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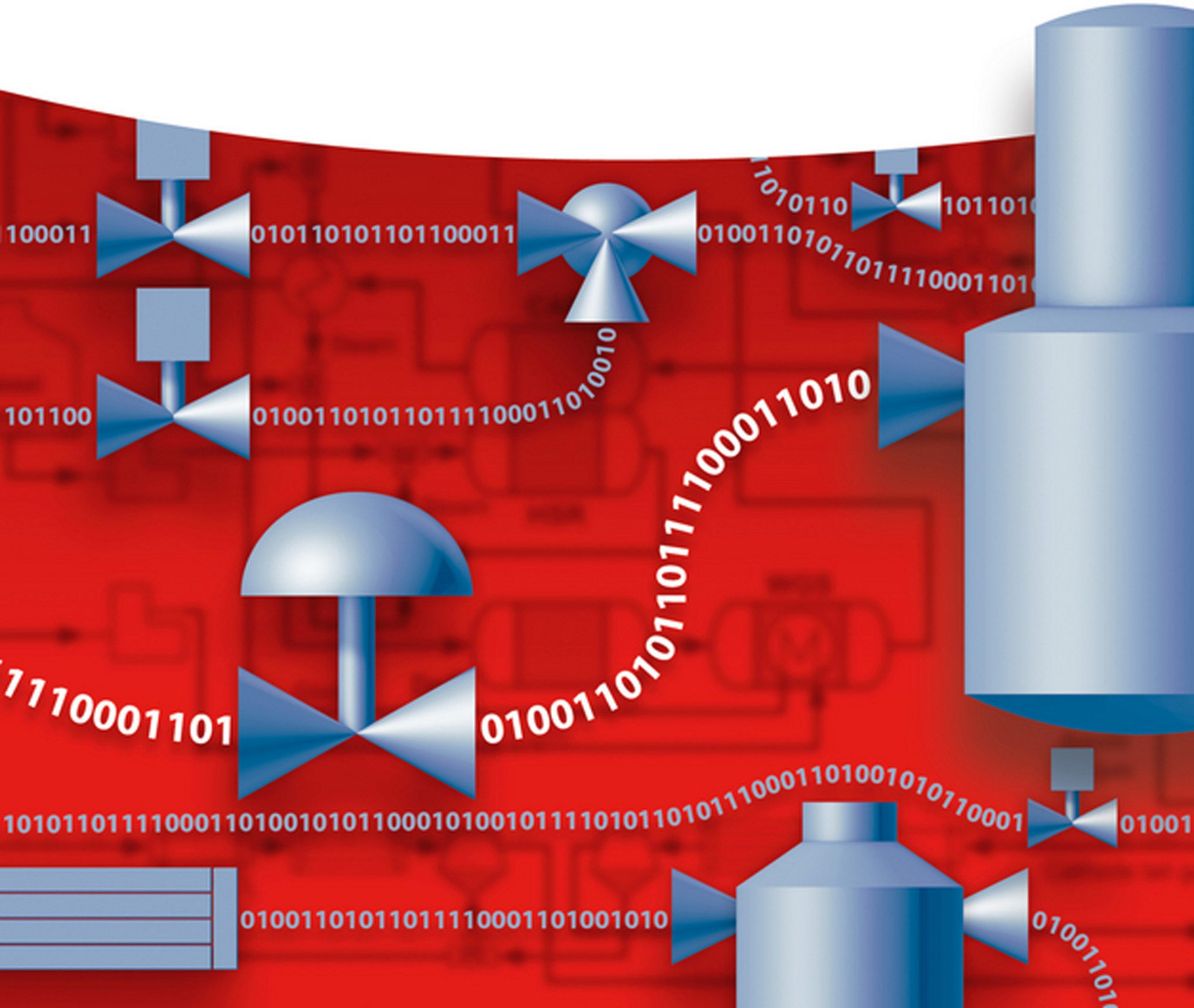

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Applied Process Control

Essential Methods

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Preface

Material in this book is sequenced for the process engineer who needs 'some' background in process control (Chapters 1–5) through to the process engineer who wishes to *specialise* in advanced process control (Chapters 1–9). The theory needed to properly understand and implement the methods is presented as succinctly as possible, with extensive recourse to linear algebra, allowing multi-input, multi-output problems to be interpreted as simply as single-input, single-output problems.

Before moving on to the more advanced algorithms, an essential practical background is laid out on plant instrumentation and control schemes (Chapters 2, 4 and 5). Chapter 3 builds modelling abilities from the simplest time-loop algorithm through to discrete methods, transfer functions, automata and fuzzy logic. By the end of Chapter 5, the engineer has the means to design simple controllers on the basis of his or her models, and to use more detailed models to test these controllers. Moreover, ability has been developed in the use of the multi-element control schemes of 'advanced process control'.

Chapter 6 focuses on observation. Whereas Chapter 3 reveals the tenuous chain of preparation of plant signals, Chapter 6 aims to make sense of them. Important issues on the plant are signal conditioning, data reconciliation, identification of model parameters and estimation of unmeasured variables.

Chapter 7 addresses more advanced control algorithms, drawing on a wide range of successful modern methods. To a large extent, continuous and discrete versions of an algorithm are presented in parallel, usually in multi-input, multi-output formats – which simply devolve to the single-input, single-output case if required. State–space, input–output, fuzzy, evolutionary, artificial neural network and hybrid methods are presented. There is a strong emphasis on model predictive control methods which have had major industrial benefits.

A review of the classical methods of stability analysis is delayed until Chapter 8. This has been kept brief, in line with reduced application in the processing industries. One recognises that stability criteria, such as pole locations, do underlie some of the design techniques of Chapter 7. Certainly, frequency domain concepts are part of the language of control theory, and essential for advanced investigation. But with the slower responses and inaccurate models of processing plants, controllers are not predesigned to 'push the limits' and tend to be tuned up experimentally online.

A review of a range of optimisation techniques and concepts is given in Chapter 9. Although not a deep analysis, this imparts a basic working knowledge, enabling the development of simple applications, which can then later be built upon. Topics covered include *linear*, *integer*, *mixed*, and *non-linear* programming, search techniques, global optimisation, simulated annealing, genetic algorithms and multi-objective optimisation. These methods, and *dynamic programming*, underlie the predictive control and optimal scheduling topics in Chapter 7, and are also important as static optimisers in such applications as supply chain, product blending/distribution and plant economic optimisers.

This book tries to make the methods practically useful to the reader as quickly as possible. However, there is no shortcut to reliable results, without a basic knowledge of the theory. For example, one cannot make proper use of a Kalman filter, without understanding its mechanism. Complex multi-input, multi-output applications will require a good theoretical understanding in order to trace a performance problem back to a poorly calibrated input measurement. Hence, an adequate theoretical background is provided.

A few distinctions need to be clarified:

1) Modelling is a particular strength of the process engineer, and is a basis of all of the algorithms – especially model predictive control. The reader needs to distinguish *state-based* models versus *input–output* models. The state-based models can predict forward in time knowing only the initial state and future inputs. Some algorithms rely on this. In contrast, *input–output* models will need additional information about past inputs and outputs, in order to predict future outputs. To use state-based algorithms on these, a state observer algorithm (e.g. Kalman filter) will be required to estimate the states.
2) The forward shift operator $z = e^{Ts}$ is used to relate discrete versions of systems to their transfer function forms $G(s)$ in the s (Laplace/frequency) domain. In a lot of what follows, this theoretical connection is not significant, and the data sampling shift parameter q could be used, but sometimes it is not in this text.
3) The text consistently uses bold characters to signify matrices [$\boldsymbol{A}$], vectors [$\boldsymbol{x}$] and matrix transfer functions [$\boldsymbol{G}(s)$, $\boldsymbol{G}(z)$]. Non-bold characters are used for scalars.

A number of examples are presented in this book in order to clarify the methods. In addition, the separate accompanying book *Applied Process Control: Efficient Problem Solving* presents 226 solved problems, using the methods of this text. These often make use of MATLAB® code which is arranged in obvious time loops, allowing easy translation to the real-time environment. There will, however, be the challenge to provide additional routines such as matrix inversion.

A simple interactive simulator program has been made available at https://sourceforge.net/projects/rtc-simulator/. It includes 20 different applications for such aspects as PID and DMC controller tuning, advanced level control, Smith prediction, Kalman filtering and control strategies for a furnace, a boiler and a hybrid system. No support is available for the simulator.

Although I have personally used a variety of methods on industrial and research applications, in writing this book I have been fascinated to discover the brilliant ideas of many other workers in the field. To all of those people who get excited about process control, I wish you an optimal trajectory.

