PROCESS MEASUREMENT & CONTROL

IN PRACTICE

Design, Specification and Implementation

SAMPLER

Edited by
John E. Edwards
Process Simulation Engineer, P & I Design Ltd

First Edition, November 2013
P&I Design Ltd
Preface

This book is based on the practical experience gained, over thirty five years, by the process, instrumentation and control engineers at P & I Design Ltd. This experience is based on design and implementation, involving a wide range of projects, in the process and control fields. This includes batch and continuous process plant, terminal storage facilities, safety systems, materials handling and secure data processing facilities installed in safe and hazardous areas. The projects required the selection and application of a wide range of instrumentation and control techniques based on standalone and centralised supervisory control systems using state of the art technologies.

The book is intended as a refresher of the basic principles for the practicing engineer and reviews the latest techniques, practices and legislation in the industry. The engineering fundamentals section has been included to provide a concise and clear introduction to the relevant applied maths, basic electrical and fluid flow principles.

Reference is made to many classic texts, industry standards and manufacturers’ data. Information has been mined from individual project reports, technical papers and contributions by specialists working in the instrumentation and control field.

Section 5 - Emergency Relief, Section 8 - Process Control, and Section 9 - Process Simulation have been adapted and developed from a previous book by J.E.Edwards, “Chemical Engineering in Practice”.

Each topic is in the form of a condensed refresher and provides useful practical information and data. Each section is numbered uniquely for contents and references, with the appendices and nomenclature being section specific. The references are not a comprehensive list and apologies for unintended omissions.

Any originality may be in the presentation format and as such some would consider this to be a Handbook or Workbook. Many modern text books do not adopt this approach which is not very helpful to the engineer challenged with providing solutions in a timely manner.

Process simulations in Section 9 are carried out using CHEMCAD™ software by Chemstations of Houston. Applications are presented from real situations whether it is design, testing or operations.

The Editor

http://uk.linkedin.com/pub/john-edwards/1b/374/924
John E.Edwards is the Process Simulation Specialist at P&I Design Ltd based in Teesside, UK.
In 1978 he formed P&I Design Ltd to provide a service in the Process and Instrumentation fields.
He has over fifty years experience gained whilst working in the process, instrumentation and control system fields.

Acknowledgements

The following personnel at P & I Design Ltd have contributed:

M.Morgan for applying his expertise in reviewing Section 3 on Liquid and Gas Hazardous Areas
D.Ransome in the preparation and review of Section 4 on Safety Instrument Systems
http://uk.linkedin.com/pub/david-ransome/14/6a2/927
P.Parkin in contributing to Section 10 Implementation
M.Edwards in reviewing all aspects associated with Information Technology
http://uk.linkedin.com/pub/mike-edwards/53/6b9/1b3

Also a special thanks to the following associates:

P.Baines of Tekna Ltd. UK in the preparation of Section10 Implementation
C.Howard for his contributions and encouragement
R.A.Mottram of MARIT Metallurgical Engineers Ltd. for Section 2.5 Materials of Construction
Section 1

Introduction

The process industry covers a broad spectrum of activities that involve the handling and treatment of gases, liquids and solids over a wide range of physical and processing conditions. The scope of this variety is shown in the following chart.

Electrical, instrument and control engineers, working in this environment, are required to provide safe, practical and timely solutions to design and operational problems frequently without having the opportunity to study the topic in great depth. This book is an attempt to provide a comprehensive review of the fundamentals, definitions and engineering principles for implementation of projects encountered in normal practice and the subsequent maintenance.

The design process should be focused on identifying potential problems and errors at the earliest stage possible in the project life cycle. Many techniques have been developed to achieve this, namely Hazard Identification (HAZID), Hazard and Operability (HAZOP), Layer of Protection Analysis (LOPA) and process simulation.

Processes can be batch, semi-continuous or continuous with the selection depending on many factors. Batch processes are used in the manufacture of a wide variety of fine and speciality chemicals and are inherently transient in nature and as such can present unique control problems.

Continuous processes are used in many industries being more prevalent in the oil and gas and petrochemicals industries. Control of continuous processes in steady state does not present any special difficulties, but during start up and process upset conditions special advanced control techniques are relied on.

To achieve consistent quality and to operate the process optimally, maximising productivity and minimising raw material and energy costs requires effective control of the key process parameters. This requires process control and data monitoring systems that matches the process to which they are being applied.

A successful outcome does not necessarily require an in depth mathematical study of the control system and its interaction with the process, which is how courses on instrumentation and control are frequently presented. Very often having a good understanding for the process behaviour and the application of common sense can provide satisfactory results.
Steady state and dynamic process simulation proves the capability to achieve stable and reproducible operating conditions with acceptable product purity, yield and cycle times to satisfy the commercial requirements and the safety and environmental issues for the regulatory authorities.

The project cost impact curves show the benefits of stress testing the designs using these techniques to avoid the increased costs of correction, even if economically possible, later in the project life cycle.

It is worth noting that many serious accidents on process plant have been as a result of failures in plant management systems and procedures and not due to failures in design. However the design and commissioning teams have the responsibility to ensure that adequate documentation and procedures are handed over to operations personnel, which should include a clear understanding of accident risks and the safety critical equipment and systems designed to control them. Maintaining a corporate memory is a serious challenge to all managements which can only be achieved by adequate systems, auditing procedures and continual training.
Section 2

Engineering Fundamentals

Contents

2.1 Maths and Units
2.2 Electrical
2.3 Fluid Flow
2.4 Process
2.5 Materials of Construction

References

Maths and Units

Fluid Flow

Process
A basic understanding of practical mathematics is frequently sufficient for an engineer working in a general design or operating process plant environment. Note, we are not considering academic study, specialist areas or research and development.

Fractions and Indices (Powers)

Two numbers, such that their product is unity, are called reciprocal numbers; for example \(\frac{4}{1}\) and \(\frac{1}{4}\). A powerful concept is to use letters in place of numbers to represent a specific parameter or variable. So we can state that \(\frac{a}{b}\) is the reciprocal of \(\frac{b}{a}\) which in our example gives \(a=4\) and \(b=1\).

If \(m\) is a positive whole number and \(a\) is any number, where \(m\) is the index (power) of \(a\), the short and convenient way of writing \(a \times a \times a \ldots\) where there are \(m\) number of \(a\)'s is \(a^m\).

The following rule \(a^m \times a^n = a^{m+n}\) always applies and it can be shown that:

For \(m = 0\), we have \(a^0 \times a^m = a^{0+n} = a^n\) making \(a^0 = 1\), note this means \(10^0 = 1\)

For \(m = 1\) and \(n = -1\), we have \(a^1 \times a^{-1} = a^{1-1} = a^0 = 1\) making \(a^{-1} = 1/a\), note this means \(10^{-1} = 1/10\)

The following can be proved using the above procedure:

\[a^2 = 1/a^2, \quad a^{\frac{1}{2}} = \sqrt{a},\text{ the square root of } a\]

Note that \((a^m)^n = a^{mn}\) always applies and it can be shown that \(x^{pq} = q\sqrt{x^p}\)

\[a^{\frac{1}{3}} = 3\sqrt{a^2}\text{ is the cubed root of } a^2\]

Number Convention SI Units – Prefixes

From the above we can deduce that fractions such as \(1/10 = 10^{-1}, 1/100 = 10^{-2}\) and so on. Numbers in the form \(10^7\) cannot be conveniently represented in computer programs, so the scientific notation is used where the \(10^7\) is now E07. Note E is not related to the mathematical constant e or the exponential function. The following conventions avoid errors in allocating the correct number of 0’s.

<table>
<thead>
<tr>
<th>Factors</th>
<th>Prefix</th>
<th>Symbol</th>
</tr>
</thead>
<tbody>
<tr>
<td>(10^{12})</td>
<td>E-12</td>
<td>pico</td>
</tr>
<tr>
<td>(10^{9})</td>
<td>E-09</td>
<td>nano</td>
</tr>
<tr>
<td>(10^{6})</td>
<td>E-06</td>
<td>micro</td>
</tr>
<tr>
<td>(10^{3})</td>
<td>E-03</td>
<td>milli</td>
</tr>
<tr>
<td>(10^{2})</td>
<td>E-02</td>
<td>centi</td>
</tr>
<tr>
<td>(10^{1})</td>
<td>E-01</td>
<td>deci</td>
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<td>(10^1)</td>
<td>E01</td>
<td>deca</td>
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<tr>
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<td>E02</td>
<td>hecto</td>
</tr>
<tr>
<td>(10^3)</td>
<td>E03</td>
<td>kilo</td>
</tr>
<tr>
<td>(10^4)</td>
<td>E06</td>
<td>mega</td>
</tr>
<tr>
<td>(10^6)</td>
<td>E09</td>
<td>giga</td>
</tr>
<tr>
<td>(10^{12})</td>
<td>E12</td>
<td>tera</td>
</tr>
</tbody>
</table>

Note 1 Tip for setting power, make equal to number 0’s so 0.00001 = \(10^{-5}\) and 100000 = \(10^5\)

Note 2 Caution refinery industry sometimes uses MM to signify \(10^6\)

Unit converters are now available from many sources so are not shown here. A typical example for energy is shown. The unit to be converted from is multiplied by the factor shown in column with the desired unit.

<table>
<thead>
<tr>
<th>from</th>
<th>Btu</th>
<th>joule</th>
<th>kWh</th>
<th>therm</th>
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</thead>
<tbody>
<tr>
<td>Btu</td>
<td>1</td>
<td>1.055E03</td>
<td>0.2931E-03</td>
<td>10E-06</td>
</tr>
<tr>
<td>joule</td>
<td>0.948E-03</td>
<td>1</td>
<td>0.2778E-06</td>
<td>9.48E-09</td>
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<tr>
<td>kWh</td>
<td>3.412E03</td>
<td>3.6E06</td>
<td>1</td>
<td>34.12E-03</td>
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<tr>
<td>therm</td>
<td>100E03</td>
<td>105.5E06</td>
<td>29.31</td>
<td>1</td>
</tr>
</tbody>
</table>

1000 Btu = \(10^3\) Btu = 1E03 Btu = 1.055E06 joule = 0.2931 kWh = 10E-03 therm
Formulae

The concept of using letters in place of numbers provides a powerful technique to define specific relationships and to perform units conversion. This approach allows the development, in many fields, of formulae to provide a statement in general terms of a whole series of particular facts or physical relationships.

Velocity, acceleration and distance

Let \( t \) represent time in seconds (s).
The reciprocal of time is called frequency \( f \) so
\[ f = \frac{1}{t} \]
with units of 1/s or in the index form \( s^{-1} \).

If a stirrer in a vessel rotates once in half a second it will rotate 2 times per second or 2/s or 2\( s^{-1} \).
\( t / 60 \) represents time in minutes obtained from units equation (s x m/60s), note the s cancels out.
\( t / 3600 \) represents time in hours obtained from units equation (s x h/3600s).

Consider an object uniformly increasing its speed (ft/s) from \( u \) to \( v \) in \( t \) seconds.
\[ \frac{(v-u)}{t} \]
represents the change in speed over \( t \) seconds the acceleration \( a \) (ft/s \( \times \) 1/s) giving units ft/s\(^2\) or ft-s\(^{-2}\).

This leads to the formula
\[ a = \frac{(v-u)}{t} \text{ ft/s}^2 \text{ or ft-s}^{-2} \]

The distance travelled in time \( t \) is given by
\[ s = ut + \frac{1}{2}at^2 \ft \]

Temperature Conversion

To convert from °F to °C we have:
\[ °F = \frac{9}{5}°C + 32 \]

A temperature of -10°C is equivalent to
\[ °F = 1.8 \times -10 + 32 = 14 \]

Pressure Conversion

Absolute pressure \( p_a \) is the pressure above a total vacuum and gauge pressure \( p_g \) is the pressure above or below atmospheric pressure \( p_{atm} \) giving:
\[ p_a = p_g + p_{atm} \text{ for } p_g > p_{atm} \]
\[ p_a = p_g - p_{atm} \text{ for } p_g < p_{atm} \]

At atmospheric pressure \( p_{atm}(\text{kg/m}^2) \), the absolute pressure \( p_a(\text{kg/m}^2) \) at the bottom of a column of liquid with a density \( \rho(\text{kg/m}^3) \) and height \( H \) (m) is:
\[ p_a = H\rho + p_{atm} \]

Conversion of \( p_a \) from kg/m\(^2\) to bar.
We know
\[ 1 \text{ kg/cm}^2 = 0.98065 \text{ bar and } 1 \text{ cm}^2 = 0.0001 \text{ m}^2 \text{ so } 1/10^4 = 10^4 \text{ kg/m}^2 = 0.98065 \text{ bar} \]

1.0197 \times 10^4 \text{ kg/m}^2 = 1 \text{ bar leading to } p = H\rho / (1.0197 \times 10^4) \text{ bar.}

\( p_{atm} = 1.01325 \text{ bar, so a } p_g \text{ of 5 barg is equivalent to } p_a \text{ of 6.01325 bar and a vacuum } p_g \text{ of 0.5 barg is equivalent to } p_a \text{ of 0.51325 bar.} \)

Power to pump liquids

The following example demonstrates the rigorous application of unit conversions and the use of a formula to calculate the power to pump a liquid. The theoretical power to pump \( Q \text{ (litres/min) of liquid, with density } \rho(\text{kg/m}^3) \) against a differential head \( h \) (m) or \( \Delta p \) (bar) is given by:
\[ P = \frac{Qh\rho}{6116 \times 10^3} = \frac{Q\Delta p}{600} \text{ kW} \]

Density of water \( \rho_{H2O} = 1 \text{ gm/cm}^3 \),
We need to convert to kg/m\(^3\) we know 1000 gm = 1 kg and 1 cm\(^3\) = 10\(^{-6}\) m\(^3\)
\( \rho_{H2O} = 1 \times 1/1000 \times 1/10^{-6} = 1000 \text{ kg/m}^3 \)
check units RHS = gm/cm\(^3\) x kg/gm x cm\(^3\)/m\(^3\) = kg/m\(^3\) = LHS

Power to pump 1000 litres/min of water 50 metres equals
\[ 1000 \times 50 \times 1000/(6116 \times 10^3) = 8.167 \text{ kW} \]

The overall efficiency \( \eta_o \), a fraction in range 0 – 1, is derived from the drive efficiency \( \eta_d \) the transmission efficiency \( \eta_t \) and the pumping efficiency \( \eta_p \) from \( \eta_o = \eta_d \times \eta_t \times \eta_p \).

