simple single loop systems shown in Fig. 2.1, to control a distillation column. The reflux rate in this system is controlled to be in a constant ratio to the feed rate.* A paper by Wallis (67) is of particular interest in that it describes a coordinated control system which uses ratio control and employs electric control equipment.

A detailed discussion of the use of ratio control in combustion control systems is given by Farquhar (73).

Adjustment of Ratio Controller—The adjustment of the controllers F.C.1 and F.C.2 depends on the changes which are expected to occur in flow (1). It will be seen that F.C.2 must be set up so that two conditions are fulfilled:

(a) F.C.2 controls flow (2) with the required precision at a steady desired value, when flow (1) is constant.
(b) Changes in desired value of F.C. 2 corresponding to changes in flow (1) do not result in excessive overshoot or resonance.

To satisfy condition (a) F.C.2 can be set up in the normal manner with a wide proportional band and short integral action time (of the order of the operating period).

To satisfy (b) resonance must be avoided by taking care that the period of the closed loop system differs by a factor of at least 3 from

* It has been common practice to control the reflux rate at a constant value, which is large enough to permit the efficiency of separation specified at the maximum feed rate expected. This ensures that the column will produce greater purity of products at lower feed rates, at the expense of reheating more distillate than is necessary to obtain the specified purity. Saving can therefore be effected in heat supplied to the boiler by maintaining the reflux rate in constant ratio with the feed rate. An argument against doing this is that the major part of the heat supplied to the boiler is used to reheat the reflux, and therefore changes in reflux rate must be accompanied by changes in heat supply to boiler. It is therefore easier to obtain steady operation of the column by maintaining reflux rate constant and it may pay to do this at the expense of supplying more heat than is necessary. Changes in heat input will then be made only to compensate for unavoidable changes in conditions, e.g. feed composition and rate of heat loss to the ambient atmosphere.

It is assumed that feed rate changes will be slow when this system is used, since two factors would otherwise militate against its success, namely (a) frequent and rapid changes in heat supply to boiler and (b) the delay in re-establishing equilibrium between vapour rate up the column and liquid flow down the column. This would call for a time delay in the transmission line from feed rate transmitter to ratio controller, to arrange for the reflux rate to change when the change in vapour rate up the column reached the top plate.

that of any cyclic changes in flow (1), and that the subsidence ratio of the closed loop system is sufficiently large to avoid excessive overshoot resulting from the most rapid changes expected to occur in flow (1).

As a guide, it can be taken that, to a sufficiently good approximation for most purposes, $e:1$ damping will result in 60% overshoot after a step change in desired value; 10:1 damping will give about 20% overshoot. (See Appendix V.)

Full treatments of the response of closed loop systems to changes in desired value are given in the literature (9), (68), (69).*

18.7 Averaging Control

It is frequently necessary to 'smooth out' disturbances in supply rates. For example, the distillation column of Fig. 2.1 is intended to produce pure bottom product. The top product will probably become the feed to a second column for further purification. If the level controller on the distillate accumulator maintains the level constant, the flow of distillate to the next still will only be constant if the feed rate, composition and all other conditions in the first still are constant. To smooth out the inevitable fluctuations in flow, an averaging level control system can be used, as shown in Fig. 18.11. (The bottom product goes to a storage vessel, and therefore fluctuations in its rate of flow are unimportant. The level in the boiler is therefore held constant.)

The averaging system consists essentially of a wide band proportional or proportional + integral controller, which calls for a comparatively small change in flow rate when the level changes in the accumulator (or in general the 'surge' or 'buffer' vessel).

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If considerable changes are expected in the pressure upstream of the control valve it is usually worth while to arrange for the level controller to adjust the desired value setting of a flow controller, which will hold the flow at the desired value demanded.

The inherent regulation† of a surge vessel and outlet valve tends to smooth out fluctuations due to changes in inflow and by making the vessel sufficiently large these fluctuations can be made as slow as required, without using a controller. The purpose of using the averaging level controller is to reduce the size of the vessel required to permit a specified rate of change of outflow to be achieved. This design problem has been treated in detail by Mason and Philbrick (70) who discuss specific examples.

* The majority of single loop process control systems are designed to maintain the controlled condition at a *constant* desired value: in the servo-mechanism field it is more frequently necessary to design systems which will maintain the controlled condition as close as possible to a constantly varying desired value.

† Defined in Section 5.2.2.