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March, 2016

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Many of the problems in this book are dealt with using the MATLAB® program, which is distributed by the MathWorks, Inc. They may be contacted at

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A few problems are dealt with in the GAMS® optimisation environment, distributed by

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Contact: sales@gams.com

Some problems make use of the LPSOLVE mixed integer linear programming software which is hosted on the SourceForge Web site at

http://sourceforge.net/projects/lpsolve/

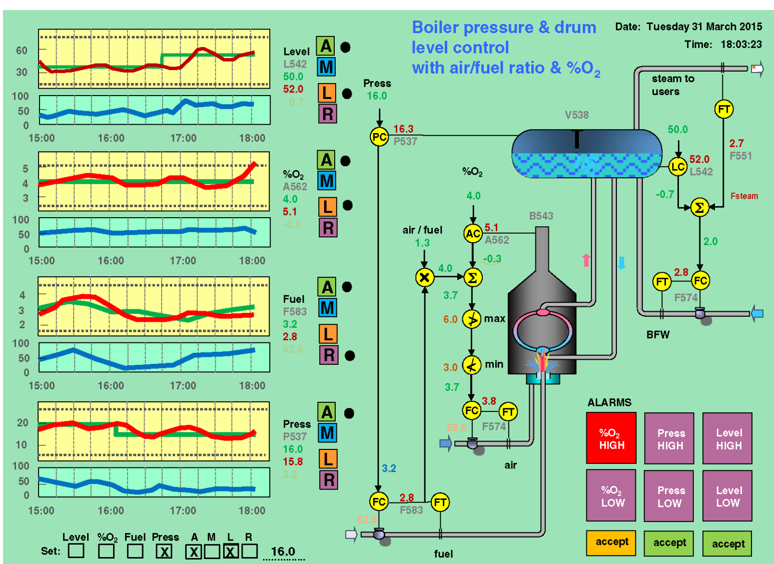
Abbreviations

A/D	analogue to digital
AC/FO	air to close/fail open
ANN	artificial neural network
AO/FC	air to open/fail closed
APC	advanced process control
ARIMAX	autoregressive integrated moving average exogenous
ARMAX	autoregressive moving average exogenous
ARX	autoregressive exogenous
BB	branch and bound
BFW	boiler feedwater
BIDP	backward iterative dynamic programming
CEM	cause and effect matrix
CRT	cathode ray tube
CV	controlled variable
CVP	control variable parameterisation
CW	cooling water
D/A	digital to analogue
DAE	differential and algebraic equations
DCS	distributed control system
DMC	dynamic matrix control
DP	differential pressure
DV	disturbance variable
E{ . . . }	expectation of . . .
EKF	extended Kalman filter
ES	evolutionary strategy
FFT	fast Fourier transform
FIDP	forward iterative dynamic programming
FIMC	fuzzy internal model controller
FIR	finite impulse response
FRM	fuzzy relational model
FRMBC	fuzzy relational model-based control
FSQP	feasible sequential quadratic programming

Fuzzy	fuzzy logic
GA	genetic algorithm
GAMS	General Algebraic Modelling System®
GM	gain margin
GPC	generalised predictive control
HP	high pressure (port)
HS	high select
I	identity matrix
I/O	input–output
I/P	current to pressure (pneumatic) converter
IAE	integral of absolute error
IDP	iterative dynamic programming
IMC	internal model control
INA	inverse Nyquist array
IO	input–output
IP	integer programming
ISE	integral of squared error
KO	knockout (separation drum)
LAN	local area network
LBT	lower block triangular
LCD	liquid crystal display
LDMC	linear dynamic matrix control
LP	linear programming
LP	low pressure (port)
LPG	liquefied petroleum gas
LPSOLVE	MILP program (http://sourceforge.net/projects/lpsolve/)
LQR	linear quadratic regulator
LS	low select
LS	least squares
MATLAB	MATLAB® program, distributed by the MathWorks, Inc.
MEK	methyl ethyl ketone
MIDO	mixed integer dynamic optimisation
MILP	mixed integer linear programming
MIMO	multi-input, multi-output
MINLP	mixed integer nonlinear programming
MIP	mixed integer programming
MIQP	mixed integer quadratic programming
MLD	mixed logical dynamical
MM	molecular mass
MPC	model predictive control
MRI	Morari resiliency index
MTBF	mean time between failures
MV	manipulated variable
NC	normally closed
NLP	nonlinear programming
NO	normally open

NSGA-II	fast non-dominated sorting genetic algorithm II
OA	outer approximation
ODE	ordinary differential equation
OHTC	overall heat transfer coefficient
P	proportional
P/I	pressure (pneumatic) to current converter
PCA	principal components analysis
PDE	partial differential equation
PI	proportional integral
PID	proportional integral derivative
PLC	programmable logic controller
PM	phase margin
PV	process variable
QDMC	quadratic dynamic matrix control
RAID	redundant array of independent discs
RGA	relative gain array (Bristol array)
RLS	recursive least squares
RTD	resistance temperature detector
RTO	real-time optimisation
SCADA	supervisory control and data acquisition
SG	specific gravity
SISO	single-input single-output
SP	setpoint
SQP	sequential quadratic programming
VPC	valve position control
WABT	weighted average bed temperature
WAN	wide area network (e.g. using telecommunication, radio)
WG	water-gauge
ZOH	zero-order hold

Old analogue control panel



Modern digital control display

1
Introduction

1.1
The Idea of Control

Plant and animal life relies on numerous control mechanisms. Here one thinks of an 'open-loop' *causal* system in which cause (input) generates effect (output). Then the 'control' part operates around this in order to modify the effect (Figure 1.1).

All autonomous entities need 'feedback' paths like this. One might shut a door to reduce the noise level, eat to reduce hunger or turn the car steering wheel to keep in a lane. The control decision can be discrete, such as when the geyser thermostat switches on or the reserve bank adjusts the prime lending rate in order to control inflation, or it can be continuous, such as the variation of one's iris with available light. The 'system' part can have various behaviours which make the decision difficult. Thus, when the shower it too hot, one cautiously increases the cold water, knowing that there is a ('dead-time') delay in the pipe. The doctor will adjust blood pressure medication in small steps as he/she waits to observe the body response. One also recognises that there are possibly multiple 'causes' at play. It may only be possible to observe a limited range of the effects, and usually one is only manipulating a small selection of the causes. Humans are not adept at coordinating multiple inputs.

What is common in these cases of feedback control is that one does not know where the inputs should be set exactly. The control problem amounts to 'given desired levels of the outputs, at what levels should the inputs be set in order to best achieve this?'. So the 'decision' is a task of *inversion*, in just the same way as one might want to find an x such that a function $f(x) = 0$. A simple control law for this mathematical task was provided by Newton (Equation 1.1).

$$x_{n+1} = x_n - \frac{f(x_n)}{f'(x_n)} \tag{1.1}$$

This is closely related to the 'dead-beat' controller (Section 7.1.1), in which one attempts to hit the target on every time step.

It is intuitive that one can expect difficulty with automatic feedback adjustments. The decision to adjust the inputs will affect the outputs which in turn will affect the next decision. If one is over-reacting on each step, the output would be driven past its desired level by successively larger amounts. So adjusting the shower water in too big steps would successively cause scalding and freezing by greater amounts. The possibility of such endless growth makes considerations of *closed-loop stability* important in the study of process control.

Applied Process Control: Essential Methods, First Edition. Michael Mulholland.

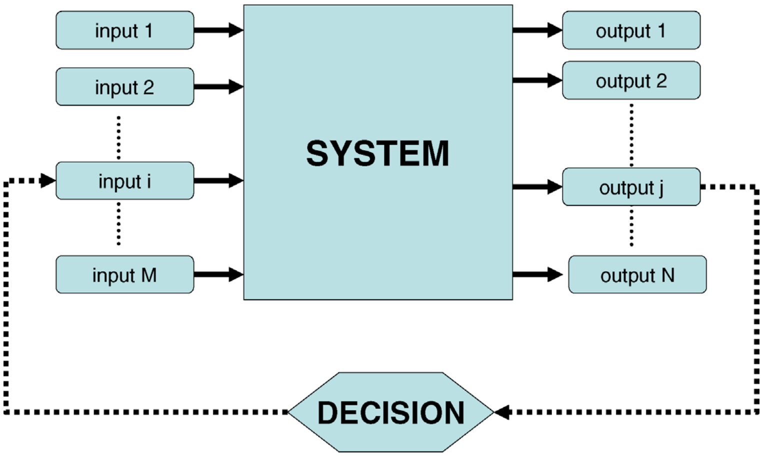


Figure 1.1 Feedback control mechanism.

A situation often arises where one tries to diminish the impact of disturbed inputs to a system by manipulating other inputs (Figure 1.2). This is 'feedforward' control. It is important to note that the outputs are not involved in this decision at all. Of course, one needs to be able to observe the relevant inputs first. Thus, the fuel to a boiler may be increasing in order to maintain pressure. Though the flue gas oxygen content may be unmeasured, a feedforward controller can increase the combustion air flow in proportion to the fuel flow in order to maintain a margin of oxygen excess. One realises that feedforward control will always require some kind of model, for example the air/fuel stoichiometric ratio. Models are never perfect, so it is likely that the relevant output may not be quite where it was planned. Often this error can be tolerated. In other cases, a feedback loop may be superimposed to provide the correction (Figure 1.3). There is nevertheless a benefit in using feedforward to eliminate most of the upset.

