1.3. <u>PROCESS MONITORING AND CONTROL</u>

An important part of any chemical process is the instrumentation that is provided to monitor and control it. The common element of the two functions, monitoring and control, is measurement. Not every quantity that is measured gets controlled but every quantity that is controlled must be measured. Many non-controlled quantities are measured in order to understand how the process is functioning and to detect if it is drifting away from normalcy. From the point of view of the process operator, the more things that are measured the better, including sometimes hard-to-measure but important quantities. From the point of view of management and the project person who installed the process, the less things measured the better because instrumentation costs money. It is often a tug-of-war between the two parties.

The following topics are examined in this section

- essentials of process control
- how instruments work and how they fail
- interaction of controls
- good control and bad control
- distributed control systems.

1.3.1. <u>Essentials of Process Control</u>

Every process requires control, whether by human intervention or by automated instrumentation or by some combination of both. Three terms that are used in process control are
- *controlled variable, CV*, the quantity which it is desired kept at some value or within some range
- *manipulated variable, MV*, the quantity that will be adjusted to keep the *CV* at its desired value
- *set point, SP*, the desired value of *CV*.

For instance it may be desired to keep a vessel's temperature (the *CV*) at a value of 120C (the *SP*) and to do so by adjusting a heating coil's steam pressure (the *MV*).

There are two broad modes of process control: *feed-back* and *feed-forward*. Operator-instituted control is usually feed-forward: for instance, it is desired to make a specific product in a process and it is known that specific process settings (e.g., temperature, pressure) are suitable, so the operator makes those settings before having any product to evaluate. Automated control is usually feed-back: the *CV* departs from the set point and

the controller takes corrective action. Some degree of feed-back is always necessary because the process is always subject to unforeseeable disturbances. However when the disturbance is foreseeable, then it may be very beneficial to apply a degree of feed-forward. For instance, consider a pair of liquids being blended in a vessel to produce a single stream of constant desired composition. If the flow of one component, 'A', is subject to change from time to time, then there is no point in waiting for a deviation measured in composition of the outlet stream before adjusting (manipulating) the flow of the second stream, 'B'. Instead a *ratio controller* is used to adjust immediately the flow of the second component.
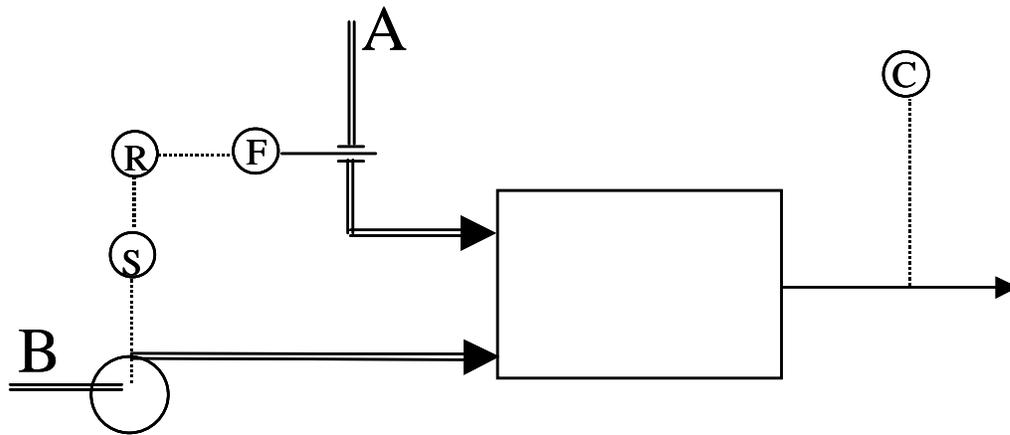

Figure 1.3.1 - Ratio Control, an Example of Feed-Forward Control

Likewise, if the set point for a controlled variable is changed, the manipulated variable can be given a feed-forward head start in helping the *CV* achieve its new value. However, the feed-*back* between *CV* and *MV* remains in place and provides the final adjustment.

By far most control comprises one manipulated variable applied to one controlled variable, i.e., a 'Single Input, Single Output' or SISO control loop. At its simplest the 'loop' consists of an *element to measure* the controlled variable, a *controller* to compare the measurement with the set point and to send an appropriate corrective signal, and a *final control element* to receive the signal and apply it to the manipulated variable. Because the measurement element and the control element act in analogue mode, if the controller acts in a digital mode then analogue-to-digital and digital-to-analogue converters must also be included in the control loop.
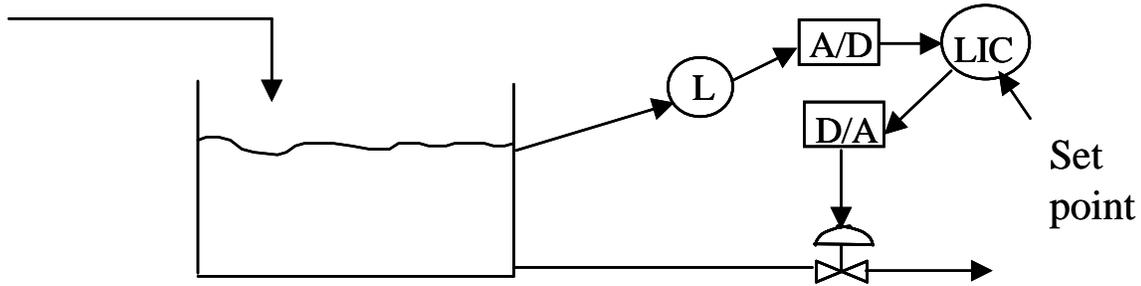
Figure 1.3.2 - Single-Input Single-Output Level Control