The dimensions of the surge vessel and the controller settings are determined by the magnitude, form and frequency of occurrence of the disturbances in inflow, and by the fluctuations which are permissible in outflow. The pattern of the disturbances to be expected must therefore be carefully considered. In general it may contain every type of disturbance superposed, but in a specific plant one type often predominates. For example, the main change in flow of distillate to the accumulator in Fig. 18.11, when the column is running steadily,

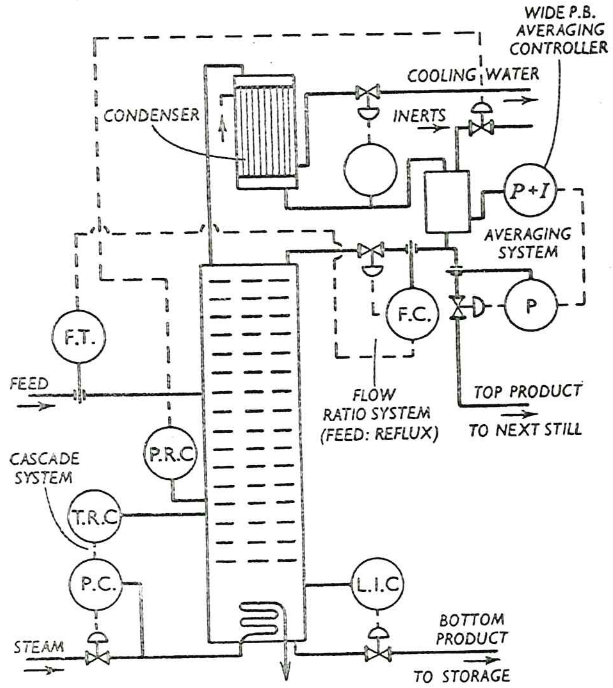

Fig. 18.11. Examples of the use of complex systems.

could be expected to be a slow increase or decrease with a superimposed small cyclic change due to the operation of the control system of the column.

Each case must be considered individually, but there are several general points to consider in all applications:

(1) In general it is an advantage to use integral action in the averaging level controller, so that the level in the surge vessel is slowly brought back to a definite position.

(2) The system must be adjusted so that the surge vessel does not become completely full or empty. A high and low level limit

device is often fitted in the controller, so that the valve is opened or shut to prevent further rise or fall of level when either limit is reached.

(3) When pressure changes occur in the system it is advisable to install a subsidiary flow controller, to ensure that the flow rate is as demanded by the averaging controller.

18.7.1 *Dimensions of Vessel and Controller Adjustments*

Since this is the only example given in the text of the quantitative design of a complete control system (i.e. plant and control equipment), it is worth while to indicate the basis of the design method. The plant part of the loop is the simplest conceivable; hence it serves well as an illustrative example because its characteristics can be calculated at once from its dimensions. Also, the design of averaging systems is important in practice and permits economic design of surge vessels, which may be very expensive, e.g. in high pressure systems or when special materials of construction are necessary.

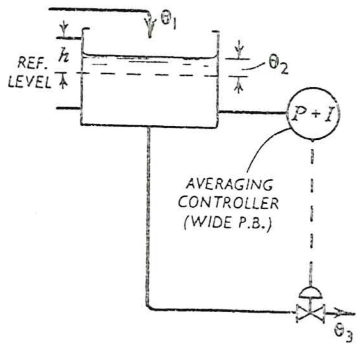

FIG. 18.12. Simple averaging control system.

The problem consists of finding the proportional action factor (K_1) and the integral action time (S) of the averaging controller, which will give the specified 'smoothness' of outflow, in terms of the dimensions of the vessel, i.e. its capacity (C) per unit depth and the total depth ($2h$) over which the level can be permitted to change.

Consider the simple system of Fig. 18.12 which employs a (P + I) averaging controller.

Let the inflow and outflow rates and the depth of liquid be θ_1, θ_3 and θ_2 respectively at time t; θ_1, θ_2 and θ_3 are measured from initial equilibrium values when the level is at the desired value (i.e. at $t = 0$, $\theta_2 = 0$, $\theta_1 = \theta_3$).

The rate of change of depth $\left(\dfrac{d\theta_2}{dt}\right)$ at time t is given by:

$$\frac{d\theta_2}{dt} = \frac{1}{C}(\theta_1 - \theta_3) \quad . \quad . \quad . \quad . \quad (18.1)$$

The output signal (V) from the controller at time t is given by:

$$V = K_1\left[\theta_2 + \frac{1}{S}\int \theta_2 . dt\right]$$

and the outflow (θ_3) from the vessel is thus

$$\theta_3 = K.V = \mu\left[\theta_2 + \frac{1}{S}\int \theta_2 . dt\right] \quad . \quad . \quad (18.2)$$

(where K is flow rate change due to unit change in V*).

By differentiating equations (18.1) and (18.2) we obtain:

$$C\,\frac{d^2\theta_2}{dt^2} = \frac{d\theta_1}{dt} - \frac{d\theta_3}{dt} \quad . \quad . \quad . \quad . \quad (18.3)$$

and

$$\frac{d\theta_3}{dt} = \mu\left(\frac{d\theta_2}{dt} + \frac{1}{S}.\theta_2\right) \quad . \quad . \quad . \quad (18.4)$$

Eliminating $\dfrac{d\theta_3}{dt}$ from equations (18.3) and (18.4), we have:

$$\frac{SC}{\mu}.\frac{d^2\theta_2}{dt^2} + S\frac{d\theta_2}{dt} + \theta_2 = \frac{S}{\mu}.\frac{d\theta_1}{dt} \quad . \quad . \quad (18.5)$$

which relates θ_2 and θ_1.