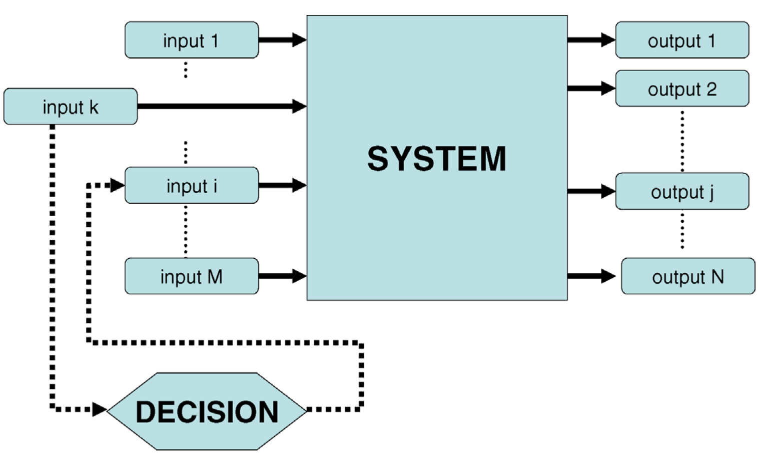


Figure 1.2 Feedforward control mechanism.

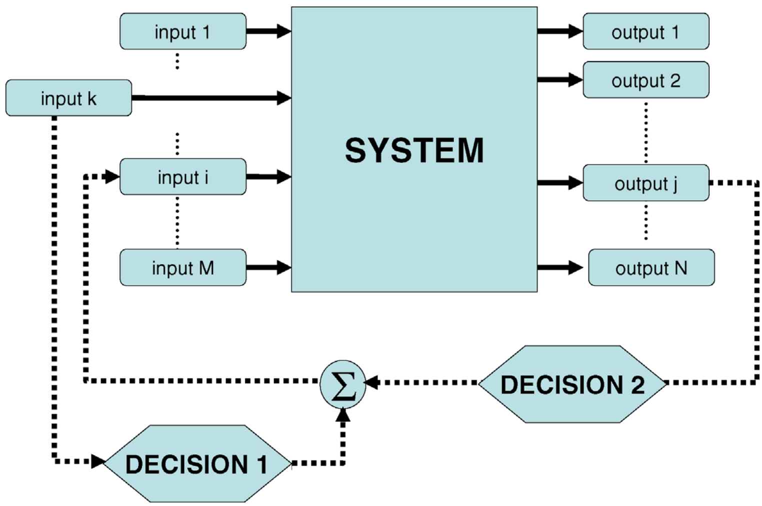


Figure 1.3 Combined feedforward and feedback.

The control engineer must learn to recognise 'information flows' in a system, that is from 'cause' to 'effect'. Sometimes these are not intuitive and have little to do with the physical arrangement. For example, the flow into a tank and the flow out of a tank could equally be used to affect the level in a tank.

1.2 Importance of Control in Chemical Processing

This book will focus on modelling, estimation, control and optimisation in the processing industries. There are unique challenges here to do with the inaccuracy of models and undefined disturbances. In addition, the widespread use of computers to handle process instrumentation in recent decades has spurred the concept of 'advanced process control' (APC), which has become a specialised process engineering domain. The objective is to take advantage of the plant-wide view of outputs, and access to inputs, of these computers, in order to enhance regulation and optimisation. In this way, industries have been able to work safely with narrower specifications and less loss (Figure 1.4). With the increasing globalisation of markets, industries which do not seek such efficiency improvements will soon find themselves uncompetitive and out of business.

In the processing industries, the automatic control aspects are viewed to constitute a pyramid of three main layers in which each layer achieves its objectives by supervising the layer below. Generally, this means that the control loop setpoints (SPs) are passed downwards (Figure 1.5).

Usually the base layer becomes the responsibility of instrumentation technicians, but more advanced inputs are required from control engineers in the upper layers. Of course, the overall control scheme, including the base layer, must be specified by engineers in the design phase. At that stage, additional specifications may be made, such as increased vessel hold-ups to facilitate 'advanced level control' (Section 4.11). Indeed, there is a growing trend to integrate the equipment design and control design at an early stage (Sakizlis, Perkins and Pistikopoulos, 2004). Increasing

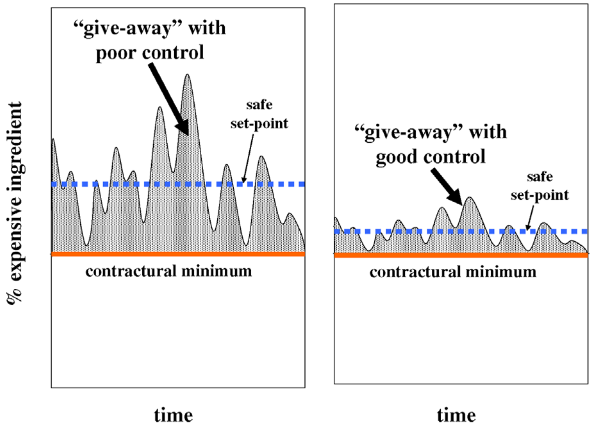


Figure 1.4 Reduction of an expensive ingredient through better control.

integration of processes through 'pinch' analysis often renders the internal regulation highly interactive, requiring special control approaches. Another lesson that has been learned is that the advanced control algorithms cannot simply be installed and left to operate without ongoing knowledgeable oversight.

The advanced algorithms focus on criteria such as throughput, product specifications and economics, not necessarily smooth process operation, and thus they can be unpopular with operating personnel. All too often such unwelcome behaviour can cause operators to switch off these algorithms. Thus, education is important, as well as investigation of downtime incidences and constant reviewing of performance. On one level, one aims to make a control scheme as simple and transparent as possible, to facilitate understanding. However, some algorithms are unavoidably complex,

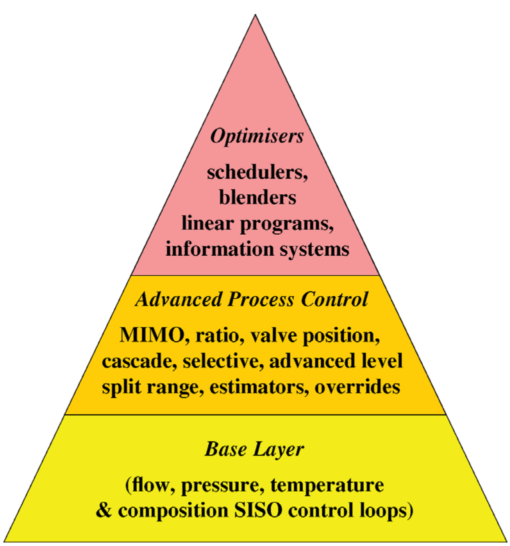


Figure 1.5 Main conceptual layers in a processing plant control scheme.

referring to a number of measurements as the basis of their output decisions. A specially trained control engineer is required to diagnose poor performance that might arise from a poorly calibrated measurement. Industries which recognise the need for ongoing care, and provide the necessary resources, have successfully increased the fractional online time of their optimisers and advanced controllers (Karodia, Naidoo and Appanah, 1999).

1.3 Organisation of This Book

With the increasing use of computers and digital communications in the processing industries, the employers' requirements of a process engineer are rapidly evolving. An engineer working day to day on troubleshooting and debottlenecking on a plant must already have a high proficiency in the use of computers for data extraction and analysis. He/she must comprehend a complex plant control scheme in order to identify the relationships between variables arising from such a scheme. Where a process engineer is specifically employed in the area of 'advanced process control', there will be even more of an expectation that good computer skills will be brought to bear on processing issues. The general brief given to such an engineer by the employer is likely to be: 'Do whatever you can with this computer know-how to safely maximise the profitability of this process!'. Well, that is a very open-ended request, extending far beyond the traditional 'process control' skills of drawing Nyquist plots to assure closed-loop stability.

At the outset, the strength of the process engineer in this environment is his/her understanding of the physics and chemistry at play, leading on to his/her ability to mathematically model the process. The model can be used effectively as the mathematical basis of the control scheme design, or simply to test proposed strategies developed in other ways. Regardless, it is the model that gives this engineer a clear insight into the process behaviour, and it provides an interpolative bridge between what is usually too few plant measurements.

Beyond the model, and now thinking of an alternative 'black box' approach, what will be important is to get an appreciation of a range of *ad hoc* methods of identification, control and optimisation which have proved useful industrially. Some of these defy mathematical treatment, and will not easily lead on to proofs of stability, which has been a major preoccupation of the field. Rather, the main purpose of mathematical treatments will be to promote understanding, and to allow one to move on quickly to useful algorithms for online implementation.

In order to establish the context of the considered algorithms, a view will initially be developed of the instrumentation and computer hardware required for their implementation. In connection with this, it will be important to understand conventions for representation of instrumentation and control on plant piping and instrumentation diagrams.

The present-day field of process control is built conceptually on remnants and artefacts of the past. For example, a computer representation of a loop controller has the equivalent switches and adjustments of the preceding panel-mounted analogue device. An engineer might claim that a control loop that oscillates a lot and won't settle down 'has a poorly damped closed-loop pole location'. One would be extremely hard put to find an application of such frequency response methods in the processing industries, yet this is the language that is naturally used. Why is this so? Well, the reason is that one's mind picture of the phenomena is built mostly on classical control theory. Though most of what is presently done in industry is based on the time domain, one ignores classical theory such as 'frequency response' at one's peril, because it is part of the language, and in many instances it is the route forward to deeper analysis and research. In some cases, for example the use of Laplace domain transfer functions, classical approaches give a much clearer view of relationships. The classical methods will thus also be used in parallel where appropriate.