The volumetric efficiency \( \eta_v = \text{actual pump flow } Q_{act} / \text{theoretical pump flow } Q_t \)
2.2 Electrical

Circuit diagrams prepared using
http://www.physicsbox.com/indexsolveelec2en.html
Ladder Logic Diagrams prepared using Wade Instruments EZ Schematics
http://www.wadeinstruments.com/index.htm

Units

The basic unit of power is the Watt (W) being the rate of work of 1 joule/sec
1 calorie = 4.18 joules and 1 HP = 746 W = 550 ft lb/sec
1 kWh = 10³ x 3600 = 3.6 x 10⁶ or 3.6E06 W-s = joules

Ω

A hydro-electric station generates 50 MWh with a turbine efficiency ηₜ of 85% and a generator efficiency η₉ of 95% giving an overall station ηₒ of 80.7%.
The water head available H is 500 ft. and water density 62.5 lb/ft³

Energy output = 50E06 x 3.6E06 = 1.8E14 W·sec x 550/746 = 1.327E14 ft-lb/s
Energy input = 1.327E14/0.807 ft-lb/s = 1.645E14

Energy available per ft³ water = 500 ft x 62.5 lb/ft³ = 3.12E04 lb/ft³
Water flow rate = 1.645E14/3.12E04 = 5.274E09 ft³/s

Units equation ft-lb/s x 1/(lb/ft²) = ft³/s

Direct Current Circuits

In a direct current (DC) circuit the power P Watts (W) transferred between a source voltage V and a load of resistance R ohms (Ω) is given by:

\[ P = V \times I \]

Where

I \quad \text{current flowing in amperes}

Ohms Law states:

\[ I = \frac{V}{R} \quad \text{leading to} \quad P = I^2R \]

When current is taken from a source voltage E the terminal voltage falls due to internal resistance r leading to the following adjustment for voltage V applied to a load R.

\[ I = \frac{E}{r+R} \quad \quad V = IR = \frac{ER}{R+r} \quad \quad V = E - Ir \]

Kirchhoff’s Laws apply in a closed circuit which state:

The sum of currents meeting at a point equals zero. \[ \sum I = 0 \]

The sum of voltage drops around the circuit is zero \[ \sum V = 0 \]

For resistances in series we have:

\[ R = R₁ + R₂ + R₃ \]

For resistances in parallel we have:

\[ \frac{1}{R} = \frac{1}{R₁} + \frac{1}{R₂} + \frac{1}{R₃} \]
2.3 Fluid Flow

Process and manufacturing facilities require extensive piping systems to distribute services and process fluids at acceptable flows and pressures for satisfactory equipment operation.

Fluid flow through a pipe, valve, fitting or nozzle will result in a pressure drop in the direction of the flow. Fluid mechanics enables the calculation of the pressure drop through a pipe, valve or nozzle as a function of the flowrate. Process equipment such as compressors or pumps, use experimentally determined performance curves to predict the flow pressure drop relationships.

The fundamental relationship for incompressible fluid flow in a circular pipe is known as the continuity equation and is given by:

\[ q = a_1 v_1 = a_2 v_2 \quad \rightarrow \quad D_1^2 v_1 = D_2^2 v_2 \quad \rightarrow \quad \frac{D_2^2}{D_1^2} = \frac{v_2}{v_1} = \beta \]

The Conservation of Energy equation due to Bernoulli, assuming no work, is given by:

\[ \frac{p_1}{w} + \frac{v_1^2}{2g} + z_1 = \frac{p_2}{w} + \frac{v_2^2}{2g} + z_2 + \sum h_L \]

Where in consistent units:

- \( q \): volume flowrate
- \( a \): pipe cross sectional area
- \( v \): fluid velocity
- \( D \): pipe internal diameter
- \( \beta \): pipe diameter ratio
- \( p \): fluid pressure
- \( w \): fluid density
- \( g \): gravitational constant
- \( z \): elevation
- \( h_L \): head loss due to friction

Subscript 1 refers to conditions at inlet and is treated as the smaller diameter
Subscript 2 refers to conditions at outlet and is treated as the larger diameter

A dimensionless number called the Reynolds Number (\( Re \)) is used to predict the fluid flowing state (laminar, transitional or turbulent) which is then used in the selection of the appropriate pressure drop correlation. Using the fps system of units, for a fluid of density \( \rho \)(lb/ft\(^3\)) and viscosity \( \mu \)(lb/ft-s) flowing in a circular pipe of inside diameter \( d \)(ft) with a velocity \( v \)(ft/s) we have:

\[ Re = \frac{v d \rho}{\mu} \]

Units equation ft/s x ft x lb/ft^3 x ft-s/lb shows that \( Re \) is dimensionless

The friction factor \( f \) is introduced which is dependent on the Reynolds Number and pipe wall roughness \( \varepsilon \). The Darcy-Weisbach equation is used to calculate, in consistent units, the friction head loss \( h_L \) of fluid flowing in a pipe of length \( L \):

\[ h_L = f \frac{L v^2}{D 2g} \]

In terms of the Darcy-Weisbach or Moody friction factor the pressure loss due to friction for a fluid with density \( \rho \) is given by:

\[ \left( \frac{dP}{dL} \right)_f = \frac{f \rho v^2}{2g D} \]

Note that the friction factor \( f \) is dimensionless:

\[ f = m \frac{s^2 m}{m^2 s^2} = 1 \]
2.4 Process

This Section reviews engineering units and their application to prevent misuse in certain applications. Mass and energy balance principles and the fundamentals of steady state and dynamic simulation are also reviewed.

Molality, Molarity, Mole Fraction and Mole Ratio (1)

Molality, \( m_i \) (mol/kg) denotes the total number of moles of solute species \( i \) per 1000g of solvent (not solution).

Molarity, \( M_i \) (mol/litre, kmol/m^3) denotes the total number of moles of species \( i \) per litre of solution, sometimes referred to as Normality \( N \) which is essentially the same as mol/litre.

Mole fraction, \( x_i \) denotes the number of moles of species \( i \) divided by the total number of moles in a solution.

\[
x_i = \frac{m_i \cdot MW_i}{1000}
\]

Mole ratio \( X_i \) denotes the number of moles of species \( i \) divided by the number of moles of species \( j \). For a binary solution we have:

\[
X_i = \frac{x_i}{x_j} = \frac{x_i}{(1 - x_i)}
\]

The weight fraction \( w_i \) of a component in an aqueous solution is calculated from:

\[
w_i = \frac{M_i \cdot MW_i}{\rho \cdot So \ln 1000}
\]

Consider a reactant species \( A \) with a flowrate \( W_A \) (kg/h) and molecular weight \( M_A \) (kg/kmol) and inert species \( B \) with a flowrate \( W_B \) (kg/h) and molecular weight \( M_B \) (kg/kmol), we have:

Mole fraction component \( A \)

\[
y_A = \frac{W_A / M_A}{W_A / M_A + W_B / M_B}
\]

Mole fraction component \( B \)

\[
y_B = 1 - y_A
\]

Mean molecular weight

\[
M_{AB} = \frac{W_A / M_A + W_B / M_B}{W_A / M_A + W_B / M_B}
\]

For a system pressure \( P=1.01325 \) bar and temperature \( T \) °K the inlet gas density (g/litre or kg/m^3) is given by:

\[
\rho_1 = \frac{M_{AB}}{22.4} \left( \frac{273}{273 + T} \right)
\]

The gas inlet concentration of component \( A \) can be converted from mole fraction to mg/m^3 using:

\[
y_A \times 10^6 \left[ \frac{M_A}{22.4} \left( \frac{273}{273 \times T} \right) \right]
\]

Weight fraction is converted to mole fraction as follows:

\[
x_i = \frac{w_i / MW_i}{w_i / MW_i + (1 - w_i) / 18}
\]
2.5 Materials of Construction

This Section reviews the selection of materials of construction as applied to the multi-purpose fine chemical and specialty chemical process sectors. The main difficulty with any general guide is that corrosion is frequently attributable to minor traces of corrosive materials in an environment and while general notes give a broad guide it cannot be assumed that all other environments are acceptable. Consideration of the precise conditions in each area of application should be made before a decision to use a material is accepted.

Assessments of plant environments on the basis of a yes/no approach should be approached with caution due to the wide range of process materials. The suitability of the material in the environment is dependent on factors such as composition, temperature, phase and duty. If the materials are assessed just on their suitability or not to a specific chemical, you will almost always finish up with only PTFE being acceptable, even though other materials might have performed satisfactorily bearing in mind that PTFE could be limited in other ways, e.g. temperature.

The table below shows chemical resistance of some materials of construction to process fluids commonly encountered and is issued for guidance only. Selection of material is the user’s responsibility.

<table>
<thead>
<tr>
<th>Process Material</th>
<th>Carbon Steel</th>
<th>St St</th>
<th>Glass Lined Steel</th>
<th>PTFE</th>
<th>PVDF</th>
<th>Halar</th>
<th>ECTFE</th>
<th>GRP Derakane 470.36</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sodium Hydroxide aqueous</td>
<td>Q</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Sodium Bromide aqueous</td>
<td>Q</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Bromine liquid/gas</td>
<td>U</td>
<td>U</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>U</td>
</tr>
<tr>
<td>Sodium Bicarbonate, aqueous</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Methanol liquid/ vapour</td>
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<td>S</td>
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<td>S</td>
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<tr>
<td>Sodium Nitrite aqueous</td>
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<td>Hexane liquid/vapour</td>
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<td>Hydrochloric Acid liquid/vapour</td>
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<td>Acetone liquid/vapour</td>
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<td>Hydrobromic Acid liquid/vapour</td>
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</tr>
<tr>
<td>Sulphur Dioxide gas</td>
<td>U</td>
<td>U</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Sodium Sulphite, aqueous</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Isopropyl Acetate liquid/vapour</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>ND</td>
<td>ND</td>
<td>ND</td>
<td>ND</td>
</tr>
<tr>
<td>Isopropyl Alcohol</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
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<tr>
<td>Sodium Hydrosulphite aqueous</td>
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<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
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</tr>
<tr>
<td>Sulphuric Acid 98% liquid/vapour</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>U</td>
</tr>
<tr>
<td>Thionyl Chloride liquid/vapour</td>
<td>Q</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>U</td>
</tr>
<tr>
<td>Monomethylamine40%liq/vapor</td>
<td>Q</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>Q</td>
<td>Q</td>
<td>Q</td>
<td>Q</td>
</tr>
<tr>
<td>Sodium Metabisulphite aqueous</td>
<td>Q</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Hydrogen Cyanide</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Potassium Hydroxide liquid</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
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<tr>
<td>NOx Gas</td>
<td>U</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>ND</td>
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<tr>
<td>Sodium Carbonate aqueous</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Carbon Dioxide gas</td>
<td>U</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Acetic Acid, liquid/vapour</td>
<td>U</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Acetic Anhydride, liquid/vapour</td>
<td>U</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>Q</td>
</tr>
<tr>
<td>Nitric Acid, liquid/vapour</td>
<td>U</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>U</td>
</tr>
<tr>
<td>Water</td>
<td>U</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Piperazine aqueous</td>
<td>ND</td>
<td>ND</td>
<td>ND</td>
<td>ND</td>
<td>U</td>
<td>ND</td>
<td>ND</td>
<td>ND</td>
</tr>
<tr>
<td>Ammonia gas/aqueous</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
</tr>
<tr>
<td>Ethyl Acetate liquid/vapour</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>Q</td>
</tr>
<tr>
<td>Dichloro Methane liquid/vapour</td>
<td>U</td>
<td>Q</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>S</td>
<td>U</td>
</tr>
</tbody>
</table>

U Unsuitable    S Satisfactory    Q Questionable    ND No data
Section 3

Liquid & Gas Hazardous Areas

Contents

3.1 Background
3.2 Principle Hazards
3.3 Hazardous Area Classification
3.4 Equipment Classification

Appendices

I Safety Case Assessment
II Gas Group Classification Data
III Hazardous Area Classification Examples

References

5. BS EN 60079-10-1:2009, "Classification of areas - Explosive gas atmospheres"
3.1 Background

This Section covers the topic of hazardous areas processing liquids, vapours and gases as applied to installations in the EU and North America. It is intended for introductory guidance but for further information reference should be made to the appropriate legislation for the country under consideration. However the basic principles discussed are still valid.

From 1985 the EU started a process for Technical Harmonisation and Standards to achieve free movement of goods within Europe. This has led to major legislative changes and introduced the potential for unintended non-compliance if the new directives are not fully understood and applied. This confusion was unfortunate as the basic principles involved remained unchanged and the existing systems in place were well understood by engineers in the field.

Most of the electrical standards have been developed over many years and are now set at international level, while standards for non-electrical equipment are only just becoming available from CEN Comité Européen de Normalisation (European Committee for Standardization) together with CENELEC European Committee for Electrotechnical Standardization.

The ATEX 95 Directive is derived from the 94/9/EC EU directive “Appareils destinés à être utilisés en ATmosphères EXplosives” (Equipment destined to be used in Explosive Atmospheres) and relates to the manufacture of equipment.

ATEX 137 is derived from the 99/92/EC workplace EU directive setting minimum requirements for protection of people.

In the UK Statutory Instrument 2776 Health and Safety, The Dangerous Substances and Explosive Atmospheres Regulations 2002(DSEAR) sets the minimum legal requirements for the protection of workers from fire and explosion risks related to dangerous substances and potentially explosive atmospheres. All new and existing plants must now be compliant.