Differentiate equation (18.4), to obtain:

$$\frac{d^2\theta_3}{dt^2} = \mu\left(\frac{d^2\theta_2}{dt^2} + \frac{1}{S}.\frac{d\theta_2}{dt}\right) \quad . \quad . \quad . \quad (18.6)$$

Substituting for $\dfrac{d^2\theta_2}{dt^2}$ and $\dfrac{d\theta_2}{dt}$ in equation (18.6) from equations (18.1) and (18.3) we have:

$$\frac{SC}{\mu}.\frac{d^2\theta_3}{dt^2} + S.\frac{d\theta_3}{dt} + \theta_3 = S.\frac{d\theta_1}{dt} + \theta_1 \quad . \quad . \quad (18.7)$$

Integration of equation (18.7) gives an expression which relates θ_1 and θ_3. To find the variations (θ_3) which will occur in the outflow due to changes in inflow, in terms of μ, S and C, equation (18.7) must be solved for the given conditions. Two cases are of primary importance, namely:

(i) step changes in inflow, and
(ii) a sustained sinusoidal change given by

$$\theta_1 = \theta_1^\circ . \sin\frac{2\pi}{T}.t = \theta_1^\circ . \sin \omega t$$

In the case of step changes the important considerations are (a) the form of recovery after an individual step change and (b) the frequency of occurrence of the changes. It will be sufficient, for the present purpose, to discuss the recovery (i.e. the transient response) after a single step change.

* K is assumed constant in spite of changes in pressure upstream of the valve due to changes in level. The valve is shown in Fig. 18.12 a sufficient distance below the vessel to make negligible the effect of level changes on pressure upstream of the valve.

In the case of a cyclic change in inflow the steady state conditions are the important consideration and particular attention must be given to the conditions which lead to resonance and consequent magnification of the inflow disturbance in the outflow.

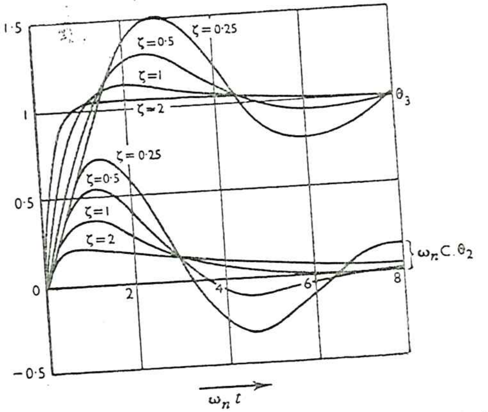

FIG. 18.13. Change in outflow (θ_3) and of level (θ_2) following step change in inflow ($\theta_1 = 1$) in averaging level control system.

The solution of equations (18.5) and (18.7) is given in the standard mathematical texts and is discussed in detail in books on process control by E. S. Smith (72) and by Farrington (9).* If these works are consulted, it will be found that the characteristics of the system represented by these equations can be presented most conveniently in terms of the natural frequency (ω_n) and the damping ratio (ζ) of the system.

In terms of the parameters of the system of Fig. 18.12.

$$\omega_n = \sqrt{\frac{\mu}{S.C}} \text{ and } \zeta = \tfrac{1}{2}\, S.\omega_n \quad . \quad . \quad . \quad (18.8)$$

The operating period (T) of the system is given by:

$$T = 2\pi.(\omega_n \cdot \sqrt{1 - \zeta^2})^{-1}$$

Step Disturbance—In Fig. 18.13 values of θ_3 and $\omega_n.C.\theta_2$ following a step disturbance ($\theta_1 = 1$) in inflow are plotted against ($\omega_n.t$) for a range of values of ζ.

From these curves values of ζ and ω_n can be chosen so that:

(a) peak disturbances in outflow will not be in excess of a specified maximum for the step changes expected in inflow.

(b) the oscillations die away before another disturbance occurs, or sufficiently rapidly to meet the needs of the process.

The selected values of ζ and ω_n determine S and $\frac{\mu}{C}$ (in expressions (18.8) above).

The next step is to consider the oscillations of the level in the same way and to determine the value of $\frac{\mu}{C}$ which will permit the use of

* See Appendix V.

the smallest surge vessel, in terms of C and $2h$. Values of $\omega_n C\theta_2$ are plotted against $\omega_n t$ in the lower set of curves of Fig. 18.13 for $\zeta = 0{\cdot}25, 0{\cdot}5, 1$ and 2.

The proportional band width of the controller, to operate with this vessel, is then found from the values of $\frac{\mu}{C}$ and C.