Working on from simple controllers, more advanced algorithms for estimation and control will be considered, finally viewing the application of optimisation algorithms. Along the way, skills will be developed in the overall instrumentation and control of a process, effectively what is necessary to specify the key plant document, namely the piping and instrumentation diagram. Methods of quantifying and describing control performance and stability will be presented, largely connecting to the classical theory.

1.4
Semantics

Some concepts and related vocabulary in process control need to be clarified initially to avoid confusion:

System: all or part of a process which can be viewed in isolation (provides output values in response to input values).

Dynamic: the mathematical description involves a derivative with respect to time, or a time delay.

Static or *algebraic*: input values immediately determine output values.

Lumped: no spatial derivatives are involved in the mathematical description.

Distributed: variations also occur in space (e.g. position within a reactor bed), requiring spatial derivatives in the mathematical description.

Order: number of time derivatives of different variables involved in the mathematical description (each higher derivative also contributes 1 to the count).

States: a selection of variables describing a system such that if their initial values are known, and all future inputs are known, all future values of the states can be predicted. Effectively these are the variables in the set of first-order derivatives describing the system, so the *order* is equal to the number of *states*.

Open loop: information generated in the output does not influence the input.

Closed loop: information generated in the output is used to influence the input.

Stable: a system is stable if its outputs are bounded (non-infinite) for all bounded inputs.

Unstable: at least one bounded input excitation can cause an unbounded output – usually manifested as exponentially increasing oscillation or magnitude. Usually this type of behaviour is restricted to a *limit cycle* or final magnitude because of the physical limits of the equipment – unless failure occurs before this point.

Step response: output variation resulting from a step in one of the inputs.

Frequency response: output characteristics when the input is a steady oscillation (varies with frequency).

Tuning: choice of free parameters for controllers, estimators or optimisers, to obtain desired performance.

Controlled variable (CV): one of the outputs for which tracking of a *setpoint* is required.

Manipulated variable (MV): one of the inputs which is available to be varied by a controller.

Disturbance variable (DV): one of the inputs which is not available for manipulation.

Dead time: this is a time delay (usually caused by a plug flow transport lag).

Inverse response: the initial direction of the response (up/down) differs from the final position.

References

Karodia, M.E., Naidoo, S.G. and Appanah, R. (1999) Closed-loop optimization increases refinery margins in South Africa. *World Refining*, **9** (5), 62–64.

Sakizlis, V., Perkins, J.D. and Pistikopoulos, N. (2004) Recent advances in optimization-based simultaneous process and control design. *Computers & Chemical Engineering*, **28** (10), 2069–2086.

2
Instrumentation

Plant instrumentation constitutes 6–30% of the cost of all purchased equipment for the plant, and thus up to 5% of total project costs for new plants (Peters and Timmerhaus, 1980). A perusal of the instrumentation trade magazines shows what a fiercely competitive market this is. Suppliers are developing a never-ending range of measurement and actuation devices based on new technologies and materials. In order to understand the context of the control schemes and algorithms addressed in this book, one needs a basic appreciation of the devices available.

2.1
Piping and Instrumentation Diagram Notation

The piping and instrumentation diagram is the most important reference document in both the construction and functioning of a processing plant (Davey and Neels, 1991). Not only is it the means of understanding the operation, but its indexing system also allows location of documents pertaining to individual plant items (such as reactors and pumps), the pipework and the measurement and control equipment and philosophy (Figure 2.1).

Conventions for showing instrumentation vary with company and national standards (e.g. BS 1646 and DIN 28004). Some examples are presented in Figure 2.2. Up until the 1980s, analogue instrumentation data handling was still common, requiring some specific conventions such as in Figure 2.3.

The development of instrumentation that could perform the required operations (such as PID control or square-root extraction) in a purely analogue context became a very sophisticated art, leading to ingenious fluidic (pneumatic) or electronic circuitry. Apart from the high purchase costs, there was a heavy burden of maintenance, so the advent of data-handling systems based on digital computers was warmly welcomed. On economic grounds, most analogue systems were quickly dumped in favour of modern DCS (distributed computer systems), SCADA (supervisory control and data acquisition) and PLCs (programmable logic controllers) (Figure 2.4). Such systems are highly configurable. No longer would the recording of a signal require a dedicated panel-mounted strip or circular chart (frontispiece). All data could be continuously recorded in a historical database, and accessed at will. Alarm limits could be defined for every signal. The control panel along which sociable operators used to stroll, tapping recorders to ensure free pen movement, disappeared, and was replaced by one or more CRT or LCD displays. The art of being an operator changed, and quite a few of the 'old school' moved on. In the digital systems, the functional distinctions 'I' and 'R' in Figure 2.3 have been lost and are omitted from tags.

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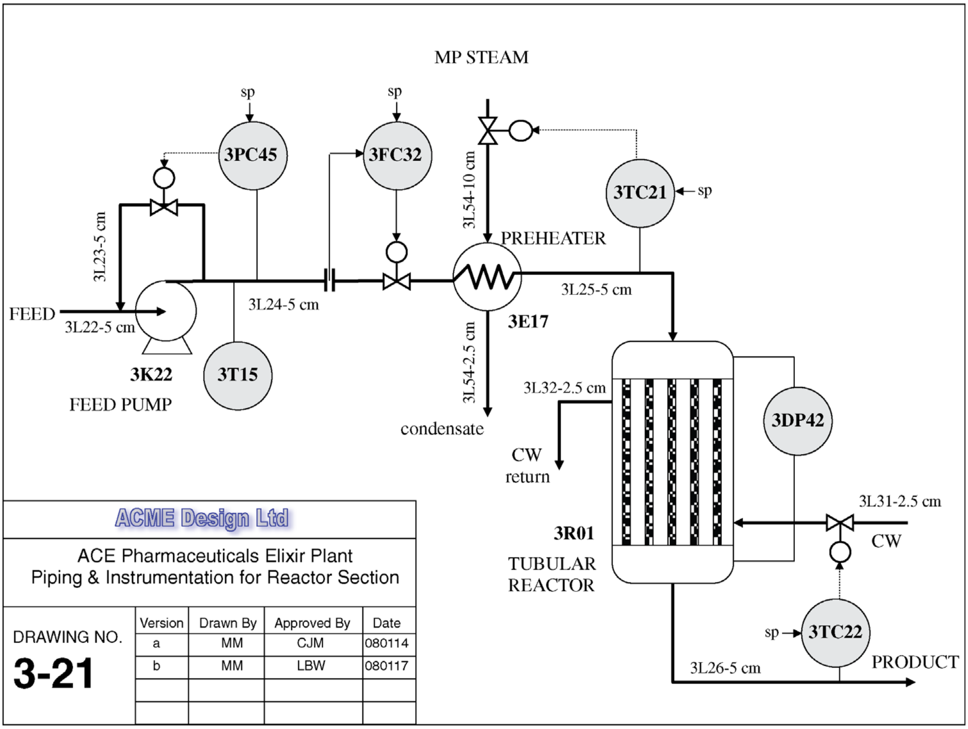

Figure 2.1 Example of a piping and instrumentation diagram.

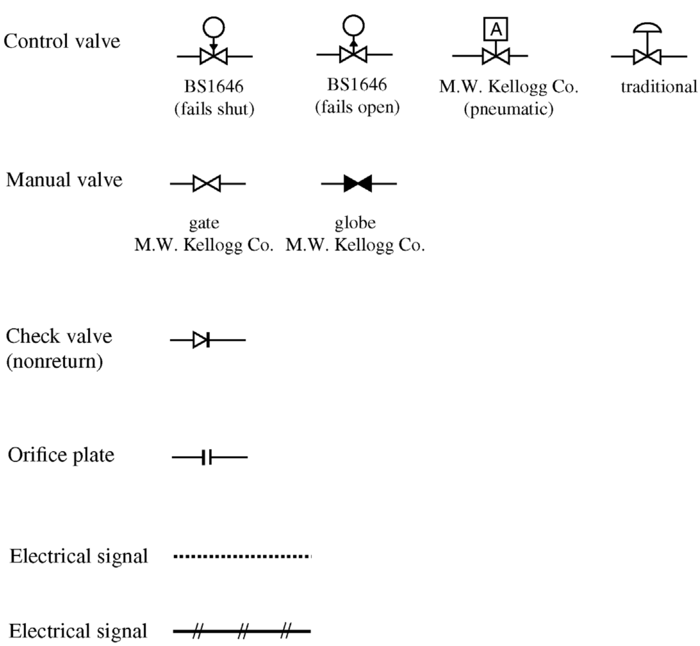

Figure 2.2 Examples of instrumentation symbols.