Other sources of advice, which describe more sophisticated approaches, are the Institute of Petroleum Model Code of Practice (Area Classification Code for Petroleum Installations, 2002), and the Institution of Gas Engineers Safety Recommendations SR25, (2001). The IP code is for use by refinery and petrochemical type operations. The IGE code addresses specifically transmission, distribution and storage facilities for natural gas, rather than gas utilisation plant, but some of the information will be relevant to larger scale users.
3.2 Principle Hazards

Explosive gases, vapors and dusts have different properties that affect the likelihood and severity of an explosion. These properties include flame temperature, minimum ignition energy, upper and lower explosive limits, and molecular weight. Additional parameters include maximum experimental safe gap, minimum ignition current, explosion pressure, time to peak pressure, spontaneous ignition temperature, and maximum rate of pressure rise. Substances are ranked into groups with similar properties to simplify the specification of equipment for hazardous areas.

Flammability of combustible liquids is defined by their flash-point which is the temperature at which the material will generate sufficient quantity of vapour to form an ignitable mixture. The flash point determines if an area needs to be classified. A material may have a relatively low auto ignition temperature yet if its flash-point is above the ambient temperature the area may not need to be classified. Conversely if the same material is heated and handled above its flash-point, the area must be classified. Properties of common solvents are shown:

<table>
<thead>
<tr>
<th>Solvent</th>
<th>Lower Flash Point °C</th>
<th>Auto ignition °C</th>
</tr>
</thead>
<tbody>
<tr>
<td>Toluene</td>
<td>4</td>
<td>480</td>
</tr>
<tr>
<td>Heptane</td>
<td>-4</td>
<td>223</td>
</tr>
<tr>
<td>Octane</td>
<td>13</td>
<td>220</td>
</tr>
<tr>
<td>Benzene</td>
<td>-11</td>
<td>562</td>
</tr>
<tr>
<td>n-Hexane</td>
<td>-22</td>
<td>225</td>
</tr>
<tr>
<td>Dichloromethane</td>
<td>13</td>
<td>413</td>
</tr>
<tr>
<td>Carbon disulfide</td>
<td>-30</td>
<td>102</td>
</tr>
</tbody>
</table>

Gases or vapours used in industry are classified into gas groups with subdivisions depending on their properties. North American NEC 500 Class I for gases and vapours is equivalent to European Group II for surface installations. Mining applications Group 1 or NEC Class II are not considered.

<table>
<thead>
<tr>
<th>Gases and Vapours Designation of Gas Groupings</th>
</tr>
</thead>
<tbody>
<tr>
<td>Typical Gas</td>
</tr>
<tr>
<td>Division Gas Groups</td>
</tr>
<tr>
<td>NEC 500 Class I</td>
</tr>
<tr>
<td>Zone Gas Groups</td>
</tr>
<tr>
<td>Group II CENELEC EN5014</td>
</tr>
<tr>
<td>Methane, Propane</td>
</tr>
<tr>
<td>D</td>
</tr>
<tr>
<td>IIA</td>
</tr>
<tr>
<td>Ethylene, coke oven gas</td>
</tr>
<tr>
<td>C</td>
</tr>
<tr>
<td>IIB</td>
</tr>
<tr>
<td>Hydrogen</td>
</tr>
<tr>
<td>B</td>
</tr>
<tr>
<td>IIC</td>
</tr>
<tr>
<td>Acetylene</td>
</tr>
<tr>
<td>A</td>
</tr>
</tbody>
</table>

The gas groups are formed in order of how volatile the gas or vapor would be if it was ignited, IIC being the most volatile and IIA being the least. The groups also indicate how much energy is required to ignite the gas by spark ignition, Group IIA requiring the most energy and IIC the least. Gas group IIB is suitable for group IIA and gas group IIC is suitable for both groups IIA and IIB. A group designation II with no A, B, or C is suitable for any gas group.

The term flammability is used to predict the ability of fuel-air mixtures to sustaining combustion in the presence of an ignition source. Flammability limits define the proportion of flammable substances in a mixture, for it to become flammable. Gas mixtures consisting of combustible, oxidizing, and inert gases are only flammable under certain conditions.

An explosive atmosphere is a mixture with air, under atmospheric conditions, of flammable gases, vapours, mists or dusts in which, after ignition has occurred, combustion spreads to the entire unburned mixture. A potentially explosive atmosphere is an atmosphere which could become explosive due to local and operational conditions. An atmosphere devoid of oxygen is neither flammable nor explosive regardless of the fuel gas concentration.

The Lower Flammable Limit (LFL) describes the leanest mixture that sustains combustion while the Upper Flammable Limit (UFL) gives the richest flammable mixture. Concentrations below LFL are too lean to burn and higher than UFL are too rich to burn. These limits are sometimes referred to as explosive limits with LFL/LEL and UFL/UEL being used interchangeably. LFL/UFL is preferred since for many flammable mixtures to result in an explosion, rapid reaction causing overpressures, needs an ignition source and containment. LFL/UFL concentrations vary greatly between combustible gases. An Oxygen enriched atmosphere lowers the LFL and increases the UFL.
Section 4

Safety Instrument Systems

Contents

4.1 Background
4.2 Risk Assessment
4.3 Design
4.4 Demonstrating Prior Use

Appendices

I General Definitions
II HAZOP Technique
III ALARP
IV LOPA
V Calculation procedures

References

2 BS EN 61511:2004, “Functional safety —Safety instrumented systems for the process industry sector”.
4 HSE, “Health and Safety at Work Act 1974”.
5 AIChE, “Layer of Protection Analysis”, CCPS Concept, 2001
6 Royal Academy Engineering, “Common Methodologies for Risk Assessment and Management”
7 HSE, “Lines of Defence/Layers of Protection Analysis in the COMAH Context”
9 D. R. Ransome, “Guideline Demonstrating Prior Use”, Chemical and Downstream Oil Industry Forum (CDOIF)
4.1 Background

This section reviews the issues involved in applying the International Electrotechnical Commission (IEC) Standards BS EN 61508-2010\(^{(1)}\) and BS IEC 61511-2004\(^{(2)}\) to Risk Management in the Process Industry.

The UK Health and Safety Executive (HSE), Management of Health and Safety Regulations 1999\(^{(3)}\), already imposes a legal requirement for organisations to carry out risk assessments of their operations. The HSE considers good relevant practice as being in place by reference to Approved Codes of Practice (AcoPs), HSE guidance, recognised standards, and industry practice appropriate to the duty holders’ activities.

BS EN 61508 is primarily concerned with Safety Instrumented Systems (SIS) whose failure could have an impact on the safety of persons and/or the environment. The consequences of failure could have a serious business impact and the SIS can be used in mitigation against this situation. The implementation of BS EN 61508 and BS IEC 61511 involves applying the principles of quality management to safety system specification, design, installation, operation, and maintenance throughout the safety lifecycle, providing a basic framework for applying a structured approach to Risk Assessment and Management within an organisation.

The safety lifecycle process of “plan, do, and review” can significantly reduce wasteful over-design of safety systems, as well as limiting unsafe under-design by an efficient structured process. There is a life cycle duty in Section 6 of the 1974 Health and Safety at Work (HSW) Act\(^{(4)}\) for ‘any article for use at work’, which applies ‘at all times when it is being set, used cleaned or maintained’.

A Royal Academy of Engineering study\(^{(5)}\) emphasised the importance of adopting an integrated risk management approach within a culture of accepting inherent risks. A risk policy, adapted to the needs of the business, should be in place for employees, setting out the limits of their responsibilities and accountabilities. The overall risk assessment and selection of options must have regard to the intended life cycle, including construction, commissioning, operation, maintenance, foreseeable modifications and eventual decommissioning or disposal.

Risk assessment can be implemented and managed using the As Low As Reasonably Practical (ALARP)\(^{(7)}\) technique which challenges the adequacy of existing measures and considers additional identified practicable measures. The ALARP principle requires that the cost of a measure be ‘grossly disproportionate’ to the benefits before the measure can be considered not reasonably practicable to implement. ALARP is inherent in the HSW Act \(^{(4)}\), specifically “Risks to workers – Health and Safety at Work (HSW) Act S2 risks” and “Risks to people not in their employ – HSW Act S3 risks”. The ALARP triangle is shown below and is discussed further in Appendix III:

![ALARP Triangle Diagram](image-url)
There are many risk analysis methods that can be employed with the graph below indicating some of the available assessment techniques. It can be seen that there is a relationship between the complexity of the analysis and the conservatism that should be used with the result.

The safety requirements for a Safety Instrumented System (SIS) involving Electrical / Electronic / Programmable Electronic Systems (E/E/PES) are determined in a systematic risk based manner taking into consideration other technology safety related systems and external risk reduction factors. BS EN 61508 is generically based and applies to all E/E/PES safety related systems irrespective of the application.

Risk reduction is achieved utilising multiple independent protection layers. Each protection layer must be independent, auditable and effective. The Layer of Protection Analysis (LOPA) uses agreed risk tolerance criteria and forms the basis for determining if sufficient layers are available and that they provide sufficient protection in regard to risk reduction.

LOPA is a very useful tool in the determination of the Safety Integrity Level (SIL) required for SIS based upon the shortfall in the risk reduction achieved to that required. Independent Protection Layers (IPLs) are applied to minimise the SIL whilst achieving ALARP.

Business and safety risk tolerance criteria (5. App E) need to be defined and agreed before LOPA is implemented. These criteria include initiating event frequencies, LOPA credits, impact event frequencies and consequence severity.

Safety shutdown systems or procedures should not be used to rectify or mitigate a potentially hazardous situation brought about by poor plant design. A major objective in the design of any plant should be to make the plant inherently safe – as far as possible by ‘designing out’ the hazards – so reducing reliance on protective systems.

Suppliers of process plant are legally required to carry out risk assessments to demonstrate that adequate arrangements are in place for preventing a major accident and for limiting the consequences of those which do occur. Contractors are legally required to carry out documented risk assessment and risk reduction reviews, appropriate to the equipment and process plant in his supply.
Section 5

Emergency Relief Systems

Contents

5.1 Introduction
5.2 Relief Sizing Fundamentals
5.3 Definitions and Assumptions
5.4 Sizing Methods Basis
5.5 Nomenclature

Appendices

I CHEMCAD Relief Sizing Tool Design Note
II CHEMCAD Relief Sizing Tool User Notes
III General Guidance Notes
IV Rupture (Bursting) Disc Technology Definitions

References

5.1 Introduction

To achieve safe operation of chemical processes, particularly when involving exothermic reactions, requires a combination of preventative and protective measures.

Preventative measures minimise the occurrence of an event and include:-
- Automatic control systems including the use of independent hardwired alarm and trip systems based on Layer of Protection Analysis (LOPA) and Safety Integrity Level (SIL) analysis.
- Provision for appropriate manual intervention.

Protective measures mitigate the consequences of a runaway reaction and include:-
- Emergency pressure relief.
- Crash cooling.
- Reaction inhibition.
- Drown out.

This section reviews the techniques associated with the design of emergency pressure relief and blowdown systems. An emergency relief system comprises the following key components:-
- Primary pressure relief device.
- Relief system piping and headers.
- Containment system.
- Treatment system.
- Atmospheric dispersion.

Specific design techniques are required for each component of the system and process modelling of the total integrated system can demonstrate the adequacy of the total design.

A pressure relief device is a relief valve, a bursting (rupture) disc or a combination of both. On chemical reactors the bursting disc is the most favoured due to being able to handle the following conditions:-
- Rapid pressure rise with full relief area available, except when vacuum support used.
- Toxic fluids where no leakage past a safety valve is permitted.
- Corrosive fluids that may cause progressive deterioration of a safety valve.
- Fluids that may deposit solids or build ups that interfere with safety valve operation.
The major disadvantages of bursting discs are:-

- Require a larger allowance between the operating pressure and the set pressure.
- If the operating pressure and the set pressures are too close, the disc can fail prematurely due to pressure pulsations.
- Loss of containment of reactor contents on operation i.e. valve does not reseat, unless used in conjunction with a safety valve.

To mitigate against the loss of containment, due to operation of a bursting disc sized for the worst case scenario, the installation of a smaller bursting disc/relief valve combination in parallel set at a lower relief pressure can be considered. This smaller system operates in the event of nuisance pressure build ups, due to maloperation, without total loss of containment. This technique has the disadvantage, due to the poor set pressure tolerance of bursting discs typically, $\pm 10\%$, of significantly increasing the set pressure of the main bursting disc to ensure the smaller disc operates first. This may lead to an undesirable situation particularly when protecting for exothermic runaway events.

The relief pressure, at which the relief device is fully open, should be set at the lowest pressure practicable consistent with preventing nuisance operations for the following reasons:-

- For most exothermic runaway reactions, the reaction rate and heat release rate increases exponentially with temperature. For a vapour pressure system, a low relief pressure means a low relief temperature and hence a relatively low rate of heat release. The relief area required is directly proportional to the rate of heat release by the reaction.
- For a relief system venting a two-phase mixture, pressure relief acts to remove reactants from the reactor. A low relief pressure allows a greater margin between the relief pressure and maximum permitted pressure, and advantage is taken of this by the sizing methods to yield a smaller relief area.

The requirements for the design of relief systems can be summarised:-

- The equipment design pressure plus permitted accumulation is not exceeded.
- The pressure relief system is as small as possible consistent with the above clause. A small relief system minimises cost, disposal requirements and the potential rate at which material could be discharged to the environment.
Relief System Piping and Headers

The flow capacity of a bursting disc is determined by the disc diameter and the discharge coefficient. However the system capacity can be limited by the discharge piping system. Key features of the relief system piping design can be summarised:-

- Branch pipe from reactor to the header to be not less than the outlet diameter of relief device.
- Branch pipe to enter main header as a 45 degree T, flow through branch.
- All 90° bends to have maximum R/D consistent with layout constraints.
- Main header to slope towards the blowdown drum and enter tangentially.
- Consideration to be given to nitrogen inerting to prevent explosive mixtures.
- Provision of adequate inspection and test facilities to ensure headers are clear.