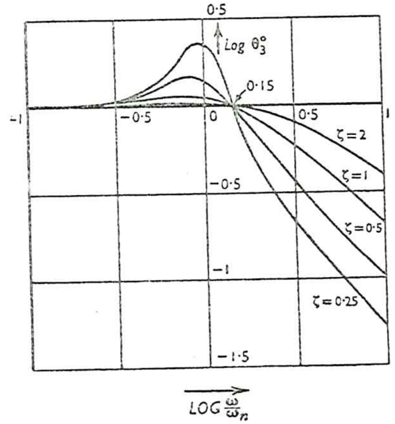

FIG. 18.14.1. Amplitude (θ_3°) of oscillation in outflow corresponding to unit amplitude sinusoidal change in inflow, in averaging level control system.

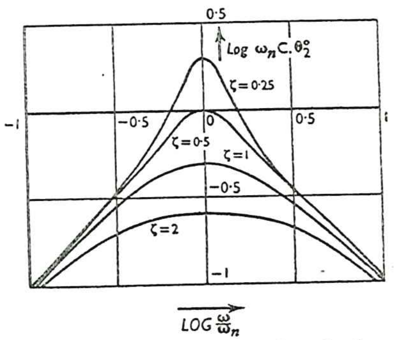

FIG. 18.14.2. Curves for the selection of the damping ratio to keep level variations within prescribed limits, for sinusoidal inflow variation, in averaging level control system.

Sinusoidal Disturbance—In Fig. 18.14.1 log θ_3° is plotted against log $\left(\frac{\omega}{\omega_n}\right)$; where θ_3° = amplitude of sinusoidal variation in outflow corresponding to an inflow disturbance of unit amplitude and frequency ω.

It will be seen that when log $\frac{\omega}{\omega_n} < 0{\cdot}15$, $\theta_3^\circ > 1$, i.e. the disturbance is amplified. The greatest amplification occurs when log $\left(\frac{\omega}{\omega_n}\right) \doteq 0$ or $\frac{\omega}{\omega_n} = 1$, and ζ is smallest.

If only one cyclic disturbance is expected, ζ can be selected to give the required attenuation of inflow disturbances. The lower the value of ζ is, the smaller the ratio $\frac{\omega}{\omega_n}$ can be made for the same attenuation. If there are disturbances of a number of frequencies, care must be taken to ensure that none of them make $\frac{\omega}{\omega_n}$ less than $0{\cdot}15$, so that they are not amplified. If this is not possible, then a high value of ζ must be used, in order to minimise the amplification. Fig. 18.14.2

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BIBLIOGRAPHY

1. Baumann, K., 'Heat Engines', *Proc. I. Mech. Eng.*, **160**, 1, 1949, 61.
2. Ritchie, E. G. and Kirov, N. Y., 'Performance of a Lancashire Boiler,' *Proc. I. Mech. E.*, **162**, 1, 1950, 27.
3. *The efficient use of Fuel*, H.M.S.O., London, 1944, p. 84.
4. Tustin, *Automatic and Manual Control*, Butterworths, London, 1952, p. 202
5. Toop, G. H., 'The Application of Open Loop Controls', *Instrument Engineer*, (Geo. Kent Ltd., Luton, England), **1**, 3, April 1953, 49.
6. Ivanoff, A, 'Theoretical Foundations of the Automatic Regulation of Temperature', *J. Inst. Fuel*, **7**, 1934, 117.
7. Westcott, J. H., 'The Frequency Response Method: its relationship to transient behaviour in control system design', *Trans. S.I.T.*, **4**, 3, Sept. 1952, 113.
8. Takahashi, Y., 'Transfer Function Analysis of Heat Exchange Processes', Ref. 4, 235.
9. Farrington, G. H., *Fundamentals of Automatic Control*, Chapman & Hall, London, 1951.
10. Farrington, G. H., 'Theoretical Foundations of Process Control', *J.I.E.E.*, **94**, pt. IIA, 1, 1947, 23.
11. Farrington, G. H., 'Automatic Temperature Control of Jacketed Pans', *Trans. S.I.T.*, **1**, 1, Jan. 1947, 2.
12. Hartley, H., 'The place of chemical engineering in modern industry', *Chemical and Process Engineering*, **34**, 8, Aug. 1953, 230.
13. Rutherford, C. I., 'Practical application of frequency response analysis to automatic process control', *Proc. I. Mech. E.*, **162**, 3, 1950, 334.
14. Ziegler, J. G. and Nichols, N. B., 'Optimum Settings for Automatic Controllers', *Trans. A.S.M.E.*, **64**, Nov. 1942, 759.
15. Callender, A., Hartree, D. R. and Porter, A., 'Time Lag in a Control System', *Phil. Trans. Roy. Soc.*, **235**, (A) 1936, 415.
16. Aikman, A. R. and Halsall, J. R., 'A portable instrument for process control analysis', *Inst. Practice*, **5**, 7, 1951, 393, and **5**, 12, 1951, 793.
17. St. Clair, D. W., Erath, L. W., Gillespie, S. L., 'Sine-Wave Generators', *Trans. A.S.M.E.*, **76**, 8, Nov. 1954, 1177.
18. Williams, F. C. and Uttley, A. M., 'The Velodyne', *J.I.E.E.*, **93**, Pt. IIA, 1, 1946, 1256.
19. Carlisle, S. S., Contribution to discussion of paper of Ref. 7 above.
20. Ream, N. S. 'The Calculation of Process Control Settings from Frequency Characteristics', *Trans. S.I.T.*, **6**, 1, March 1954, 19.
21. Young, A.J., *Process Control*, The Instrument Publishing Co. Inc., 1954, p. 21.
22. Bane, W. T., 'Design Charts for an On-Off Control System', *Trans. S.I.T.*, **5**, 2, June 1952, 52.