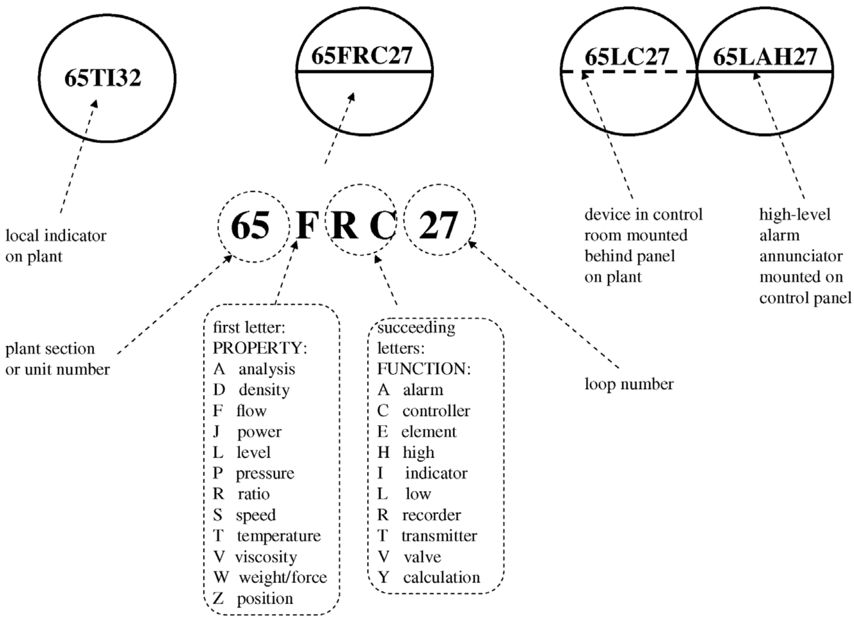

Figure 2.3 Instrumentation tag bubble notation.

The growth of the digital systems out of the old analogue systems constrained manufacturers (probably needlessly) into a sort of item-per-item replacement. Clearly, this would reduce the shock for existing personnel. The square-root extraction, ratio, control, alarm and trip calculations stayed in functional blocks which could be interconnected just like the old analogue signals. The computer 'faceplates' of PID controllers had the recognisable adjustments from the old control room panel. However, beyond the direct replacement of the old functionality, industries realised that they were now in a completely new situation where every measurement and every control actuator could be accessed simultaneously, and there was almost no constraint on the complexity of any calculation. This opened up a whole new world for control engineers to take the concept of 'advanced process control' a lot further.

2.2 Plant Signal Ranges and Conversions

One legacy of the analogue era is the handling of measurement and actuator signals in conventional ranges such as

4–20 mA;

3–15 psig;

20–100 kPag.

With new fieldbus devices, instruments can communicate with the computer control system digitally, avoiding the need to convert current or pneumatic signals into the accepted ranges.

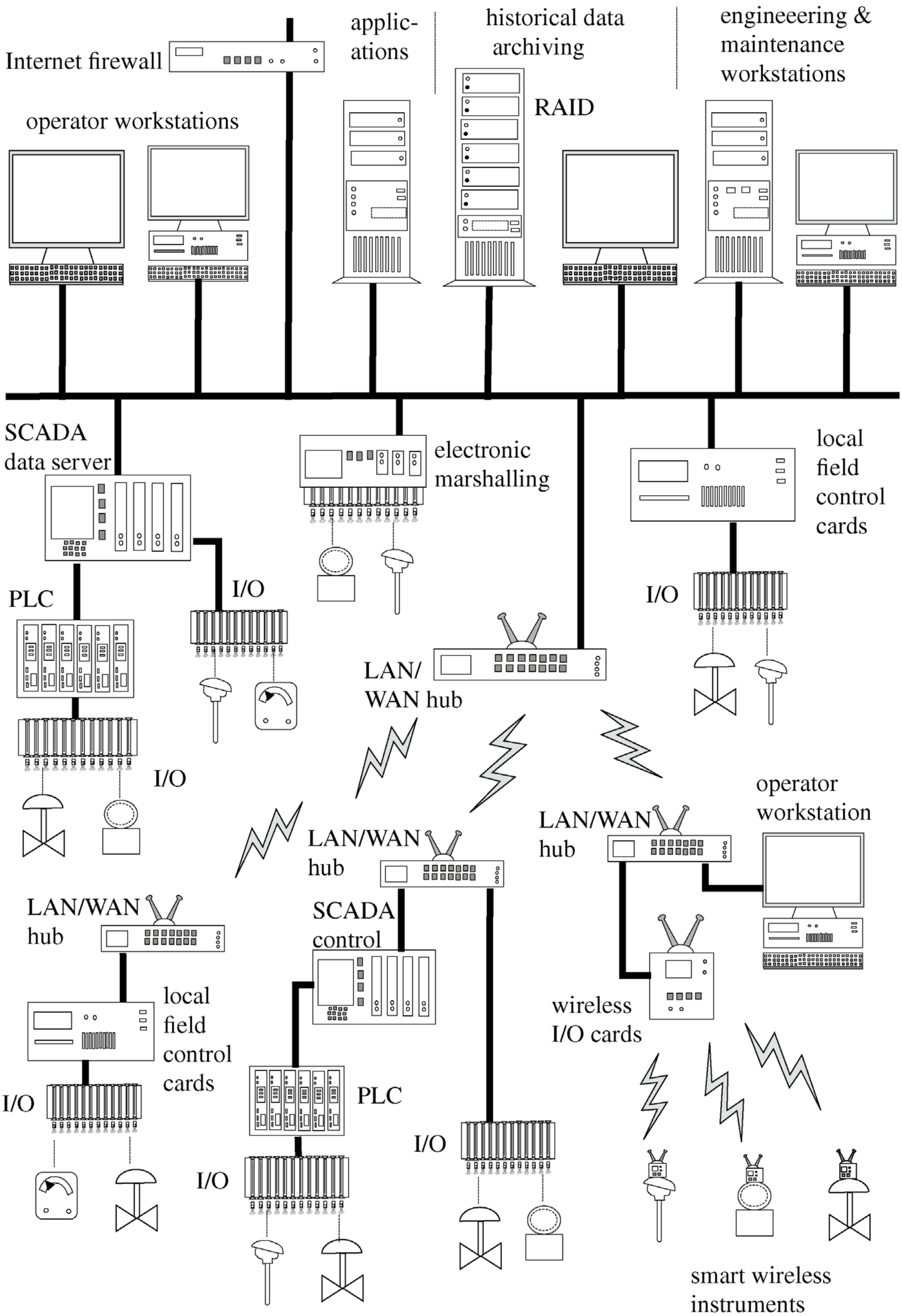

Figure 2.4 Distributed control system structure illustrating a range of features.

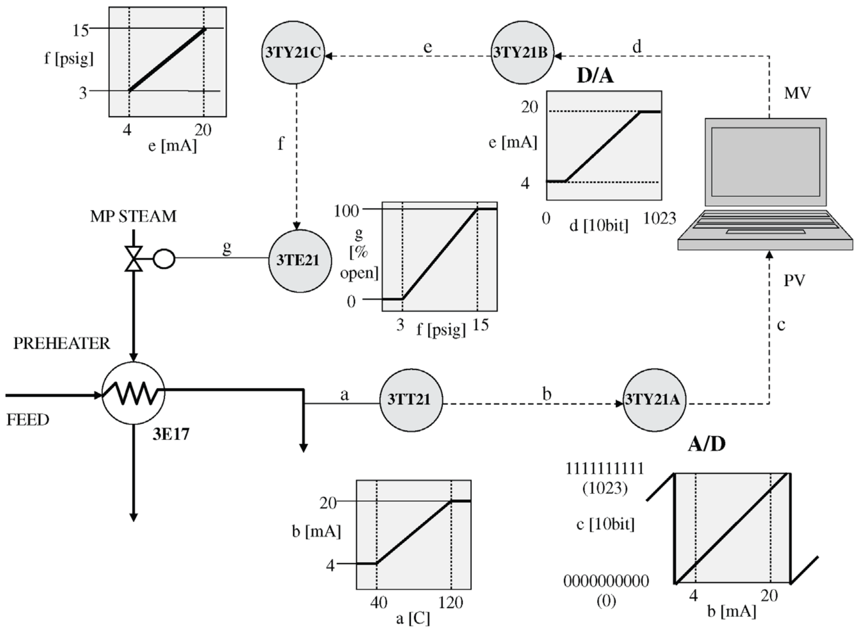

Figure 2.5 Typical signal conversion sequence.

Nevertheless, much instrumentation is still being designed around the analogue signal range concept, and one has to be aware that any device will anyway have lower and upper saturation limits. On large plants, these analogue signals are handled as close to the associated plant equipment as possible in signal substations. Signal conditioning (such as smoothing) and A/D or D/A (analogue-to-digital or digital-to-analogue) conversion are handled there, so that communication with distributed and centralised computer facilities thereafter occurs on digital buses (e.g. coaxial, fibre-optic or 'wireless' transmissions).