Containment System

In many instances the discharge stream from an emergency relief system is a two phase vapour liquid mixture. The stream is routed to a blowdown/knock-out drum designed to disentrain the liquid from the vapour to allow discharge to atmosphere or for downstream treatment. The justification for the blowdown drum is:-

- Prevents release of hot, toxic and corrosive liquid resulting in potential safety hazards and environmental damage.
- Prevents release of flammable droplets which could lead to a vapour cloud explosion.
- Allows for downstream treatment of toxic vapours in a wet scrubber, flare or incinerator.

There are many designs of blowdown drum depending upon the circumstances but key features include:-

- Tangential inlet into a vessel of sufficient diameter to effect good vapour – liquid separation.
- Total volume sufficient to hold the estimated carryover, typically two times the volume of the largest reactor connected to the relief system.
- Adequate instrumentation monitoring for level and pressure detection.
- Appropriate facilities for drainage and material handling.
- Appropriate facilities for quenching reaction mixtures.

Sizing of the blowdown drum is carried out in accordance with API 521. The basic design method\(^3\) involves the calculation of the allowable vapour velocities for the components under consideration which allows the vapour flow area to be set. A drum diameter can then be determined on the basis that the vapour flow area occupies half of the drum area. Refer Appendix IV.

The drum volume allowed for the disentrained liquid is based on the following criteria:-

- For non-foaming systems the volume should be equal to the maximum working volume of the largest reactor connected to the system.
- For foaming systems the volume should be a minimum of 1.5 times the maximum working volume of the largest reactor connected to the system.
- As a general rule in multi-purpose batch plants the minimum blowdown drum volume should be equal to 2 times the maximum working volume of the largest reactor connected to the system.
Section 6

Measurement

Contents

6.1 Flow
6.2 Level
6.3 Pressure
6.4 Temperature
6.5 Analytical

References


Introduction

Reliable and accurate measurement sensor selection is one of the key parameters in the successful operation of a process plant.

Measurement sensor technology is a key driver in the development of modern industrial processes. The technologies, to measure and transmit process parameters, such as flow, level, temperature and pressure, have developed significantly since the 1960’s.

Impulse lines, used to connect instruments to the process, are found less frequently on new installations and are being replaced on existing ones. Where used, they require specialist knowledge during design, installation and maintenance for reliable measurement. Modern techniques have simplified maintenance and enhanced plant performance.

Transmission technology development has allowed universal application of self-powered two wire 4-20 mA dc signals. This has eliminated power supply and special sensor cabling for magnetic flow meters, thermocouples and resistance bulbs, reducing cost and simplifying installation.

In the 1980’s, microprocessors facilitated the transition from signal to information based process automation. Smart transmitters provided bi-directional digital communication and diagnostics capability with HART® (Highway Addressable Remote Transducer) protocol. The 4-20 mA and HART digital signals are transmitted over the same wiring, providing a centralised capability to configure, calibrate, characterise and diagnose devices in real time together with reporting capability. Data can be captured from multi-parameter devices without additional hardware, providing predictive maintenance capability.

Development in fieldbus digital communication allows field devices to be connected using a single cable bus structure, reducing cabling, installation time and cost. Fieldbus is a device level network that sacrifices speed for security.

There are several protocols available with Modbus®, PROFIBUS PA and FOUNDATION™ being the most common. Modbus® was the earliest protocol and has connectivity with Ethernet and other fieldbus protocols. PROFIBUS PA was developed in Europe and with PROFINET has Ethernet connectivity. FOUNDATION™ fieldbus was developed in North America by ISA, suppliers and users.

Fieldbus technology is more complex and costly, requiring suppliers to provide sensor options to meet the different standards. Fieldbus selection is guided by plant layout, sensor interface capabilities and data management infrastructure. An industry working group is currently developing the capabilities of Electronic Device Description (EDD) technology. An understanding of the fundamentals of process control is essential before beginning work on dynamic simulation. Process Control Systems by Shinskey[1] is a landmark text in which Process Control is explained from first principles, with due consideration given to real processes in regards to mass and energy balances. The cases presented in this Section provide an introduction to dynamic simulation and simple control loop applications.
6.1 Flow

Accurate flow measurement is a key element in process productivity. Various types of flow meters are used, with the orifice plate and differential pressure transmitter being predominant prior to 1965. New materials, manufacturing techniques and micro-electronics have resulted in significant developments.

Measurement accuracy is quoted for a given turndown ratio (max/min reading) as % span (max–min reading), % full scale or % actual, with the latter being the highest specification for same value. Accuracy is important for stock and custody monitoring, with reproducibility, the ability of the sensor to reproduce its reading, being important for control.

**Differential Head Flowmeters**

The differential pressure measured and unrecovered pressure loss across a square edge concentric orifice plate is dependent on the pressure tap location; as shown in the diagrams below. It can be seen that full flow taps (2½D and 8D) measures the permanent pressure loss and should be used for restriction orifice calculations.

The fundamental relationships for orifice plates are:

For liquids:

\[
Q = K \cdot d^2 \cdot C \sqrt{\frac{h}{\rho_f}} \quad \text{and} \quad W = K \cdot d^2 \cdot C \sqrt{\frac{h}{\rho}}
\]

For gases

\[
W = K \cdot d^2 \cdot C \sqrt{\frac{h \cdot p_f \cdot M}{T_f}}
\]
**Sizing Methods**

The methods are based on “Principles and Practice of Flow Meter Engineering” by L.K.Spink, Foxboro Company, 1967. Section 9 Case 5.11 details sizing using CHEMCAD.

The equation for non-viscous liquid flow is given by:

\[
S = \frac{W_m}{N D^2 F_s F_m (G_f)} h_m
\]

Where:
- \(W_m\) maximum rate of flow (lb/h)
- \(D\) inside pipe diameter (in)
- \(F_s\) ratio of area of device bore at flowing temperature to that at 68ºF
- \(F_m\) manometer correction factor (=1 for diaphragm transmitters)
- \(N\) constant for units adjustment (N=2835 for lb/h)
- \(G_f\) specific gravity of liquid at flowing temperature, water at 60ºF=1.0
- \(h_m\) maximum differential pressure (in wg)

\[
F_s = 1 + 2\alpha (t_f - 68)
\]

Where:
- \(\alpha\) coefficient of thermal expansion orifice material (in/in ⁰F)
  Typical value for 18/8 SS is 9.5E-06 and for Monel is 7.0E-06
- \(t_f\) flowing temperature (⁰F)

The orifice resistance coefficient is given by:

\[
K_r = \frac{1 - \beta^2}{C^2 \beta^4} \\
C = \frac{C_d}{(1 - \beta^4)^{0.5}}
\]

Where:
- \(C\) orifice flow coefficient
- \(d\) orifice bore
- \(\beta\) \(d/D\) (For better measurement try and keep in the range 0.3 to 0.6)

The equation for viscous liquid flow is given by:

\[
S = \frac{W_m}{N D^2 F_s F_m F_c (G_f)} h_m
\]

The application of the viscosity correction factor \(F_c\) for plant operational measurements and control is rarely justified. Viscosity limits for 1% calculation tolerance vary in the range of 1 to 8 cps depending on the \(\beta\) ratio, keeping <0.6, and pipe size. \(F_c\) can vary in the range of 1.0 to 1.09. Refer to L.K.Spink Flow Handbook for more information.
6.2 Level

Liquid level measurement is the key to reliable and safe process plant operation. Normally flows are held steady whilst levels are allowed to change within limits, requiring reproducibility. Accuracy is important for tanks used for stock and custody control.

The hydrostatic continuous, indirect, level method measures the pressure due to liquid level and density plus over-pressure. The sensor measures the pressure difference between this pressure and a reference pressure, normally atmospheric, so is not preferred for vacuum and pressure service. Instruments can be flanged mounted or rod insertion type, the latter not being recommended for turbulent conditions. Typical accuracies claimed are ±0.2% reading, and is dependent on process fluid properties and conditions.

The displacer continuous or point level method measures the change in buoyancy via a torque tube or lever arrangement. The continuous measuring range is set by the displacer length immersed in the tank or external cage, which is preferable on noisy applications. The point method uses a float with the range being limited by the length of the float arm.

The nucleonic point or continuous, non-contact, level method measures the signal strength of a radioactive source beamed across a vessel. Independent of fluid properties and has typical ranges of 0.24 to 3.36m. Typical accuracies claimed are ± 2% reading. This is the preferred method for controlling level in flash vessels and reboilers under all temperature and pressure conditions.

The radar point or continuous level method measures the travel time of an impulse transmitted and reflected from the liquid surface. Interference echoes resulting from tank internals and agitators are suppressed and signals characterised to give liquid volume. The sensor has no contact with the liquid but is exposed to head space conditions, which do not affect the measurement. The reflectivity requires the liquid dielectric constant ($\varepsilon_R$) to be ≥1.4 (hydrocarbons 1.9 - 4.0, organic solvents 4.0 - 10 and conductive liquids >10). The antenna and signal conditions are adjusted to suit the process, with guided radar being used for low $\varepsilon_R$ and turbulent conditions. Method is suitable for custody transfer with accuracy ±0.5mm being claimed.

The capacitance point or continuous level method is suitable for liquids which can act as dielectrics. The measurement is more sensitive when the difference $\delta\varepsilon_R$ between the liquid and the vapour space or between the two liquids is higher. Special designs, involving coated and twin probes, are used when $\delta\varepsilon_R < 1.0$, conductivities $> 100$ μmho, coating effects or vessel material is non-conducting. Typical accuracies claimed are ±0.25% span and is dependent on fluid properties, so is not suitable for changing conditions. Maximum conditions 200°C at 100 bar and 400 °C at 10 bar.

The ultrasonic point or continuous level measurement is based on the time-of-flight principle. A sensor emits and detects ultrasonic pulses which are reflected from the surface of the liquid. The method is non-invasive, with some types being non-contact, and is not affected by $\varepsilon_R$, conductivity, density or humidity. Maximum conditions 150°C at 4 bar.

Load cells, based on strain gauge or piezoelectric, measure the weight of the process vessel plus contents. Individual load cell accuracy of 0.03% full scale is achievable but overall performance is dependent on correct installation practices preventing external forces, due to associated piping and equipment. For vessels with jackets, agitation and complex piping it is difficult to obtain an acceptable accuracy. When the container can be totally isolated, as in final dispensing and filling applications, precision weighing can be achieved.

The vibrating tuning fork principle is used to detect point liquid level but is unsuitable in viscous and fouling applications. Maximum conditions 280 °C at 100 bar.

The conductivity point level method requires a liquid conductivity > 0.1 μmho and is frequently used on utility and effluent pump control systems.
Level Application Summary

It is essential to understand the physical property variations of the process fluids and the phase changes that may occur within the process during normal and abnormal conditions. Boilers, flash vessels and distillation column bottoms involve boiling liquids, resulting in noisy levels.

Displacers in external cages frequently are used on steam generators and flash vessels, provided the process fluids are of low viscosity and relatively clean.

The non-contact nuleonic method will prove most reliable for distillation column bottoms, where reproducibility is more important than absolute accuracy. While expensive, it can be more than justified given its value in providing stable column operation and in preventing reboiler fouling due to loss of level.

Avoid the use of impulse lines in level systems if the process pressure varies and there’s a tendency for solids’ formation due to freezing, precipitation or polymerization. Purging the lines with inert gas or process compatible fluids will have limited success and is high maintenance. An alternative solution, which is less expensive than the nuleonic method, is to use remote diaphragm seals as shown in the diagram.

Nuleonic level detection provides a powerful tool to perform on-line process diagnostics. Typical applications include monitoring level profiles in tray towers, distribution in packed beds, locating level build-up and blockages in vessels, and general flow studies.
The following sequence gives an indication of comparative costs, with the lowest first:

Conductivity → Capacitance → Tuning fork → Hydrostatic → Displacer → Ultrasonic → Load cell → Radar → Nucleonic
6.3 Pressure

Pressure Terminology and Units

Process pressure is defined as the force applied to a surface area with typical units as follows:

<table>
<thead>
<tr>
<th>from</th>
<th>to</th>
<th>psi</th>
<th>kg/cm$^2$</th>
<th>bar</th>
<th>mm Hg</th>
<th>atm</th>
</tr>
</thead>
<tbody>
<tr>
<td>psi</td>
<td>1</td>
<td>0.07031</td>
<td>0.06895</td>
<td>51.715</td>
<td></td>
<td></td>
</tr>
<tr>
<td>kg/cm$^2$</td>
<td>14.223</td>
<td>1</td>
<td>0.9807</td>
<td>735.6</td>
<td></td>
<td></td>
</tr>
<tr>
<td>bar</td>
<td>14.504</td>
<td>1.0197</td>
<td>1</td>
<td>750.06</td>
<td></td>
<td></td>
</tr>
<tr>
<td>mm Hg</td>
<td>0.01934</td>
<td>0.00136</td>
<td>0.00133</td>
<td>1</td>
<td></td>
<td></td>
</tr>
<tr>
<td>atm</td>
<td>14.696</td>
<td>1.0332</td>
<td>1.01325</td>
<td>760</td>
<td>1</td>
<td></td>
</tr>
</tbody>
</table>

Pressure is measured in terms of absolute, gauge or vacuum pressure. The relationship between these definitions is shown in the diagram below.

Absolute pressure $p_a$ is the pressure above a total vacuum and gauge pressure $p_g$ is the pressure above or below atmospheric pressure $p_{atm}$ giving:

$$p_a = p_g + p_{atm} \text{ for } p_g > p_{atm}$$

$$p_a = p_g - p_{atm} \text{ for } p_g < p_{atm}$$

So we have the following absolute pressure relationships:

1 atm = 14.696 psi = 1.01325bar which is equivalent to 0 psig and 0 barg
30 psig = 44.696 psia = 3.082 bara

It is recommended that absolute pressures are stated as “psia” and “bara” to prevent confusion. Many literature sources quote absolute and gauge pressures interchangeably leading to confusion.