    Takahashi, Y., 'Interference of Two-Position Controllers', I.S.A. Conference, Sept. 1954, Paper 54-41-1.

    Sartorius, H. (p. 421), Oldenbourg, R. C. (p. 435), Oppelt, W. (p. 449), Ref. 4.
23. North, J. D., 'The Human Transfer Function in Servo-systems', Ref. 4, p. 473.

    Bartlett, F., 'Human Control Systems', *Trans. S.I.T.*, **3**, 3, Sept. 1951, 134.
24. Gilliland, R., 'Electronic Control Systems', *Instruments and Automation*, **27**, 12, Dec. 1954, 1962.
25. Janssen, J. M. L., 'Analysis of Pneumatic Controllers', Ref. 4, p. 189.
26. Gould, L. A. and Smith, P. E., 'Dynamic Behaviour of Pneumatically Operated Control Equipment', (Part II), *Instruments*, **26**, 7, 1953, 1026.
27. Ziegler, J. G. and Nichols, N. B., 'Optimum Settings for Automatic Controllers', *Trans. A.S.M.E.*, **64**, 1942, 759.

27. (*continued*)
Hazebroek, P. and van der Waerden, P. L., 'Theoretical Considerations and the Optimum Adjustment of Regulators', *Trans. A.S.M.E.*, **72**, April 1950, 309–322.
Ford, R. L., 'Determination of Optimum Process Controller Settings and the Confirmation by Means of an Electronic Simulator', *Proc. I.E.E.*, **101**, 2, 80, April 1954, 141–155.
Oldenbourg, R. C. and Sartorius, H., 'A Uniform Approach to the Optimum Adjustment of Control Loops', *Trans. A.S.M.E.*, **76**, 8, Nov. 1954, 1265.
Wolfe, W. A., 'Controller Settings for Optimum Control', *Trans. A.S.M.E.*, **73**, May, 1951, 413.
Jones, S., 'The Determination of the Best Form of Response for a Servo when an Extraneous Random Disturbance is Present with the Error Signal', Ref. 4, p. 139.
Sartorius, H., 'Das Optimierungsproblem in der Regelungstechnik', *Regelungstechnik*, **1**, 1953, 74.
Altenheim, H. J., 'Optimale Regelung einer Regelstrecke mit Ausgleichsgrad, Anlaufzeit und Totzeit', *Regelungstechnik*, **1**, 1953, 232.
Janssen, J. M. L., 'Control-System Behaviour Expressed as a Deviation Ratio', *Trans. A.S.M.E.*, **76**, 8, Nov. 1954, 1303.
28. Bergen, S. A., 'Pneumatic Force Balance Instrumentation', *Inst. Practice*, **8**, 4, 1954, 312.
29. Kirk, D. B., 'Nozzle Flow Characteristics in Pneumatic Force-balance Circuits', *Trans. A.S.M.E.*, **70**, 2, 1948, 111.
30. Moore, C. B., 'The Solution of Instrumentation Problems by the Pneumatic Null-Balance Method', *Instruments*, Sept. 1945.
31. Clarridge, R. E., 'New Transet Tri-Act Controller', *Taylor Technology*, **3**, 3, 1950, 8.
32. Aikman, A. R. and Rutherford, C. I., 'The Characteristics of Air-operated Controllers', Ref. 4, p. 175.
33. Williamson, H., 'Theory and Design of Compound-action Pneumatic Controllers' *Trans. S.I.T.*, **6**, 4, Dec. 1954, 162.
34. Boundy, J. R. and Bergen, S. A., 'An Electronic Process Controller', *Proc. I.E.E.*, **98**, Pt. II, 1951, 609.
35. Boundy, J. R., 'Electrical Telemetering and Automatic Process Control', *Electronic Engineering*, **25**, 302, April 1953, 154.
36. Redding, R.J., 'Intrinsic Safety', *Electrical Times*, **124**, 322, 6th Aug. 1953, 235.
Boundy, J. R., 'Electrical telemetering and process control', *Electronic Engineering*, **25**, 302, 303, April and May 1953, 154, 198.
37. Roper, C. G., 'Electrical Transmitting and Transducing at Process Measurements', *Instruments*, **25**, 3, March 1952, 338.
Roper, C. G. and Engleberger, J. F., 'Electro-mechanical D.C. amplifier', *Electronics*, **20**, 11, Nov. 1947, 117.
38. Davis, E. T., 'An Electronic Control System for Industrial Control Applications', *Instruments*, **24**, 1951, 642.
39. Klepp, E. C., 'Recent Developments in Instruments', *J. Inst. Fuel*, **25**, Jan. 1953, 295,
Klepp, E. C., 'Elliott Trimode', *Elliott House Journal*, **1**, March 1951, 20.
40. Oetker, R. 'Finneuer elektropneumatischer regler', *Feinwerktechnik*, **55**, 1951, 217.
41. Oppelt, W., 'Einige Näherungsformelm zur Berechnung von Regelvorgängen', *Trans. Inst. & Meas. Conf.*, Stockholm 1952, p. 419.
42. Dushkes, S. Z. and Cahn, S. L., 'Analysis of some hydraulic components of Regulators,' *Trans. A.S.M.E.* **74**, May 1952, 595.
43. Howe, W. H., 'Some Recent Developments in Process Control', Ref. 4, p. 217.