A typical signal conversion sequence is given in Figure 2.5. Note that the information flow, starting from the temperature sensor (a), passes through a sequence of calibrations, in this case each involving an intercept and a slope. Each of these is subject to drift and error, so it is only a foolhardy engineer who simply accepts the engineering values presented in the computer display. A nonlinearity arises from the saturation conditions determined by the conventional signal ranges and the A/D and D/A conversion windows. For example, a 20 mA signal (b) might represent an original temperature measurement (a) anywhere above 120 °C. Regarding the 20% offset of the minimum transmission signal values from zero (e.g. 4 mA, 3 psig), this was originally to aid signal loop continuity checking, since the most likely fault would be an open circuit.

For a pneumatically actuated control valve (air-to-open) as suggested here, the intercept and slope (*zero* and *span*) are determined by two nuts on the valve stem which position and tension the return spring. It is conceivable that wear or incorrect set-up might result in a 3 psig signal not quite closing the valve, say leaving it 4% open. There is no feedback of this position, so the operator would be unaware that live steam is still in the system, for example. Often the nuts might be set to close the valve at, say, 4 psig, to ensure a really tight shut-off when a 0% (3 psig) open position is requested. These are but some of the pitfalls to be expected, so the engineer would do well to maintain a healthy scepticism regarding the validity of data on the system and should cross-check wherever possible.

2.3
A Special Note on Differential Pressure Cells

The differential pressure (DP) cell is the ubiquitous workhorse of industrial instrumentation, playing a part in the measurement of flow, pressure and level (Figure 2.6). It is a transmitter in the sense that it receives a pressure signal, converts it linearly into the desired signal range (e.g. 4–20 mA, 20–100 kPag) and retransmits it. In cases where it transmits a current signal, it is sometimes referred to as a P/I transducer (pressure/pneumatic-to-current). Flow measurement techniques relying on the creation of a pressure difference (e.g. orifice plates) usually attempt to work with as small a Δp as possible to minimise pumping power. So in this case the DP cell might receive a signal of, say, 0–20 ″WG, and retransmit in the 3–15 psig range. Here one can think of the device as a high-gain amplifier. Conversely, the pressure inside an ammonia synthesis loop (e.g. 150 barg) might be compared with atmospheric pressure, and the Δp retransmitted in the 3–15 psig range.

DP cells will be installed close to the point of measurement, so there could still be a need to use a completely pneumatic device (as above) for intrinsic safety in the presence of flammable gases. Commercial electrical DP cells are usually provided in explosion-proof housing, so these are likely to be acceptable in hazardous areas as well.

A DP cell might have a very sensitive diaphragm to measure flow or furnace draught (e.g. 20 ″WG), yet all DP cells are supplied in extremely robust steel housing, to allow for the measurement to occur at a high pressure (e.g. flow in an ammonia synthesis loop). In this case, special precautions need to be taken to ensure that the diaphragm is not inadvertently exposed to a high Δp, for example if one of the impulse lines is disconnected from the plant piping.

Figure 2.7 highlights some important considerations of DP cell installation. When it is being attempted to measure small Δp values, as in the use of an orifice plate for flow measurement, a

(a) (b)

Figure 2.6 Installed DP cells: (a) electronic and (b) pneumatic (measuring pressure).

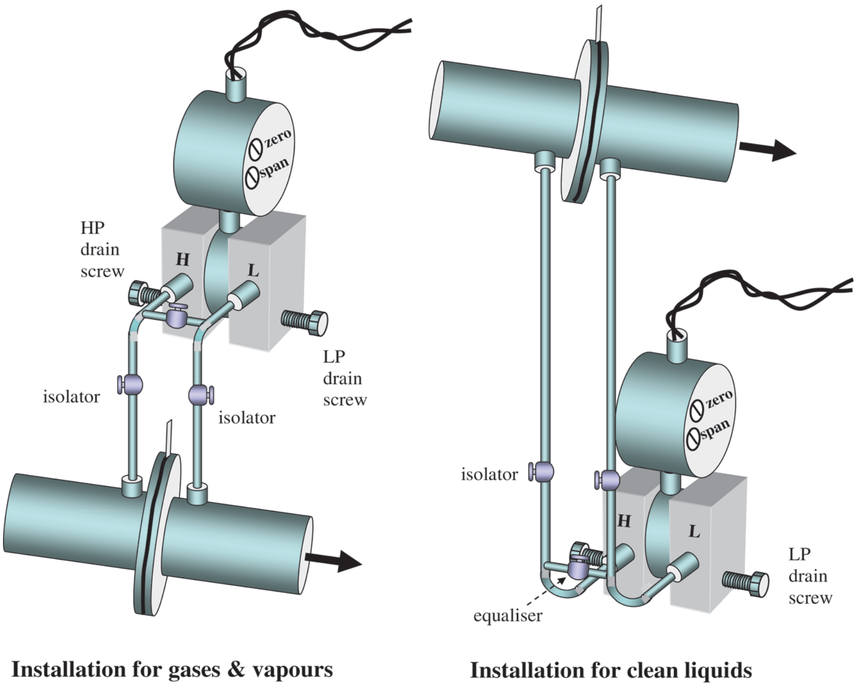

Figure 2.7 DP cell installation for flow measurement.

situation where the impulse lines themselves can exert significant and unknown pressure through vapour locks (in the case of liquids) or liquid slugs (in the case of gases and vapours) must be avoided. After all, the cell is seeking to measure the equivalent of only a few inches of water. Even a horizontal connection, if it undulates, becomes subject to the *sum* of all slugs trapped in the troughs. Thus, for gases and vapours the lines must be arranged to allow clear drainage of any condensate back into the duct, whilst for liquids one seeks a continuity of liquid all the way back into the liquid flow duct. Assuming the liquid duct is above atmospheric pressure, at start-up one loosens the drain screws on the two cell chambers, to allow the liquid to run freely through each chamber and displace any air present. Such an operation of course must take due regard of any excessive Δp that might arise across the diaphragm.

An equalising valve is normally installed between the HP and LP ports of the cell. This is the means of setting a zero Δp for the purpose of zeroing the cell. With the equalising valve open, the zeroing adjustment is turned until the cell transmits at the threshold of its signal range, namely 4 mA, 3 psig or 20 kPag. Similarly, the span adjustment can be used to effectively set the top end of the scale, for example by matching correctly to a known applied Δp.

Electronic DP cells may use potentiometric, capacitance, piezoelectric, strain gauge, silicon resonance or differential transformer techniques for interpretation of the diaphragm deflection. In the common 4–20 mA type, the device acts as a variable resistance in the current loop, and actually draws its own power from the same loop.

The principle of a pneumatic DP cell is illustrated in Figure 2.8. The diaphragm deflection varies the distance of a flapper from a nozzle, causing a variable back-pressure. This signal cannot be used directly, as additional movement of air out of the nozzle cavity will vary the calibration. Instead, a

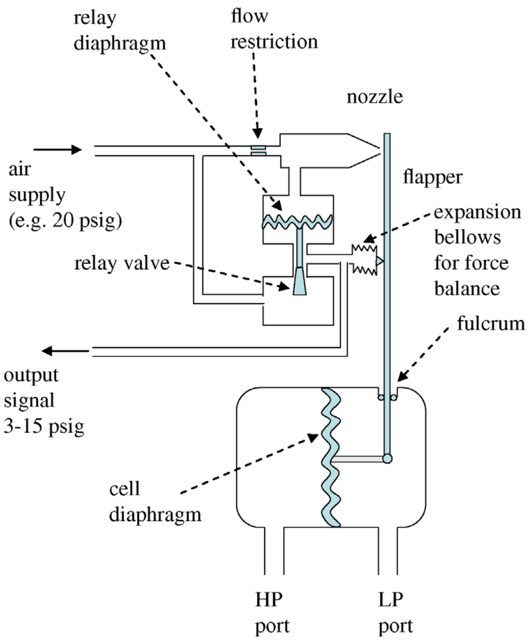

Figure 2.8 Principle of a pneumatic DP cell.

relay is used to create a balancing pressure by supplying air to an expansion bellows acting on the same flapper. The latter source is able to supply a greater flow of air at a pressure proportional to the nozzle pressure.

2.4 Measurement Instrumentation

An initial division of measurement instruments is into the categories 'local' and 'remote'. A local device needs to be read at its point of installation, and has no means of signal transmission. Typical candidates are thermodials and Bourdon-type pressure gauges. Occasionally, manometers might be used to indicate differential pressures. Up until the 1980s, local measurements were prolific, and important, with operators patrolling the whole plant at hourly intervals, jotting down the readings on a clipboard. Increasingly now the operational picture is built up entirely from electronic information, whether it is updated on graphical mimic diagrams of the plant, or archived for future analysis. It seems that the additional investment in signal transduction, marshalling, conversion and capture is worthwhile in comparison with manual reading, transcription and data entry.

In the processing industries, there are few measurements requiring the high speeds of response often needed in electrical or mechanical systems. Nevertheless, it is important to bear in mind the impact of the response time constant of an instrument considered for each application. For example, a flue gas O_2 measurement might output a smoothed version of the actual composition. A brief O_2 deficiency might not be seen, but could be enough to start combustion when the uncombusted vapours meet O_2 elsewhere in the ducting.