In level measurement a liquid of density $\rho$ and level $H$ gives a static head pressure of $H\rho$ and for a tank at atmospheric pressure $p_a$ the level transmitter HP connection pressure will be $(H\rho + p_a)$. If the pressure transmitter LP side is connected to the tank vapour space the transmitted signal is proportional to $H\rho$.

If the HP and LP connections are continually immersed in the liquid level the transmitted signal can be used to determine the liquid density.
6.4 Temperature

An excellent practical reference work on thermocouple and resistance thermometry can be downloaded from www.te-direct.com

Resistance Temperature Detector (RTD)
http://www.omega.com/temperature/tsc.html

RTDs provide good accuracy, excellent stability and repeatability. Platinum RTDs are very stable and are not affected by corrosion or oxidation. RTDs are relatively immune to electrical noise especially around motors, generators and other high voltage equipment. Each type of temperature sensor has a particular set of conditions for which it is best suited. RTDs offer a wide temperature range with -50 to 500°C for thin-film and -200 to 850°C for wire-wound and can be used in all but the highest-temperature industrial processes. The DIN or IEC standard is considered the world-wide standard for platinum RTDs. This standard, DIN/IEC 60751 (or simply IEC751), requires the RTD to have an electrical resistance of 100.00 Ω at 0°C and a temperature coefficient of resistance (TCR) of 0.00385 Ω/Ω/°C between 0 and 100°C.

<table>
<thead>
<tr>
<th>Temperature</th>
<th>Accuracy</th>
</tr>
</thead>
<tbody>
<tr>
<td>°C</td>
<td>±0.56</td>
</tr>
<tr>
<td>Ohms</td>
<td>±1.3</td>
</tr>
<tr>
<td>-200</td>
<td>±0.32</td>
</tr>
<tr>
<td>-100</td>
<td>±0.12</td>
</tr>
<tr>
<td>0</td>
<td>±0.30</td>
</tr>
<tr>
<td>100</td>
<td>±0.48</td>
</tr>
<tr>
<td>200</td>
<td>±0.64</td>
</tr>
<tr>
<td>300</td>
<td>±0.79</td>
</tr>
<tr>
<td>400</td>
<td>±0.93</td>
</tr>
<tr>
<td>500</td>
<td>±1.06</td>
</tr>
<tr>
<td>600</td>
<td>±1.13</td>
</tr>
<tr>
<td>700</td>
<td>±1.17</td>
</tr>
<tr>
<td>800</td>
<td>±1.28</td>
</tr>
<tr>
<td>850</td>
<td>±1.34</td>
</tr>
</tbody>
</table>

RTDs are manufactured in three and four wire configurations with single or duplex arrangements. If directly connected to receivers over long distances the cable requires careful sizing and routing to achieve satisfactory performance. Head mounted transmitters eliminate this problem allowing the standardisation of cabling systems.

6.5 Analytical
This section is a brief review of the common liquid and gas analytical measurement methods used in the process industries. It is such an extensive topic that it is not covered in detail here and other sources, such as manufacturers’ data, should be referred to for more specific information.(3)

**Liquid Analysis**

The table below gives a summary of the common analytical liquid measurements used in the processing industries.

<table>
<thead>
<tr>
<th>Liquid Analysis Measurement</th>
<th>Units</th>
<th>Technology</th>
<th>Applications</th>
</tr>
</thead>
<tbody>
<tr>
<td>Conductivity</td>
<td>mho/cm</td>
<td>Electrolysis</td>
<td>Contamination in aqueous solutions, Purified and cooling water monitoring, Steam condensate monitoring</td>
</tr>
<tr>
<td>pH</td>
<td>pH</td>
<td>Ionisation</td>
<td>Effluent treatment, Process control</td>
</tr>
<tr>
<td>Density or SG</td>
<td>kg/m$^3$</td>
<td>Buoyancy - float, Static head - dp cell, Nucleonics, Vibration</td>
<td>Composition e.g. NH$_3$ and H$_2$SO$_4$ solutions, Interface detection</td>
</tr>
<tr>
<td>Turbidity</td>
<td>Photodetector</td>
<td></td>
<td>Filter-centrifuge solids break through, Water purity</td>
</tr>
</tbody>
</table>

**Conductivity**

A conductivity cell consists of two electrodes of area $A$(cm$^2$) separated by distance $L$(cm), is used to measure the conductance $C$(mho) which is given by:

$$C = \frac{L}{A}$$

A standard cell has an $A=1$cm$^2$ and $L=1$cm measures the specific conductance, called conductivity, with units of mho/cm.

This measurement is used extensively in industry for measuring water quality from water treatment plants and is of great importance in contributing to the protection of steam generators from corrosion. It is also used to monitor the quality of recirculated steam condensate and cooling water to detect for process equipment failure resulting in acidic material contaminating the systems leading to major damage.

**Hydrogen ion concentration – pH**

pH is the hydrogen ion concentration in a solution and represents the degree of the solution acidity or alkalinity. pH is defined by:

$$pH = -\log_{10}[H^+]$$

This definition results in pH being represented by a scale of 1 to 14. Acidic solutions have pH<7, alkaline solutions have pH>7 and neutral solutions have pH=7.

pH is determined using two electrodes, a measuring electrode, typically made of glass containing a constant pH solution(buffer) and conductor, and a reference electrode to complete the circuit. The measurement requires temperature compensation.

The electrode assembly can be installed as a tank immersion unit or in a bypass sample loop flow through arrangement. On line measurement requires frequent sample checking and calibration as the probes can become quickly fouled and sluggish. It is recommended that an off line fully calibrated back up system be kept available as calibration with buffer solutions is time consuming, particularly with the immersion assemblies.

This measurement is extensively used to monitor the condition of utility streams and for the measurement and control of process effluents. Process effluents present a difficult challenge due to contamination with hydrocarbons and large variations in pH conditions.
pH Control

pH control of effluent streams is complicated by the characteristics of the titration (neutralisation) curve where large volumes of reagent are required at extreme pH values as compared to very small volumes on approach to neutrality. See diagram below.

A single equal % reagent control valve will have a rangeability of ~50 where it can be seen that a rangeability in excess of 200 will be required. To achieve this a large control valve has to be used in sequence with a small control valve.\(^{(4)}\)

If the large valve is selected to be 20 times the size of the smaller valve a flow rangeability of 700 can be achieved. The large valve should have a linear characteristic and the small valve an equal % characteristic.

A feedforward control system has been shown which provides anticipatory control for the variation in incoming feed conditions. Influent conditions monitored are flow and pH and the effluent pH is monitored to give feedback control. A similar scheme is based on influent flow and reagent flow in ratio control with effluent pH feedback trim which will not be as effective as the scheme shown.

The control equation has been derived \(^{(4)}\) as follows:

\[
\frac{m}{F} = \log_{10}\left(\frac{aF}{10^{pH}}\right) - \frac{100}{p}\left(r - pH\right)
\]

Section 7

Final Control Devices

Contents

7.1 Control Valves
7.2 Actuators
7.3 Control Valve Performance
7.4 Control Valve Specification

References


Appendices

I Control Valve and Actuator Types
II Valve Flow Coefficient Units
7.1 Control Valves

The selection of a control valve, with an appropriate operating characteristic, valve size and actuator power, is essential for satisfactory control loop performance. Appendix I shows details of control valve and actuator assemblies.

For a detailed review of valve sizing issues refer to Emerson Process Management, Fisher Control Valve Handbook, 4th Edition. Special consideration should be given to piping installation factors which can affect the valve performance and are covered in detail in this reference.

As a general “rule of thumb” control valves, fitted with full size trims, are usually sized to be less than the line size of diameter D, typically ½D. This results in valves being fitted between pipe reducers. Line size valves, fitted with reduced trims, simplify installation but with a potential increase in installation cost.

The control valve must be sized to cover the anticipated maximum and minimum operating limits of the process. Valve rangeability, maximum flow to minimum flow ratio, of 50:1 is normal. The maximum flowing quantity should be set at 15% to 50% above the maximum flow required by the process. An initial estimate would be to set a full size trim valve at half the line size. Line sized valves, fitted with reduced trims, can avoid the use of piping reducers which create an additional pressure drop. This simplifies the piping design and installation but these benefits need to be offset against the increased valve cost.

The gain of a valve is the rated flow under nominal process conditions at full stroke. The flow characteristic is the relationship between the flow through the valve and the % rated stem travel as it is varied from 0 to 100%. An equal % characteristic produces a change in flow by a certain % of present flow for a given increment in stem position and compensates for gain variations elsewhere in the control loop. An equal% characteristic has an interesting feature in that changing the valve size does not affect the loop gain, which is directly proportional to flow, making valve sizing less critical. Refer to section 6.1 for more details.

The liquid flow through a valve is a function of the available pressure, size and opening:

\[ F = C_v \frac{\Delta P}{\rho} \]

Where:
- \( C_v \) flow coefficient of the valve
- \( \Delta P \) pressure drop across valve
- \( \rho \) specific gravity of flowing liquid

Simulation Section 9 Cases 3.09 and 3.10 review the sizing methods for liquids, steam and gases. The consequences of an undersized valve are self-apparent. An oversized valve will only operate over part of its travel, requiring a wider proportional band, which results in poor control and at worst case no control.

The selection of the valve pressure drop, for design, needs to consider the maximum and minimum inlet and outlet pressures that will be experienced. In many cases, the process will set the inlet and outlet pressures and the valve must be sized accordingly. However for liquid transfers using centrifugal pumps a valve pressure drop of ~30% of total system pressure, at full flow, will give reasonable control. For valves installed in extremely long or high pressure drop lines at least 15% up to 25% of system pressure drop should be taken.

If the pressure in the valve vena contracta drops below the liquid vapour pressure, bubbles will form and limit flow through the valve. If the outlet pressure remains below the vapour pressure the bubbles remain, and the liquid is said to have flashed. If the outlet pressure rises above the vapour pressure the bubbles will implode, producing cavitation, which sounds like sand is flowing through the valve. Flashing and cavitation can cause severe damage to the valve and even pipework and should be avoided if possible by increasing the inlet pressure or selecting a valve type that has a larger critical flow factor, \( C_r \). A change in flow direction by installing a streamlined angle valve will achieve a significant increase in \( C_r \).
Control Valve Sizing

The methods are based on “Control Valve Sizing” by Masoneilan Company, 6th Edition, which is entirely compatible with ISA SP39.3, “Control Valve Sizing Equations for Compressible Fluids”. Refer to Appendix II for discussion on sizing coefficient units.

Liquid

The fundamental equations are presented as follows:

The valve sizing coefficient \( C_v \) metric equations for non-viscous liquid flow are given by:

For sub-critical flow where \( \Delta P < C_f^2 \Delta P_s \)

\[
C_v = 1.16 \frac{G_f}{\Delta P^{0.5}}
\]

Where:
- \( q \) liquid flow rate (m\(^3\) / h)
- \( C_f \) critical flow factor from manufacturers’ data
- \( G_f \) specific gravity of liquid at flowing temperature, water at 15°C=1.0
- \( \Delta P \) actual pressure drop (bar)

For critical flow where \( \Delta P \geq C_f^2 \Delta P_s \)

\[
C_v = \frac{G_f}{C_f \sqrt{\Delta P_s}}
\]

\[
\Delta P_s = P_1 \left( 0.96 - 0.28 \frac{P_v}{P_c} \right) P_v
\]

Where:
- \( P_1 \) upstream pressure (bar)
- \( P_2 \) downstream pressure (bar)
- \( P_v \) fluid vapour pressure at flowing temperature (bar)
- \( P_c \) critical pressure (bar)
- \( \mu \) fluid viscosity (cps)

Laminar flow can result at high viscosity or when the valve \( \Delta P \) or \( C_v \) is small.

Calculate turbulent flow \( C_v \) and laminar flow \( C_v \) and use the larger value as the required \( C_v \).

For laminar flow we have:

\[
C_v = 0.032 \left( \frac{\mu q}{\Delta P} \right)^{0.667}
\]

Refer to Section 9 Simulation Case 6.01
Section 8

Process Control

Contents

8.1 Fundamentals
8.2 Heat Exchange
8.3 Gas Absorption
8.4 Continuous Distillation
8.5 Batch Distillation
8.6 Batch Reactors
8.7 Utility Systems

References


Fundamentals

Heat Exchange

Gas Absorption
   (The British Library Reference WQ6 7604)
   (British Library Reference W36 7183)

Continuous Distillation

Batch Distillation

Batch Reactors

Utilities
2. Spirax Sarco, www.spiraxsarco.com/resources
8.1 Fundamentals

In 1967 Shinskey presented a novel approach to the understanding of process characteristics and the associated control systems by studying the problem in the time domain and the application of basic mass and energy balances.

The basic principle of control system design is to achieve mass and energy balances regardless of the process. Control loop interaction should be avoided or at the best minimised for stable control.

Each process parameter exhibits unique properties that determine the control strategy to be applied.

- Flow and liquid pressure periods are fast and noisy with no dead time, requiring a low gain response with reset adjustment to maintain a desired value.
- Gas pressure exhibits a high capacity with no dead time, requiring a high gain response which is adequate to maintain a desired value. This is considered to be the easiest parameter to control.
- Liquid level period can be fast or slow and sometimes noisy (boiling) with capacity, requiring medium gain and sometimes with reset adjustment to maintain a desired value.
- Temperature period can be medium or slow with capacity and dead time, requiring variable gain, reset adjustment and anticipatory response.

Control systems should not be used to overcome shortcomings in plant design. To avoid this, the process plant design and the control philosophy should be considered throughout the design process. For example, if a processing unit takes a long time to achieve stable and optimal operating conditions the feeds should be held constant by the use of surge tanks in which the level is allowed to float between acceptable limits.