P. L., 'Theoretical Considerations and
gulators', *Trans. A.S.M.E.*, 72, April

um Process Controller Settings and the
onic Simulator', *Proc. I.E.E.*, 101, 2, 80,

'A Uniform Approach to the Optimum
ans. A.S.M.E., 76, 8, Nov. 1954, 1265.
r Optimum Control', *Trans. A.S.M.E.*,

Best Form of Response for a Servo
sturbance is Present with the Error

blem in der Regelungstechnik', *Rege-*

einer [illegible] elstrecke mit Ausgleichsgrad,
echnik, ., 1953, 232.
Behaviour Expressed as a Deviation
v. 1954, 1303.
ace Instrumentation', *Inst. Practice*, 8,

teristics in Pneumatic Force-balance
948, 111.
mentation Problems by the Pneumatic
, Sept. 1945.
ct Controller', *Taylor Technology*, 3,

'The Characteristics of Air-operated

of Compound-action Pneumatic Con-
4, 162.
Electronic Process Controller', *Proc.*

ng and Automatic Process Control',
l 1953, 154.
calTimes, 124, 3222, 6th Aug.1953,235.
ng and process control', *Electronic*
May 1953, 154, 198.
and Transducing at Process Measure-
52, 338.
'Elect[illegible]echanical D.C. amplifier',

ystem for Industrial Control Applica-

n Instruments', *J. Inst. Fuel*, 25, Jan.

House Journal, 1, March 1951, 20.
ischer regler', *Feinwerktechnik*, 55,

n zur Berechnung von Regelvorgän-
ckholm 1952, p. 419.
lysis of some hydraulic components
May 1952, 595.
nts in Process Control', Ref. 4, p. 217.