2.4.1 Flow Measurement

2.4.1.1 Flow Measurement Devices Employing Differential Pressure

Bernoulli's equation for incompressible fluids gives

$$\frac{\rho v^2}{2} + \rho g h + p = \text{constant} \tag{2.1}$$

which expresses the conservation of energy in the flow direction (ρ: density; v: velocity; h: height; g: gravity; p: pressure). For small pressure changes, this is also an adequate representation of gas flows. So between two points in a flow system the changes must balance as follows:

$$\frac{\rho}{2}\Delta\left[v^2\right] + \rho g \Delta h + \Delta p = 0 \tag{2.2}$$

For systems in which some of the energy is used up in overcoming frictional resistance,

$$\frac{\rho}{2}\Delta\left[v^2\right] + \rho g \Delta h + \Delta p = \Delta p_{\mathrm{f}} \tag{2.3}$$

Measurements under turbulent conditions show that Δp_{f} is almost proportional to the square of the total flow rate, whether it be flow through ducts, restrictions, plant equipment, particulate beds or free objects moving through fluids. For example, the Darcy equation for pressure loss along a duct is

$$\Delta p_{\mathrm{f}} = \left(\frac{4f' L}{D}\right)\rho \frac{v^2}{2} \tag{2.4}$$

where L is the duct length and D is the equivalent diameter. There is a minor secondary dependence on the velocity v since the friction factor f' is generally correlated using the Reynolds number $N_{\mathrm{Re}} = \rho v D/\mu$. However, as an approximation, frictional pressure losses are usually thought of in terms of a number of kinetic 'velocity heads' (one head $= v^2/2g$). For example, the disorderly expansion of a flow with sudden enlargement of a duct incurs a loss of approximately one velocity head, that is $\Delta p_{\mathrm{f}} = \rho g(v^2/2g)$.

Venturi Meter

As noted above, the flow rate in a duct, or stream velocity passing an object, can be estimated from the pressure drop occurring along a duct section or across the object. Usually these features will cause a net loss of pressure due to friction. However, the venturi flow meter is an attempt to obtain a measurable pressure difference, yet minimise the overall frictional loss. This becomes important in low-pressure systems such as vacuum distillation pipework, furnace draught or flue gas ducting, where the friction loss might compete with the available delivery pressure, alter the density or affect the vapour–liquid equilibrium significantly. It is common to see tapered venturi duct sections in place on the suctions of forced-draught furnace fans, for the purpose of combustion air flow measurement.

Assuming that the velocity profiles at positions 1 and 2 in Figure 2.9 are uniform, and that the flow is frictionless, a general equation for the total mass flow of incompressible and compressible fluids through a venturi tube is

$$w = \rho_2 A_2 v_2 = C_{\mathrm{d}} \rho_2 A_2 \sqrt{\frac{-2\int_1^2 (1/\rho)\mathrm{d}p}{1 - \left(\dfrac{\rho_2 A_2}{\rho_1 A_1}\right)^2}} \tag{2.5}$$

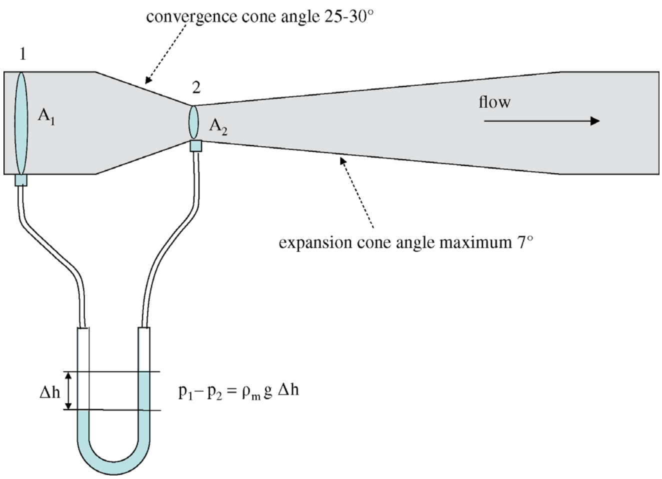

Figure 2.9 Venturi flow meter.

For an ideal gas under adiabatic conditions, this gives

$$w = C_d \rho_2 A_2 \sqrt{\frac{2\left(\frac{p_1}{\rho_1}\right)\left(\frac{\gamma}{\gamma-1}\right)\left[1-\left(\frac{p_2}{p_1}\right)^{(\gamma-1)/\gamma}\right]}{1-\left(\frac{A_2}{A_1}\right)^2\left(\frac{p_2}{p_1}\right)^{2/\gamma}}} \tag{2.6}$$

where $\gamma = c_p/c_v$ and the pressures are absolute. For p_2/p_1 close to 1, this reduces to

$$w = C_d A_2 \sqrt{\frac{2\rho_1(p_1 - p_2)}{1-\left(\frac{A_2}{A_1}\right)^2}} \tag{2.7}$$

which is also valid for an incompressible fluid. For well-designed venturi meters, the discharge coefficient C_d is found to be around 0.98 for both compressible and incompressible fluids.

Orifice Plate

The orifice plate is the most common flow measurement device employed in the processing industries. There are variations in the installation (e.g. flange tappings versus corner tappings versus upstream/downstream tappings, bevelled edge versus square edge, upstream/downstream free run, etc.), some complying with strict standards.

The principle is the same as for the venturi meter, except that here the precise area of the liquid jet represented by the flow streamlines at the *vena contracta* in Figure 2.10 is not known. Using instead the area of the orifice itself, one obtains for an incompressible fluid

$$w = C_d A_O \sqrt{\frac{2\rho(p_1 - p_2)}{1-\left(\frac{A_O}{A_1}\right)^2}} \tag{2.8}$$

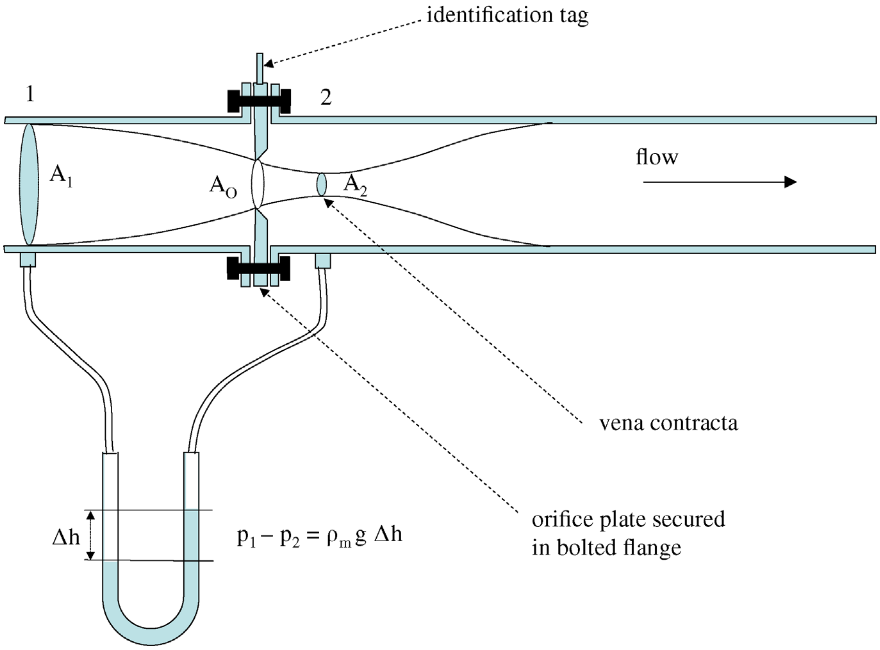

Figure 2.10 Orifice plate flow meter.

but now the discharge coefficient is around 0.61 (for high Reynolds numbers), the difference from its value in Equation 2.7 largely compensating for the deviation between A_2 and A_O.

Equation 2.6 suggests that adjustment of Equation 2.8 for compressible flow will depend primarily on γ and the ratios A_O/A_1 and p_2/p_1. Indeed, data are available correlating a compressibility factor Y in terms of γ, d_O/d_1 and $\Delta p/p_1$ $(= 1 - p_2/p_1)$.

For a particular fixed installation, users generally rely on the square-root behaviour

$$w = k\sqrt{\rho_1(p_1 - p_2)} \tag{2.9}$$

where it is seen that

$$k = YC_d A_O A_1 \sqrt{\frac{2}{A_1^2 - A_O^2}} \tag{2.10}$$

where the compressibility factor $Y \rightarrow 1$ for an incompressible fluid.