Processes can be batch, semi-continuous or continuous with the selection depending on many factors. Batch processes are used in the manufacture of a wide variety of fine and speciality chemicals and are inherently transient in nature and as such can present unique control problems.

A basic control loop consists of an instrument measuring a process variable and converting it into a 4 to 20 mA dc signal for transmitting to a controller. The controller produces a 4 to 20 mA dc output signal, based on the error between measurement and set point and the control settings Proportional, Integral and Derivative (PID). The output signal is connected to a regulating device, such as a control valve or variable speed drive. A typical single loop is shown:

![Single Loop Control Diagram]

The key points to consider when setting up a control loop with a controller and control valve are:

The measurement from a process is normally transmitted using a 4 to 20 mA dc signal representing the process variable calibrated range. For example, a flow loop could be calibrated 4 mA = 0 kg/h and 20 mA = 50000 kg/h. An elevated zero can be used where 4 mA = 10000 kg/h or a suppressed zero 4 mA = -50°C.
The measurement signal on modern systems, is normally linearised with signal conditioners, for example flow measurement with an orifice plate gives a signal proportional to the square of flow. The measurement unit for level is Length and for mass flow is Mole/Mass.

The controller set point value is the desired process value in the same units as specified for the measurement.

Controller Action determines the controller output (OP) in response to the error between Set Point (SP) ± Measured Variable (MV). Negative feedback control is required which behaves as follows:

For Reverse Action: If MV > SP the OP ↓ and if MV < SP the OP ↑
For Direct Action: If MV > SP the OP ↑ and if MV < SP the OP ↓

It is essential that this concept is understood because an incorrect selection will result in Positive Feedback with the controller output continually increasing or decreasing.

Control Valve failure action is determined by the process e.g. a heating valve usually fails closed and a cooling valve usually fails open. A fail closed valve requires an increasing controller output 4 to 20 mA output to open (Signal to Open) and a fail open valve requires an increasing controller output to close (Signal to Close). An incorrect match of controller action and valve action will result in Positive Feedback.

Tuning parameters for noisy control loops e.g. liquid flow, liquid pressure and level use a wide Proportional Band (PB in %) as controller gain is 100/PB. In this case, short Integral Action Times (Ti) are used to eliminate SP±MV error. Derivative is never used on noisy signals as Td operates on rate of change of error de/dt where rapid changes in error will result in large rapid changes in controller output leading to control valve slamming and damage.

Single loop control systems have limited application with more advanced control systems requiring multiple loop configurations. The more common of these multiple loop systems are shown below:

<table>
<thead>
<tr>
<th>Multiple Loop Control Systems</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Cascade Loop</strong></td>
</tr>
<tr>
<td>Tx</td>
</tr>
<tr>
<td>Master XIC Slave CV</td>
</tr>
<tr>
<td><strong>Ratio Loop</strong></td>
</tr>
<tr>
<td>Tx</td>
</tr>
<tr>
<td>Wild PV Ratio Control PV XIC CV</td>
</tr>
<tr>
<td><strong>Auto Select Loop</strong></td>
</tr>
<tr>
<td>Tx</td>
</tr>
<tr>
<td>Hi Lo Select CV CV CV CV XIC</td>
</tr>
</tbody>
</table>

Heating and cooling valves are frequently used in split range such that the cooling valve is fully open at 0% controller output and the heating valve is fully open at 100% controller output. Split range control valve operation and function is discussed in more detail in Section 7.
8.2 Heat Exchange

Shell and tube heat exchangers are used extensively throughout the process industry and as such a basic understanding of their performance is important to both process and instrument engineers. This section provides a condensed review of the key terms and parameters involved in the design and operation of shell and tube heat exchangers.

The heat transfer performance of shell and tube heat exchangers involves many interacting parameters, which can be summarised as follows:

- Process fluid state on shell side and tube side
- Stream temperature specifications
- Shell side and tube side pressure drop
- Shell side and tube side fluid assignment
- Selection of heat transfer models and fouling coefficients for shell side and tube side
- Heat exchanger Tubular Exchanger Manufacturers Association (TEMA) layout and number of passes.

The basic layout for a countercurrent shell and tube heat exchanger is shown:

![Diagram of a shell and tube heat exchanger](image)

The fundamental equations for heat transfer across a surface are given by:

\[
Q = UA \Delta T_{lm} = \dot{w} \ C_{p(t)} (t_2 - t_1) = W \ C_{p(s)} (T_1 - T_2) \quad \text{or} \quad W = L
\]

Where:

- \( Q \): heat transferred per unit time (kJ/h, Btu/h)
- \( U \): the overall heat transfer coefficient (kJ/h-m² °C, Btu/h-ft²-°F)
- \( A \): heat-transfer area (m², ft²)
- \( \Delta T_{lm} \): log mean temperature difference (°C, °F)
- \( C_{p(t)} \): liquid specific heat tube side, (kJ/kg·°K, Btu/lb·°F)
- \( C_{p(s)} \): liquid specific heat shell side, (kJ/kg·°K, Btu/lb·°F)
- \( \dot{w} \): tube side flow rate, W: shell side flow rate (kg/h, lb/h)

The log mean temperature difference (LMTD) \( \Delta T_{lm} \) for countercurrent flow is given by:

\[
\Delta T_{lm} = \frac{(T_1 - T_2) - (T_2 - t_1)}{\ln \left( \frac{T_1 - t_1}{T_2 - t_1} \right)}
\]

Where:

- \( T_1 \): inlet shell side fluid temperature
- \( T_2 \): outlet shell side fluid temperature
- \( t_1 \): inlet tube side temperature
- \( t_2 \): outlet tube side temperature
8.3 Gas Absorption \(^{(1, 2, 3)}\)

Introduction

This section reviews gas absorption and liquid stripping processes. It discusses thermodynamics and the basic chemical engineering relationships for vapour liquid equilibrium (VLE), mass and energy balances.

Methods are discussed for interpreting vapour liquid equilibrium data, design procedures and determining satisfactory operating condition.

Gas absorption is used extensively throughout the process industries to achieve a variety of process objectives:

- Gas purification processes such as acid gas treating or sweetening.
- Water removal and drying of gas streams.
- Removal of reactant off-gases from batch chemical reactions.
- Removal of volatile organic compounds from inert gas streams.
- Removal of malodorous compounds from inert gas streams.

Liquid stripping is used extensively throughout the process industries to achieve a variety of process objectives:

- Air stripping for removal of CO\(_2\) from water.
- Air stripping for removal of volatile organic compounds from drinking water.
- Regeneration of amine solution to remove CO\(_2\).
- Steam stripping to remove volatile compounds.

Mass Balance \(^{(1)}\)

At steady state the overall material balance must be satisfied for each component. Usually the gas inlet stream condition and the liquid inlet composition are determined by the process, with the liquid flow rate being the only variable available to adjust the dependent outlet streams' compositions.
At any point in the tower the gas stream, \( G \) (mole/area-time), consists of the diffusing species, mole fraction \( y \), partial pressure \( p \) or mole ratio \( Y \) and the non-diffusing relatively insoluble gas \( G_s \) and are related as follows:

\[
Y = \frac{y}{1-y} = \frac{p}{p_t-p} \quad G_s = G (1-y) = \frac{G}{1+Y}
\]

The liquid stream \( L \) (mole/area-time), at any point in the tower consists of the dissolved species, mole fraction \( x \) or mole ratio \( X \) and essentially non-volatile solvent \( L_s \) and are related as follows:

\[
X = \frac{x}{1-x} \quad L_s = L (1-x) = \frac{L}{1+X}
\]

Since the inert gas \( G_s \) and solvent liquid \( L_s \) remain essentially unchanged in the tower the overall material balance is based on them and yields:

\[
G_s (Y_1 - Y_2) = L_s (X_1 - X_2)
\]

This is the equation of the operating line with slope \( L_s/G_s \) which passes through ordinates \( (X_1, Y_1) \) at the bottom (gas inlet, liquid outlet) and \( (X_2, Y_2) \) at the top (gas outlet, liquid inlet).

Consider a tower with gas inlet \( Y_1 = 0.0401 \), liquid inlet \( X_2 = 0.01492 \), gas outlet \( Y_2 = 0.00383 \) and liquid outlet \( X_1 = 0.04263 \); calculating for \( L/G \) gives:

\[
\frac{L}{G} = \frac{0.0401 - 0.00383}{0.04263 - 0.01492} = 1.31
\]

The composition of both product streams is dependent on \( L/G \). Setting the composition of either product results in control of the composition of the other product due to the relationship between \( X \) and \( Y \) being fixed.

**Simplified Mass Transfer Theory**

For low concentrations of a solute in the liquid phase and if solute forms a simple solution, Henry’s Law applies:

\[
p^* = H \times x
\]

The partial pressure of solute in the gas phase is a function of the gas composition:

\[
p = y \times p
\]

Combining gives the equilibrium relationship for the vapour phase concentration in equilibrium with the liquid phase:

\[
y^* = \frac{H \times x}{p} \quad \text{slope } m \text{ relates solute in gas } y^* \text{ at equilibrium to solute in liquid } x \quad m = \frac{H}{p}
\]

The two film theory considers that gas-film and liquid-film resistances are in series giving:

\[
\frac{1}{K_G} = \frac{1}{k_G} + \frac{m}{k_L} \quad \frac{1}{K_L} = \frac{1}{m k_G} + \frac{1}{k_L}
\]

A process is gas film controlled when \( m \) is small making \( K_G \approx k_G \), and a process is liquid film controlled when \( m \) is large making \( K_L \approx k_L \).

If the system follows Henry’s Law the equilibrium line will be straight and if the amount of solute transfer is small and the concentrations are low, the gas and liquid flows are almost constant giving a straight operating line which connects the inlet and outlet concentrations.
8.4 Continuous Distillation \(^{(1, 2, 3)}\)

Introduction

This section reviews the continuous distillation process by considering the basic chemical engineering relationships for vapour liquid equilibrium, mass and energy balances and their impact on control system design and operation.

Distillation is used extensively throughout the process industries to achieve a variety of processing objectives including:

- Separation and purification of hydrocarbon mixtures
- Purification of organic solvents
- Water removal and drying.
- Solvent recovery.

Distillation is probably the most difficult process to simulate and control because:

- Separation is a multi-variable process.
- Distillation units are frequently subject to process upsets from upstream units.
- Feed conditions, temperature and vapour fraction, have a dramatic effect on performance.
- Column pressure variations affect performance.
- Control loops are interactive rendering stable control difficult if incorrect selection.

Mass Balance \(^{(1)}\)

At steady state the overall material balance must be satisfied for each component. Only one product stream can be set independently making the other product stream a dependent variable. In general, distillate flow is manipulated, directly or through varying reflux ratio, requiring the bottom outlet flow to be adjusted to maintain the material balance, which is achieved by holding the reboiler liquid level constant.

The material balance based on the more volatile component (MVC) yields:

\[
\frac{D}{F} = \frac{z - x}{y - x}
\]

This shows that the flow ratio \(D/F\) determines the relative composition of each product such that if feed composition \(z\) changes then \(D/F\) must be manipulated to maintain constant \(x\) and \(y\).
Separation Criteria \(^{(2)}\)

The separation factor \(S\) for a binary mixture based on the MVC is defined as:

\[
S = \left( \frac{y}{1-y} \right) \left( \frac{1-x}{x} \right)
\]

\[
y = \frac{S x}{1 + x (S-1)} \quad \text{and} \quad x = \frac{y}{y + S (1-y)}
\]

The relative volatility \(\alpha\) of an ideal binary system, consisting of components 1(MVC) and 2, is the ratio of the pure component partial pressures at the same temperature.

\[
\alpha = \frac{p_1^0}{p_2^0}
\]

\(\alpha\) can be obtained by using the Antoine equation or the simplified Cox equation.

Antoine Equation (mm Hg,ºK,ln)

\[
\ln p = A - \frac{B}{C + T}
\]

Cox Equation (mmHg,ºK,log\(_{10}\))

\[
\log_{10}(p) = A_c - \frac{B_c}{T} + 230
\]

To convert from \(\ln\)→\(\log_{10}\) for the Cox equation \(A_c=(A/2.303), B_c=(B/2.303), C_c=(273+C/2.303)\).

Consider a tower separating equi-molar mixture of MEK and MIBK into fractions of 1 mol % MIBK in MEK and vice versa. Feed \(z=0.5\), distillate \(y=0.99\) and bottoms \(x=0.01\); calculating for \(S\) gives:

\[
S = \frac{(0.99)(0.99)}{(0.01)(0.01)} = 9801
\]

\[
\frac{D}{F} = \frac{0.5 - 0.01}{0.99 - 0.01} = 0.5
\]

We can conclude the following:

The composition of both product streams is dependent on \(D/F\).
Changes in feed composition can be compensated for by manipulating \(D/F\).
For a constant separation, control of the composition of either product will result in control of the composition of the other product due to the relationship between \(x\) and \(y\) being fixed.

Theoretical Stages

The Fenske equation can be used to estimate the minimum number of stages \(N_m\) at total reflux:

\[
S = \alpha^{N_m}
\]

Substitution gives;

\[
N_m \ln \alpha = \ln \left( \frac{y}{1-y} \right) - \ln \left( \frac{x}{1-x} \right) = \ln S
\]

For the separation of MEK and MIBK assuming a relative volatility \(\alpha=3.0\)

\[
N_m = \frac{\ln 9801}{\ln 3.0} = 8.36
\]

For a batch distillation we can expect an initial \(S\) value of 99 requiring \(N_m\) of 4.18.

Experience has shown that about 2.5\(N_m\) theoretical stages are required for economic operation. In a tray column, efficiencies of 60 to 70% are not uncommon requiring 3.8\(N_m\) bearing in mind that the reboiler represents one theoretical stage \(^{(2)}\).
8.5 Batch Distillation \((1, 2, 3)\)

**Introduction**

Batch distillation techniques are used extensively in the Specialty, Fine Chemicals and Pharmaceutical Industries to achieve a variety of processing objectives, namely:

- Removal of impurities/unwanted components from reaction mixtures.
- Water removal and drying.
- Solvent change between reaction stages on multistage syntheses.
- Concentration prior to crystallisation.
- Heat removal control for strong exothermic reactions at reflux.
- Solvent recovery.
- Fractionation of complex mixtures.