44. Aikman, A. R., MacMillan, J. and Morrison, A. W., 'Static and Dynamic Performance of Industrial Thermometers', *Trans. S.I.T.*, 5, 4, 1953, 138.
45. McAdams, W. H., *Heat Transmission*, McGraw Hill, New York, 1954.
46. Iberall, A. S., 'Attenuation of oscillatory pressures in instrument lines', *J. of Research of N.B.S.*, 45, July 1950, 85.
Moise, J. C., 'Pneumatic Transmission Lines,' *J.I.S.A.*, 1, April 1954, 35.
47. Bradner, M., 'Pneumatic Transmission Lag', *Instruments*, 22, 7, July 1949, 618.
48. Aikman, A. R., 'The influence of measuring and transmission lags', Ref. 4, p. 205.
49. Profos, P., 'Die behandlung von regelproblemen vermittels des frequenzganges des regelkreises', Diss. E.T.H., Zurich 1943.
50. Carslaw, H. S. and Jaeger, J. C., *Conduction of Heat in Solids*, Oxford University Press, 1947.
51. Prinz, D. G., 'Contributions to the theory of automatic controllers and followers', *J. Sci. Inst.*, 21, 4, 1944, 53.
52. Behar, M. F., 'The primary element response', *Instruments*, 21, 8, Aug. 1948, 691.
53. Hornfeck, A. J., 'Response characteristics of thermometer elements', *Trans. A.S.M.E.*, 71, Feb. 1949, 121.
54. Aikman, A. R., 'Frequency response analysis and controllability of a chemical plant', *Trans. A.S.M.E.*, 76, 8, 1954, 1313. Kopiert.
55. Ziegler, J. G. and Nichols, N. B., 'Valve Characteristics and Process Control', *Instruments*, 22, 1, 1949, 75.
56. Walton, H. R., 'Control Valves: Design', *Trans. S.I.T.*, 5, 1, March 1953, 10.
57. Rockwell, R. A., 'Valve rangeability', *Instruments*, 25, 8, 1952, 1074.
58. Brockett, G. F., 'Correlation of Valve Sizing Methods', *Instruments*, 25, 6, 1952, 752.
59. Balls, B. W., 'Control Valves: Application', *Trans. S.I.T.*, 5, 1, March 1953, 20.
60. Ziegler, J. G., 'Cascade Control Systems', *Proc. Texas A. & M. Symposium on Instrumentation*, Jan. ~~1954, and~~ *Taylor Technology*, 6, 3, 1954, 18.
Ziegler, J. G., 'An evaluation of Cascade Control Systems', *Oil & Gas Journal*, 48, April 1954, 95. Bestilt 27/10-23 Svar: Referansen er feil.
61. Fishwick, W., 'Time-lags in Thermometer Pockets', *Trans. S.I.T.*, 4, 1, March 1952, 29.
62. Porter, A., 'Basic principles of automatic control systems', *Proc. I. Mech. E.* 159, 1948, 25.
63. Young, Ll., 'Control of Boilers fired by Solid Fuels in Suspension', *Proc. I. Mech. E.*, 166, 1, 1952, 91.
64. Tivy, V. V. St. L., 'Automatic Control of Fractionating Columns', *Petroleum Refiner*, 27, 11, Nov. 1948, 123.
Boyd, D. M. (Jnr.), Fractionation Instrumentation and Control', *Petroleum Refiner*, 27, 10 and 11, Oct. and Nov. 1948, 533 and 594.
65. Reidel, T. C., 'How to instrument the fractionation section in fuel processing plant', *Oil & Gas Journal*, 52, 51, 26th April, 1954, 136.
Lefkowitz, I., 'Completely Automatic Control System for the Distillation Process', *Instruments*, 25, 6, June 1952, 773.
66. Farquhar, J. C., 'Some aspects of cascade and multi element control systems', *Proc. 3rd World Petroleum Congress*, Sect. 6, Reprint 7.
67. Wallis, S. W. J. and Simpson, R. F. C., 'Concept and practice of coordinated control systems', Proc. I.S.A., 1st International Congress, Philadelphia, 1954, Paper 54.9.1
68. Brown, G. S. and Campbell, D. P., *Principles of servo-mechanisms*, John Wiley, New York, 1948.

69. Chestnut, H. and Mayer, R. W., *Servomechanisms and Regulating System Design*, John Wiley, New York, 1951.
70. Mason, C. E. and Philbrick, G. A., 'Mathematics of surge vessels and automatic averaging control', *Trans. A.S.M.E.*, **63**, 1941, 589.
71. Hoyt, P. R. and Stanton, B. D., 'Use of field test data for control system analysis', *Instruments*, **26**, 8, Aug. 1953, 1180.
72. Smith, E. S., *Automatic Control Engineering*, New York, 1944.
73. Farquhar, J. C., 'Automatic Combustion Control', *Trans. S.I.T.*, **6**, 1, 1954, 29.
74. Box, G. E. P. and Wilson, K. B., 'On the experimental attainment of optimum conditions', *T. Roy. Statistical Soc.*, **13**(B), 1, 1951, 1.
75. Davies, O. L., *Design and Analysis of Industrial Experiments*, Oliver and Boyd, London, 1954.
76. Clarridge, R. E., 'An improved pneumatic control system', *Trans. A.S.M.E.* **73**, April 1951, 297.
77. Toop, G. H., 'A Technique for Minimizing Overshoots in Discontinuous Processes', *Inst. Practice*, **6**, 6 April, 1951, 297.
78. Dickenson, B. W. and Brooks, R. M., 'Neutralisation of Acid Wastes', *Ind. Eng. Chem.*, **42**, 4, 1950, 599.
79. Schoenemann, K., 'The Chemical Yield in Continuous Processes', *Dechema Monograph Band*, 21, 1952.
80. Coulson, J. M. and Richardson, J. F., *Chemical Engineering*, Pergamon Press, London, 1954.
81. Fishenden, M. and Saunders, O. A., *An introduction to Heat Transfer*, O.U.P., London, 1950.
82. Gee, R. E., Linton, W. H., Maier, R. E. and Raines, J. W., 'The Use of Computers in Kinetic Calculations', paper presented at A.I. Chem. Eng. Conf., Washington, March 1954. Tech. Section, No. 6.
83. Report of A.I.E.E. Feedback Controls Systems Committee, *Electrical Engineer*, **70**, October 1951, 905.
Philbrick, G. A., 'Unified Symbolism for Regulatory Controls', *Trans. A.S.M.E.*, **69**, Jan. 1947, 47.
84. Young, A. J. 'The Development of Automatic Process Control', *Trans. Inst. & Meas. Conf.*, Stockholm, 1952, p. 428.
85. 'Automatic Control Terminology', *Trans. A.S.M.E.*, **74**, 6, June 1952, 486. [Report of Terminology Committee of A.S.M.E. (I.R.D.)].
86. Young, A. J., 'A.S.M.E. and B.S.I. Terminologies for Process Control', Proc. I.S.A., 1st International Congress Philadelphia, 1954, Paper 54-IRD-10.
87. Chaplin, A. L., 'Some application factors affecting the regulation of pH in Industry', *Instruments*, **22**, 7, July 1949, 579.
88. 'Trends in Computers: Automatic Control and Data Processing', Proc. Western Computer Conference, Los Angeles, Feb. 1954, published by A.I.E.E., New York.
89. Cross, R. E., 'Trends in Machine Tool Automation', *Inst. & Automation*, **27**, 5, May 1954, 768.
90. June, A. S., *et al.*, 'The Automatic Factory—a critical examination', 8 parts, *Inst. & Automation*, **27**, 9 (p. 1476) to 2 (p. 277) 1955.
91. Diebold, J., *Automation*, D. van Nostrand Co., New York, 1952.
92. Metral A. R., *Revue Générale de Mecanique*, Vol. 34—see particularly 'Machines Universelles ou Machines Spéciales ou Machines Transferts', **34**, 17, May 1950, 155.
93. 'Automation in American Industrial Plants', Report of I.R.—11,072, U.S. Dept of Commerce, Office of Technical Services, Washington, D.C., 1954.
94. Redding, R. J., 'Electronic Process Control in an Oil Refinery', *Inst. Practice*, **6**, July 1952, 597. Also Marchment, W. T., *J.I.S.A.*, **1**, 3, March 1954, 22.
95. Stewart, L., 'Electronically Controlled Refinery,' *Inst. and Automation*, **27**, 12, Dec. 1954, 1949.