Calibration of Orifice Plate, Venturi and Similar Flow Meters

In practice, k in Equation 2.9 might be estimated using measured plant displacements. Note that the total volumetric flow rate is obtained as

$$F = k\sqrt{\frac{(p_1 - p_2)}{\rho}} \tag{2.11}$$

with the assumption that $\rho_1 \approx \rho_2$, and it is sensible to calibrate *in situ* to determine the approximately constant k. Indeed, most installations have no compensation for variations in density ρ, with

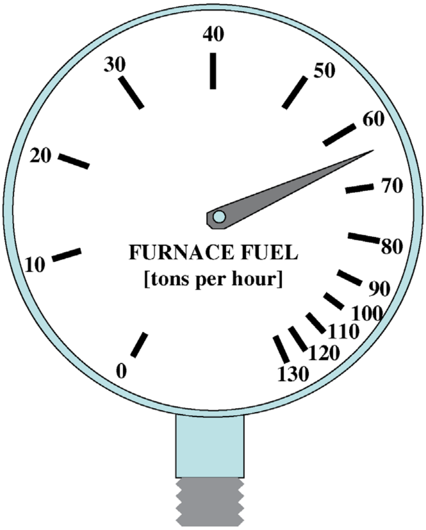

Figure 2.11 Square-root extracting flow gauge.

a fixed multiplying constant representing $k\sqrt{\rho}$ determined at plant *design conditions* applied to $\sqrt{\Delta p}$ for the form in Equation 2.9.

$$w = K\sqrt{\Delta p} \tag{2.12}$$

Prior to the widespread use of digital computers, the signal for Δp would often arrive at a display panel to be shown on the dial of a simple pressure gauge as in Figure 2.11. The graduations on the dial were simply arranged to extract the square root of the signal, and apply the multiplier K.

Engineers need to be very wary of this type of mass flow or volumetric flow indication based on a fixed K value. The density of streams on a plant will of course vary with composition, as well as operating pressure and temperature in the case of gases and vapours. The indicated flows must be corrected relative to the calibration condition as in Equations 2.13–2.14 for mass w and volumetric F flows respectively.

$$w_{\text{ACTUAL}} = w_{\text{INDICATED}}\sqrt{\frac{\rho_{\text{ACTUAL}}}{\rho_{\text{CALIBRATION}}}} \tag{2.13}$$

$$F_{\text{ACTUAL}} = F_{\text{INDICATED}}\sqrt{\frac{\rho_{\text{CALIBRATION}}}{\rho_{\text{ACTUAL}}}} \tag{2.14}$$

It follows for ideal gases that

$$w_{\text{ACTUAL}} = w_{\text{INDICATED}}\sqrt{\frac{M_{\text{ACTUAL}}}{M_{\text{CALIBRATION}}}\cdot\frac{P_{\text{ACTUAL}}}{P_{\text{CALIBRATION}}}\cdot\frac{T_{\text{CALIBRATION}}}{T_{\text{ACTUAL}}}} \tag{2.15}$$

$$F_{\text{ACTUAL}} = F_{\text{INDICATED}}\sqrt{\frac{M_{\text{CALIBRATION}}}{M_{\text{ACTUAL}}}\cdot\frac{P_{\text{CALIBRATION}}}{P_{\text{ACTUAL}}}\cdot\frac{T_{\text{ACTUAL}}}{T_{\text{CALIBRATION}}}} \tag{2.16}$$

where M is molecular mass, P is absolute pressure and T is absolute temperature.

A properly calibrated and installed orifice flow meter working at its design conditions can be expected to be accurate to within 2–4% of full span, whilst the accuracy of a venturi meter is around 1% of full span.

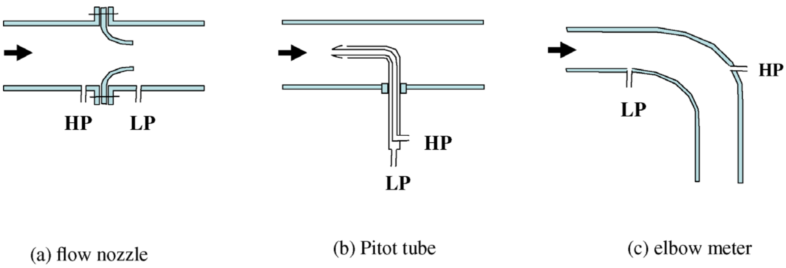

Figure 2.12 Some other flow measurement devices based on differential pressure.

Other Flow Measurement Devices Employing Differential Pressure

Figure 2.12 shows three additional flow measurement devices relying on differential pressure. The flow nozzle (a) can be expected to comply with the general equations for venturi or orifice plate meters, and has an accuracy of 2% of span. The Pitot tube (b) has a dynamic port on the tip at which the local pressure will rise by $\rho v^2/2$ according to Bernoulli's equation (Equation 2.3), as a result of the velocity being reduced to zero. The static ports just behind the tip provide the offset which allows the local v to be resolved. The device can be simplified using a separate tapping on the pipe wall for the static measurement. The elbow meter (c) is based on the same principle as the Pitot tube, and achieves an accuracy of 5–10% of full span.

Flow Control Loop Signals

It is useful at this stage to review a detailed piping and instrumentation diagram of a flow control loop, in order to identify the various elements and interconnections arising in the configuration. In Figure 2.13, where only analogue signals are shown, a measurement device requiring square-root extraction is considered, as above. Depending on the audience for an instrument and control

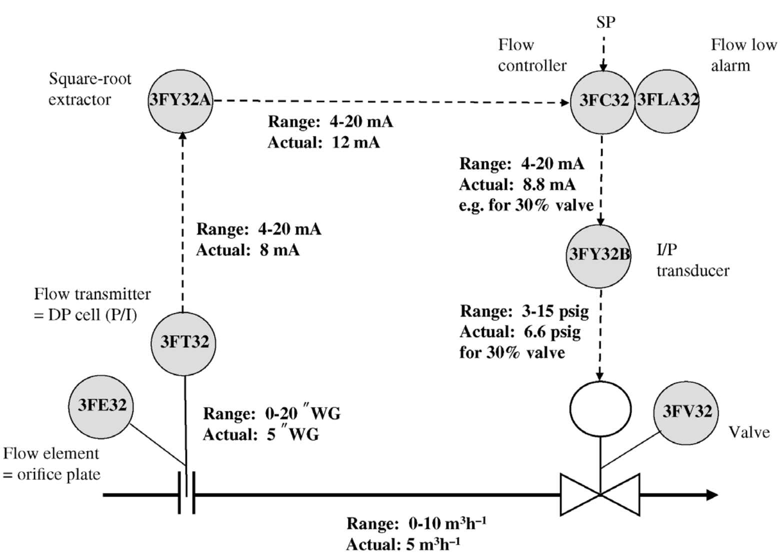

Figure 2.13 Analogue signals around a flow control loop.

scheme representation of this nature, one includes more or less detail. For example, Figure 2.13 might be shown with just a single '3FC32' bubble, and it would generally be understood that the various conversions and the square-root extraction are occurring in this simplified loop. In modern control systems, there are many optional settings for such instrumentation (such as ranges and controller parameters), and this information is usually held in a database, amounting to a number of pages for one loop.

It is interesting to follow the sequence of conversions around the control loop in Figure 2.13 for an operating flow of $5\,\mathrm{m^3\,h^{-1}}$:

$$\begin{aligned}
&\text{From orifice plate:} && \left(\frac{5-0}{10-0}\right)^2(20-0) && = 5''\mathrm{WG}\\
&\text{From DP cell:} && \left(\frac{5-0}{20-0}\right)^2(20-4)+4 && = 8\,\mathrm{mA}\\
&\text{From square-root extractor:} && \left(\frac{8-4}{20-4}\right)^{1/2}(20-4)+4 && = 12\,\mathrm{mA}\\
&\text{From controller (e.g. for 30\% valve):} && 0.3\times(20-4)+4 && = 8.8\,\mathrm{mA}\\
&\text{From I/P converter (for 30\% valve):} && \left(\frac{8.8-4}{20-4}\right)(15-3)+3 && = 6.6\,\mathrm{psig}
\end{aligned}$$

2.4.1.2 Other Flow Measurement Devices

As with most industrial instrumentation, there is a plethora of devices based on alternative technologies for flow measurement. The diversity is driven along by manufacturers' attempts to develop new niches based on patents. Indeed, some are well suited to particular environments such as dirty fluids and high-accuracy measurement. Some of these devices are shown in Figure 2.14.

Gas mass flow (thermal) devices (g) work on the principle of adding heat to the stream, and detecting its temperature change. In constant-current I mode, the flow rate is inferred from ΔT, and in constant ΔT mode from I. *Coriolis* meters (h) provide both a flow rate and density. The flow passes through a U-shaped tube which is subjected to a lateral vibration by electromagnets. The flow momentum resists the lateral movement as it enters, and increases that of the other arm of the U as it leaves – causing a phase lag between the two arms which is related to the mass velocity. In turn, the vibration amplitude depends on the stream density.

In Table 2.1, Swearingen (1999, 2001) compares the attributes of selected flow measurement devices. Here 'variable area' refers to rotameter-like devices in which an object is displaced (e.g. against gravity or a spring) in order to increase the flow area.

2.4.2 Level Measurement

2.4.2.1 Level Measurement by Differential Pressure

Differential pressure is often used to determine level. Three scenarios are shown in Figure 2.15.

Measurements of liquid level for vessels open to atmosphere are obtained from $\Delta p = \rho g h$, with atmospheric pressure acting on both the liquid surface and the LP port of the DP cell. For enclosed vessels, it is necessary to locate a second sensing point above the liquid surface, to eliminate the effect of any absolute pressure fluctuations. This will involve an extra vertical section of impulse