The simplest form of batch distillation involves a single flash separation which is used where a large difference in volatility exists between the components. Such a distillation does not need a fractionating column between the batch still, normally a stirred jacketed reactor, and the condenser. Simple batch distillation, also referred to as pot to pot, provides one theoretical plate of separation.

A typical batch distillation, arrangement utilizing a stirred batch reactor as the reboiler (still), packed column, overhead condenser, rundown cooler and accumulator(receiver), is shown, together with the nomenclature used in this section.

When the difference in volatility between the components to be separated is small or when operating over narrow composition ranges a rectification section is required between the still and the condenser. Overhead facilities are also required to provide control of reflux ratio and layer separation when handling heterogeneous azeotropes.

In operation, the system is brought to steady state under total reflux. Overheads are continuously withdrawn in accordance with the reflux control strategy. Cuts (fractions) are taken by switching to different accumulators(receivers), following a time, temperature or overhead composition strategy.
Batch distillations can be operated as follows:

- Constant reflux ratio, giving varying overhead composition. Distillation is continued until desired composition is achieved in the still or the distillate receiver.
- Varying reflux ratio, giving constant overhead composition. As the distillation proceeds the still is depleted of the lighter component with the reflux ratio continually increasing. The distillation is terminated at a maximum economic reflux ratio or when the desired still composition is achieved or when still heat input cannot sustain the reflux ratio.
- Repetitive total reflux. The unit is operated at total reflux until equilibrium is established; then distillate is withdrawn as total draw off for a short period of time before returning to total reflux. This technique is useful when required to separate a low-boiling trace component.
- Minimum time by varying reflux ratio. This provides the most cost effective mode of operation consistent with achieving the desired separation.

The distillation is normally operated at atmospheric pressure. However reduced pressure operation is sometimes required to achieve the desired separation, to reduce operating temperature for economic reasons or for processing temperature sensitive materials.

For multipurpose operation careful consideration is required when selecting column internals to achieve acceptable column loadings and operational turndown. The boil up rate achievable with stirred jacketed reactors is dependent upon many factors including operational temperature difference, jacket heating media and heat transfer considerations, refer to Section 6.6.

**Operation and Control**

The operation and control of a batch distillation is constrained by the mass and energy balances, with the primary objective being to minimise distillation time. If conditions are fixed at the still then conditions at the top of the column can be varied and vice-versa.

A common strategy is to control the heat input at the still to maximise the distillate draw rate and sustain the desired reflux ratio. The still bottoms temperature will increase as the distillation proceeds reducing the temperature driving force. If the heating medium is steam the heat input can be controlled by the flowrate or pressure and if heat transfer fluid the heat input can be maintained by increasing the temperature. Alternatively if the heating medium conditions are fixed, the distillation pressure can be reduced to increase the temperature driving force, if necessary, in incremental steps as the distillation temperature increases throughout the distillation.

At constant pressure, the top and bottom temperatures are an indication of composition. If the product is the still residues the distillation is continued until the bottom temperature reaches a target limit. If the product is the distillate the column top temperature is controlled by increasing the reflux ratio as the distillation proceeds to maintain the target temperature. There will come a point when the still heat input will not be able to sustain the increasing reflux ratio at which point the column top temperature will start to increase and the distillation is stopped.

The operating instructions can determine the distillation by specifying temperature cut off values or time steps can be given with an established distillation. If repetitive batch distillations are to be carried out it may be desirable to shorten the distillation time by the use of intermediate fractions between the component fractions.

The basic instrumentation required will include still bottoms and column top temperatures, still pressure and possibly control, reflux and distillate flow control and level measurements as determined by the plant configuration. Still heat input control will be determined by the heating medium used and the jacket services configuration.

To start the distillation, apply heat to the still and continue until stable conditions at total reflux have been achieved. Now start distillate draw at fixed reflux ratio for bottoms composition or variable reflux for top composition. If there are distillate cuts involved separate receivers will be required.

It is not desirable to store hot liquids e.g. hydrocarbons, nor to cool the reflux as this causes internal reflux which leads to additional still heat input, so rundown coolers should be installed appropriately.
Utility systems are a key part of the process plant operating infrastructure providing heating, cooling, compressed air, vacuum, water, inert gases and ancillary systems. A high level of reliability is required for consistent production and process safety.

Utility site energy usage can be as high as 70% with up to ~15% being lost when equipment efficiency and operations are taken into consideration. The application of combined heat and power (chp), to produce electricity, can significantly improve the efficiency. Process simulation can provide a powerful tool in the optimisation of utility systems.

Total integration of utilities with the production facilities will result in significant economic benefits. The sizing of individual utility units must consider the turndown capability in relation to the maximum and minimum demands of the processes, redundancy issues for maintenance and future expansion. Typical turndown ratios are in the 3 to 5 range, with the efficiency reducing significantly at low turndowns. The economics of running large units, sized based on maximum demand, as opposed to operating smaller units in parallel requires careful consideration; utility units are frequently oversized. For example, if a single process unit has the requirement for a small duty at very low temperature, consider a local refrigeration unit, sized accordingly, rather than make the central refrigeration unit operate at this low temperature.

Utility distribution networks need to be designed to ensure the supply is at a satisfactory pressure, flow and temperature, for all process units under all operating conditions. There are few process plants that do not experience bottlenecks due to poorly designed networks. Header pressure control using pressure control loops at the supply to return connections and flow conservation devices at the individual process units will minimise distribution problems.

The following summary is not comprehensive but highlights some issues that should be given consideration in the design and operation of utility systems.

**Heating Systems - Steam**

Steam generated at about 10 barg (dry saturated temperature 184°C) is suitable for most process applications including tank heating, steam ejectors, distillation and stripping. Process temperatures up to ~310°C would require steam pressures of 100 barg making hot oil the preferred option due to the lower operating pressure. Steam generated at superheated conditions can be used for driving steam turbines with the exhaust steam then passing to the process. If superheated steam pressure is not reduced through steam turbines the steam should be desuperheated before passing to process heating duties.

When correctly applied, steam provides the most efficient heating medium. Typically steam heat transfer film coefficients, as compared to those for heat transfer fluids, can be greater by factors of 5 to 10. This can have a significant impact when considering boil up rates on distillation processes. The following summarises some general application points:

- Generate and use steam at the lowest pressure consistent with providing adequate ΔT for heat transfer.
- Match steam generation to process usage to minimise or prevent venting to atmosphere. Provide back pressure control on steam generators to prevent carry over of condensate into the steam mains to avoid severe corrosion and stress corrosion cracking.
- Pressure let down stations, using control valves or orifice plates, waste energy and can create superheated steam which is inefficient for heat transfer. Note that the heat given up by superheated steam down to the dry saturation temperature is small compared to the enthalpy of evaporation released when at saturation.
- Reducing pressure through steam turbines releases work which can be used for electricity generation or pumping applications. Superheated steam is good for this.
Saturated steam is best for heat transfer provided that air is not allowed to accumulate. Correctly located steam traps with air removal capability are required.

Typical steam line sizing velocities for saturated steam are 20-30 m/s (0-1 barg), 30-40 m/s (>3.5 barg), superheated steam (>13.5 barg) 45-60 m/s and hot condensate 0.5-1 m/s.

Condensate is highly corrosive and should be removed efficiently using appropriate steam traps. Condensate build up in heat transfer equipment will reduce the area for heat transfer. If steam flow to a unit is being manipulated to control temperature, condensate removal should be by means of a float trap, essentially a level controller, to prevent control hunting. Alternatively, steam traps based on the venturi orifice principle, have no moving parts, provide smooth control, air removal capability and improved efficiency.

Condensate can become contaminated, as a result of leaks from process equipment, requiring the location of the hot well to be considered carefully and the provision of condition monitoring equipment.

**Heating Systems - Heat Transfer Fluids**

Heat transfer fluids can be used in the temperature range -100°C to 400°C and provide the benefit, when correctly applied of being single phase liquid systems.

Single phase systems avoid the requirement for automatic routing valves to connect different services and provides seamless heat-cool changeover for temperature control systems.

Eliminates the requirement for water treatment, reduces the corrosion associated with water based systems and maintenance costs.

Heat transfer fluid specific heats are much lower than those for water systems. This requires significantly higher flow rates for the same heat removal capability.

Heat transfer fluid viscosities can increase dramatically at low temperatures affecting heat transfer capabilities.

The presence of water can result in steam formation at high temperatures and freezing at low temperatures. Methods to remove water include running at high temperature and venting off the steam at high points which can result in pump damage due to cavitation, passing through molecular sieves and the use of an additive, such as a glycol, to capture the water.

Contamination with water due to heat exchanger failure can result in serious outages. The facility to isolate individual systems will allow the leakage source to be identified more readily.

**Cooling Systems**

Use the highest temperature coolant consistent with providing adequate ΔT for heat transfer; do not use refrigerated coolants when cooling water can achieve adequate cooling. Optimise batch process cycle times by switching coolants when appropriate.

The variation in ambient conditions can affect cooling tower efficiencies resulting in issues associated with vacuum plant.

On low temperature systems the heat gain from ambient and energy from pumps is significant. Use high efficiency insulation and consider variable speed drives.

Protect refrigeration equipment from peak loads by using hot-cold side tanks for thermal storage. To conserve energy consider the use of variable speed drives on main circulation pumps to match process demand.

Monitor the freezing points of water based refrigerants using ethylene glycol or propylene glycol (preferred in the food and pharma industries) to check for dilution from other water services.

Cooling water systems impose an onerous requirement for water treatment, monitoring for microbial contamination and avoidance of legionella. Careful consideration should be given to the sanitisation of water based systems.
Section 9

Process Simulation

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9.1 Introduction

This section presents cases on a variety of applications, discussed previously, using the CHEMCAD simulator by Chemstations Inc., Houston.

Simulators provide process control engineers with a powerful tool to assist in the design, optimisation and operation of processes. Prior to the introduction of process simulators, this activity involved a team of engineers working many man hours and maybe not achieving a definitive result. The simulator characteristics are determined by the process. Batch and semi-batch processes require dynamic simulation whereas continuous processes are studied with both dynamic and steady state simulators.

Steady state simulators model plant behaviour using continuous, time interval based, steady state approximations and require the least effort to set up and calibrate the flowsheet. Dynamic simulation can also be considered as a series of steady state simulations.

Dynamic simulators are based on real time or accelerated dynamic simulation principles and can provide a true representation of the plant behaviour, including retention times and thermal lags. Continuous steady state with varying throughput, start up, shutdown, feed and composition change can all be studied. Dynamic simulators can also be used for operational plant troubleshooting, control loop tuning and real time optimisation. Dynamic simulations require more information and are more difficult to set up.

A process simulation involves taking the input stream flow rates, compositions and thermodynamic conditions, performing a series of iterative calculations as the streams are processed through Unit Operations and recycles, finally leading to the output stream flow rates, compositions and thermodynamic conditions.

The iterative process can be based on

- Direct substitution, the value is recalculated after a single iteration
- Wegstein acceleration, the value is recalculated after three iterations - speed up frequency 3
- Dominant Eigen value, the value is recalculated after four iterations - speed up frequency 4

The number of iterations is limited, default value 40, until the convergence parameters are satisfied, nominally set at an absolute value of 0.001.

The chart below shows the steps involved in setting up a simulation.
9.2 Dynamic Simulation

The following provides a summary of key points to achieve rapid results from the dynamic models.

Run time and step size must be selected to ensure consistency with real time behaviour. Dynamic simulation is provided with a speed up facility called Operational Time Scale (OTS) which allows for simulation speed up over real time by a factor up to 50.

Under “Run – Convergence” select
“Run one time step for dynamic simulation” to allow study of model behaviour.
“Run Data Map at each dynamic time step” to transfer data between Excel at each step.

Control valve rangeability based on the ratio of the maximum to minimum controllable flows depends on the entire control loop, use 50 for equal % characteristic and 10 for linear characteristic.

The position of the control valve in the model is critical and the rules can be summarized:-

It requires inlet (P_i) and outlet (P_o) pressures to be specified where P_i is taken from the inlet stream. P_o can be specified directly or overtaken indirectly by an assigned UnitOp in the model.

A mass balance must be maintained around the control valve. To achieve this, locate immediately after a Feed Arrow (direct flow adjust), or after a Divider (indirect flow regulation). Dynamic Vessel, Dynamic Reactor and Dynamic Column can also be assigned for indirect flow regulation. The optimal mass flowrate transfer is used to set up Indirect Flowrate Regulation.

If a value for valve flow coefficient C_v is not known use the Equipment Valve Sizing option which will provide a value at stated stream conditions. The CHEMCAD valve data base smallest valve C_v is 9.0; a smaller valve C_v requirement is determined by separate calculation which can be linked to the model using the Excel Data Map facility. An incorrectly sized valve will result in dynamic model convergence problems and an unstable condition in the model.

If the controller input goes out of the calibrated range the control output continues to change resulting in unreal conditions.

The control valve can be operated in various operating modes which allow the simulations to be driven to the desired condition. These features allow control valve positions to be set manually or automatically from an assigned controller.

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9.3 Simulations
Process Instrumentation and Control Simulations

Case 3.01 Control of Dead Time

Dead time (transport lag or distance-velocity lag) is the interval after the application of a control device output change for which no process response (measurement) is observable. It occurs when mass or energy is transported along a path length with velocity constituting the delay.

A 1 minute dead time process will cycle with a 2 minute period under proportional only control.

The gain of a dead time process is 1 so to achieve a loop product gain of 1 the proportional band must be set at 100%. Controller gain is 100/P.

To dampen oscillations the proportional band is increased to 200% to obtain ¼ amplitude damping.