A commonly occurring example of an open loop system is found in the flow ratio control system of Fig. 2.5 (a). It will be seen that any change in pressure drop across valve V_2 will cause the flow rate through V_2 to vary, for any given setting called for by F.C.1. The ratio of the two flows will therefore not be held constant.

When the ratio of the two flows is critical, the system of Fig. 2.5 (b) is used. Here the proportional controller or flow transmitter F.C.1 alters the desired value set in the flow controller F.C.2, so that F.C.2 holds the flow rate at the desired value as set by F.C.1.

Open loop control systems are not generally employed in process control, except where conditions are very stable or to reduce disturbances before they enter a main, closed loop system. Examples of such use are given in the literature (5).

2.4 Conclusion

It has been emphasised that a control system is essentially a 'closed loop' system, of which the plant forms an integral part.

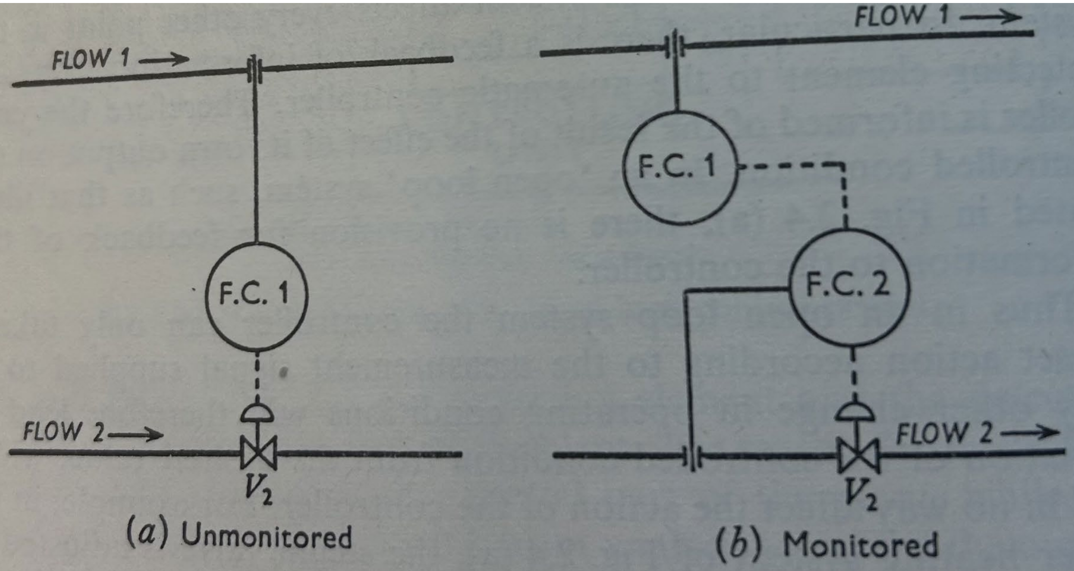

FIG. 2.5. Flow ratio control systems.