Dead time is a difficult element to control and process design should concentrate on minimising or eliminating if practical.
Section 10

Implementation

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10.5 Maintenance

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I Static Earthing
II Equipment Ingress Protection (IP) Rating
III Process Equipment Specifications
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V Instrument Specifications
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References
10.1 General Considerations

To design and build a process plant facility involves many engineering disciplines, including process, mechanical, instrumentation, electrical and control engineers who are required to work in a coordinated manner as a project team. The main work phases involved in the design process are summarised in the following chart.

The process concepts are defined using Process Flow Diagrams (PFD) to develop the mass and energy balances. The PFDs show the process stream compositions and conditions of flow, pressure and temperature together with all feed, recycle, product and waste streams. Individual processing units, referred to as Unit Operations (UnitOps), are included and define the basic processing requirements for reactions, purifications and separations. This phase allows the development of the process economics and identifies hazards, production requirements and environmental challenges.

Process and Instrumentation Diagrams (PIDs) are developed to show all utility systems and processing units. The PIDs show the process and ancillary equipment with all utility connections, process piping, valves, pressure relief devices, instrumentation and control loops.

Equipment, piping and instrumentation are individually identified with unique Tag Numbers which will probably include information giving the processing area/unit, the type of equipment and in the case of piping the line number, material and size.

Equipment numbering systems vary throughout different industries and companies but the need for a consistent and unique system cannot be over emphasised. The more complete the information the numbering system conveys provides a major contribution to inter discipline communication during the design, construction, operation and maintenance phases. Any change to the numbering system mid-project will definitely lead to confusion, increased costs and project delays.

Document control is crucial to the satisfactory running of the design process and should include the relevant signatories in accordance with in house procedures. All reports, specifications and drawings should be uniquely numbered showing revision status and date of issue.
A summary of the documentation that can be generated for a typical process plant is shown in the diagram. The documentation should be carefully structured to ensure adequate cross referencing and to avoid repeating design specific information on different reports, specifications and drawings to simplify the revision process.

Procedures for checking, approving and changing should be clearly established as an auditable document history is essential for legal, safety and commercial reasons.

One method is to maintain a master set of documents which are highlighted with all the changes at each revision stage. In the early stages of a project, when the design is being evolved, it is advisable not to initiate a rigid change control procedure too early as this can lead to confusion and delays. One method is to use a numeric revision numbering system prior to initiating the change procedure to be followed by a letter revision numbering system when the change procedure is initiated.

Many documentation structures and procedures can become inflexible which if applied to a process design still in the evolution phase can lead to excessive costs and project delays. Design change procedures can become onerous and defeat their purpose, causing confusion and delays not to mention expense.

The documentation in support of the process, instrumentation, control and electrical phases should be structured, where possible, so that it will be useful throughout the life of the plant. Documentation essential for the build phase should be kept up to date only until completion of the build phase since it has no further use once plant operations start.
10.2 Control System Structures

Modern control systems are mainly based on Distributed Control System (DCS) or Programmable Logic Controller (PLC) in combination with Supervisory Control and Data Acquisition (SCADA) (PLC/SCADA). A typical control system structure is shown:

PLC/SCADA has evolved into a total Process Control System (PCS) structure which is capable of discrete and advanced control functions. PLC/SCADA systems can be successfully implemented by Small to Medium Enterprises (SME) who provide flexibility in system development and implementation which allows for the system to be readily expanded.

DCS systems tend to be supplied and implemented by major international manufacturers which make the end user entirely dependent on single source support. This approach requires a stable design situation supported by a comprehensive User Requirement Specification of the total facility, frozen at an early stage. Changes in the course of the project can cause significant delays and be costly. The development of PLC/SCADA and DCS can be summarised in the following:

Initial PLC applications were in discrete control of manufacturing processes where most of the inputs and outputs were digital. Little processing power is needed for handling digital signals so operation was very fast being ideally suited for sequence control of machine tools, process shutdown and alarm systems and general manufacturing processes.

Initial DCS was designed primarily to control processes involving large counts of input and output analogue signals (4-20mA or 0-10V) representing process variables of temperature, flow, pressure, pH, conductivity and others. Advanced control strategies can be programmed into DCS to provide accurate control of critical processes involved in food and pharmaceuticals manufacturing. Speed was not as critical but accuracy and complexity was.

The "distributed" part of the name was used to describe that the control functions could be spread out among many sub-assemblies located in remote locations. Single mainframe computers were first applied to control plants in the 1960s but were vulnerable in that a single failure could cause a total plant shutdown. Distributed control gave a more robust and resilient solution.

Since the early 1990's computers have developed at an incredible rate involving increased memory and processing speed. There is no longer a clear distinction between DCS and PLC/SCADA which can now perform simple to advanced process control functions. PLC input/output capabilities have been developed to handle all types of signals and can be implemented in a flexible distributed format and is now referred to as Programmable Automation Controller (PAC).
10.3 Design

This is a major topic involving many standards and guidelines and it is not possible to treat in detail here. The section covers the highlights of key issues and describes some methods and practices that have been applied successfully over a wide range of industry sectors and projects.

Concepts

Following initial process development in the laboratory and pilot plant the User Requirement Specification, Process Flow Diagram and Process and Instrumentation Diagram are the starting point for the development of the design concepts for the process control system.

The nature and extent of the process will determine if local or centralised control is suitable. In most modern facilities a centralised control room will be considered most appropriate requiring a decision between a conventional control system, with individual displays and alarms or SCADA/PLC or DCS.

A sound technique is to develop these concepts into single line Overview System Diagrams which will show the hazardous area classifications, the location of equipment racks, control panels, instruments, equipment, displays, inter-connections and communications concepts.

The process plant layout is driven by factors external to control system considerations. In the initial stages the layout will depend on factors such as geography, location, accessibility and external facilities. The nature of the process will determine the location of the utilities, individual processing units, storage, feed stock and product shipment facilities.

The selection of the control room location is critical for the safe and satisfactory operation of the facility. The optimum location for operational purposes will probably result in the control room being located centrally within the processing facilities which may prove to be unacceptable for personnel safety considerations.

If a single centralised control room is considered unacceptable the distributed control features of both PLC/SCADA and DCS allow for satellite control rooms to be connected by secure communications networks. The final location(s) will be determined by the process logistics, safety and operational economics.

The importance of developing comprehensive and meaningful numbering systems to identify instruments, equipment racks, panels, junction boxes, cables, wiring and terminals cannot be over emphasised.

The more information a numbering system can convey the better for all phases of a project and subsequent operations. A well thought out system can minimise the requirement for referring to additional documentation to navigate around a control system.

To avoid having a running number involving all documents the numbering systems should be developed to distinguish between document types. Alpha numeric numbering systems provide a simple method of achieving this objective by applying prefixes or suffixes, for example reports-RPT, specifications-SPC, P&I Diagrams-PID and so on.
Process Diagrams

Process Flow Diagrams (PFD) show the process plant in schematic format indicating the main items of equipment and the process and utility stream connections. Mass and energy balances are shown for each process and utility stream.

Process and Instrumentation Diagrams (P&I) are developed from the PFD and show much greater detail. The P&I Diagram is the basic working document generated and allows for the implementation of the detailed engineering design, construction and operation of the plant. This diagram is critical as it is the source document that facilitates interaction between all the disciplines involved in the project.

Standards for graphical representation of process, plant and instrumentation are employed on the diagram. Larger organisations have tended to develop their own standards but International standards are also available, including:

- BS 1553 Part 1: Graphical Symbols for Piping Systems and Plant
- ISO 3511/I-1977: Symbols for Process Measurement Control Functions and Instrumentation
- ANSI/ISA-5.1-2009: Instrumentation Symbols for Identification

The symbols used are to be included within a legend. Any variations to standards and custom symbols should be highlighted within this legend. Abbreviations should be clearly defined.

When graphical symbols are used that are similar or identical in shape or configuration and have different meanings, then adequate steps must be taken to avoid misinterpretation. This requirement is critical where symbols taken from different disciplines are intermixed and their misinterpretation might cause danger to personnel or damage to equipment or have economic consequences.

If P&I Diagrams are provided with sufficient detail it can minimise the requirement for extensive detail drawings such as interfaces between piping and process equipment. The extent of the detail shown will vary between organisations but the following practices have been successfully applied in the design and build of medium size facilities.

All process and utility equipment are shown. Interconnections between diagrams clearly identified with routings on to and from the diagram with a suitable symbol. If a process stream is routed to atmosphere the discharge location requirement is identified to prevent injury to personnel or recycling of obnoxious materials into ventilation or drainage systems. Appendix VI shows a typical utility system P&I Diagram.

The extent of supply of package units is shown indicating piping connection points and specifications.

 Separate diagrams are used for the utility systems showing routings to the process equipment being supplied.

 All items of equipment, valves, fittings, instrumentation and piping are identified with a unique number. Manual valve symbols indicate the valve type and show an application identifier with size. Set values for relief valves and other safety devices are shown.

 Piping identification includes an application identifier, direction of flow, line size, pressure and temperature rating and materials of construction. Piping size changes are shown and pipe specification changes identified. Specification changes are extremely important and can have a serious safety and economic impact if shown incorrectly. Trace heating and insulation requirements should be included.

 Main instrumentation, control loops, alarms, interlocks and control devices are shown with an indication of controller action and loop failure conditions.
10.4 Testing and Commissioning

As equipment is received from suppliers it is important that there is an adequate acceptance procedure in place to ensure compliance with the purchase order and equipment specifications. In the case of instrumentation this may include carrying out an initial functional test and calibration. It is important to identify any procurement errors as early as possible as impact on project schedule could be severe.

In the initial phases, control system testing will involve carrying out factory acceptance testing of individual control panels, equipment racks and ancillary cubicles for compliance with the design drawings. If the system includes software and graphics the functionality is validated against the user requirement and functional specifications using pre-agreed testing protocols.

Once the system is installed on site testing then involves the checking of the cabling systems and then establishing power to individual circuits following a set procedure.

The documentation used to control all phases of testing varies between organisations but the guiding principles are the same. The testing should be carried out in an orderly manner with the progress being noted and controlled using an agreed and consistent method. Methods used, include the use of test sheets/schedules and/or highlighting a set of working copy drawings. Once a fault is identified the rectification and re-testing should be carefully controlled.

Control loop testing involves ensuring the components comply and are calibrated in accordance with the specifications and that the loop functionality regarding control and failure actions are correct.

Instrumentation connected to the process should be checked to ensure compliance with the specified connection details including isolation and insertion requirements. Caution should be used during pressure testing and line cleaning operations to protect against damage to instrumentation as some devices can be damaged by overpressure.

Package Units

Package units are designed, built and installed by specialist suppliers and contractors. Normally the unit will be pre-assembled and pre-wired ready for connection to the plant piping, power and cabling systems. The unit will have been tested and accepted at the factory so testing on site should only involve ensuring that the external connections are in accordance with prior agreements and drawings.

If software is involved in the unit operation it is normal practice for the package unit supplier to include this in their scope of supply. It is not advisable to incorporate package unit software into the main plant system as this can lead to conflicts of interest.
10.5 Maintenance

The importance of generating and maintaining accurate records of any maintenance activity cannot be over emphasised. The records are required to ensure compliance with legal requirements for certain items of equipment or systems and will be referred to in the event of any abnormal operation or incident.

Instrument Calibration

Instrumentation equipment should be calibrated at a frequency to be determined by the type of equipment and importance relative to the safety and economics of the operation. Safety Instrument Systems, that carry a SIL rating, will require special consideration to ensure compliance with the relevant standards and codes of practice; refer to Section 4.

A system should be in place where calibration records are held on a database with secure access only provided to authorised personnel. Calibration certificates should be generated and retained as determined by the plant quality instructions.

All details relating to the site, operating unit and instrument are entered into the database. Only authorised persons should be allowed to enter or modify data on the database.

The database should retain details of test equipment used in instrument calibrations.

A single instrument calibration form is used to record details of individual instrument calibration. Each record details calibration date, calibration frequency together with accuracy details. The test instrument database is automatically linked to the calibration databases. This ensures that test instruments out of calibration cannot be used for calibration of instruments.

Where possible the automatically calculated 5 point rising and 5 point falling calibration figures are utilised, if this is not possible, then details of the calibration are entered into the report section. If any adjustments are made to the instrument, then the adjustments made box must be clicked.

Full instrument loop calibrations will be required during on-going maintenance and operational checks and should be recorded in the database. Data entry uses the Loop Calibration form but only 3 points are checked.

Instrument calibration certificates and loop calibration certificates should be retained on summary registers.

Maintenance - Demonstrating Prior Use

Prior Use of a component within a Safety Instrumented System (SIS) can be claimed by the end user, provided certain requirements are fulfilled. This is discussed in detail in Section 4.4.

The collection and analysis of reliability data from the field is required in order to confirm the claimed reliability of a component within a SIS. BS EN 61511:1 – Clause 3.2.60 Defines Prior Use as:

“When a documented assessment has shown that there is appropriate evidence, based on the previous use of the component, that the component is suitable for use in a Safety Instrumented System”

What is essential in providing successful data collection is that all events relevant to the SIS are recorded and analysed. This not only includes failure data, but also proof testing, activations (both genuine and spurious), maintenance activities, anything which occurs with the system and component.

In order to perform reliability calculations it is necessary to build up a history of how long each component has been in service, its operating hours, how often it has been activated, tested, maintained together with a description and record of any failures.
Section 11

General Engineering Data

Contents

Dished End Dimensions
Commercial Steel Pipe Dimensions
Stainless Steel Pipe Dimensions
Tube Dimensions
Plastic Lined Pipe Dimensions
Pipe Roughness Coefficients
Valve and Pipe Fitting Diagrams
Control Valve Sizing Coefficients
Worcester Ball Valve Coefficients
Atomac Valve Coefficients
Typical Overall Heat Transfer Coefficients
Typical Fouling Resistance Coefficients
Shell Tube Count Data

Reference

### Depth Ratio = 0.249 to 0.317

![Diagram of storage tank ends]

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