What is the form of corrective action taken by a controller to keep the controlled variable at its set point? The original controllers operated pneumatically and consisted of combinations of nozzles, bellows and valves, so the nature of the control action had to be suited to the capabilities of these devices. With the advent of electronic controllers and then of digital devices, the scope for sophisticated control action opened up. However most control still starts with the original scheme, which is described here. It consists of

- *proportional* action whereby the signal sent by the controller is proportional to the amount of deviation of *CV* from *SP*

- reset or *integral* action whereby the signal increases with not only the amount but also with the time that *CV* deviates from the *SP*

- *derivative* action whereby the signal adds an extra component proportional to the speed with which *CV* is departing from *SP*.

Taken together these three actions constitute proportional-integral-derivative or PID control. The 'tuning' of a controller consists of setting the magnitudes of these three actions. Sometimes the third (derivative) action is suppressed, particularly if there are rapid fluctuating disturbances in the system which would make this action unstable. Proportional control by itself is insufficient to drive *CV* all the way back to *SP* if there is some sustained disturbance affecting *CV*. Integral action is needed to keep increasing the corrective signal until it is sufficient to drive the *CV* back to its correct value. Unfortunately the size of the integral signal accumulates while the *CV* has been off-spec, so that when it does return to the correct value the action persists and drives *CV* the other way. This phenomenon of 'reset wind-up' therefore limits the aggressiveness of integral action. Proper tuning determines the optimum size of the integral action.

Mathematically, the PID control equation is

Corrective signal = K_p x { error + K_I x integral error d(time) + K_D d(error)/d(time) }

(1.2.1)

where **error** is the difference between *CV* and *SP*.

It is seen that the value of $\mathbf{K_P}$ affects the strength of all three control actions.

The inverse of $\mathbf{K_I}$ known as 'reset time' and $\mathbf{K_D}$ is called 'derivative time'.

In practice this equation is modified in subtle ways, such as distinguishing whether the error is due to an actual deviation by *CV* or to a change in *SP*. In more sophisticated control the value of $\mathbf{K_P}$ may be made a function of the magnitude of the **error**: this change allows for a non-linear relationship between corrective signal and its effect.

1.3.2. <u>How Instruments Work and How They Fail</u>

As seen in Figure 2.2, the control loop comprises at least three, generally five and sometimes more elements. Controllers and digital-analogue converters are generic devices that function well-removed from the actual process and that are unlikely to fail. Final control elements are closer to the process and have more of a tendency to mal-function: control valves, for instance, may plug or may stick. But the instruments that require the greatest attention are those that do the measurements. They are the focus of this section.

The following table shows most of the combinations of *controlled variables* and *manipulated variables* that are found in chemical processes.

Table 1.3.1 – Common Control Loops in Chemical Processes

Controlled Variable	Manipulated Variable
Temperature	Heating fluid pressure
	Heating fluid flow rate
	Heating fluid temperature
	Heating fluid valve opening
	Electrical current
	Electrical voltage
Fluid flow rate	Valve opening
	Pump speed
	Pressure or pressure differential
Liquid level	Rate of in-flow or out-flow
	Valve opening
	Pump speed
Pressure	Valve opening
	Pump speed
	Relief flow

Other quantities that are controlled directly or by inference are compositions including pH.

It is seen that the same *CV* may be controlled by a choice of *MV*'s. Where the *MV* is itself a process quantity such as flow or heating-fluid temperature, the implication is that the *MV* becomes the *CV* of a secondary control loop and is controlled by some concrete action like a valve motion.

In this section measurement methods are reviewed for the four control quantities listed in Table 2.2.1. The principle is described and precautions are given that helps avoid 'failure' of the instrument to deliver a reliable measurement.

Temperature

The simplest and most common device is the *thermocouple*. It makes use of the fact that when wires of two dissimilar metals are joined at two point of unequal temperature, a small voltage difference appears between the two ends. One end is the point where the temperature is to be measured. Theoretically, the other end should be kept at fixed temperature but in practice and variation is compensated for electronically. The temperature at the measurement point is inferred from the electromotive force (EMF) at the other end.

E.M.F.
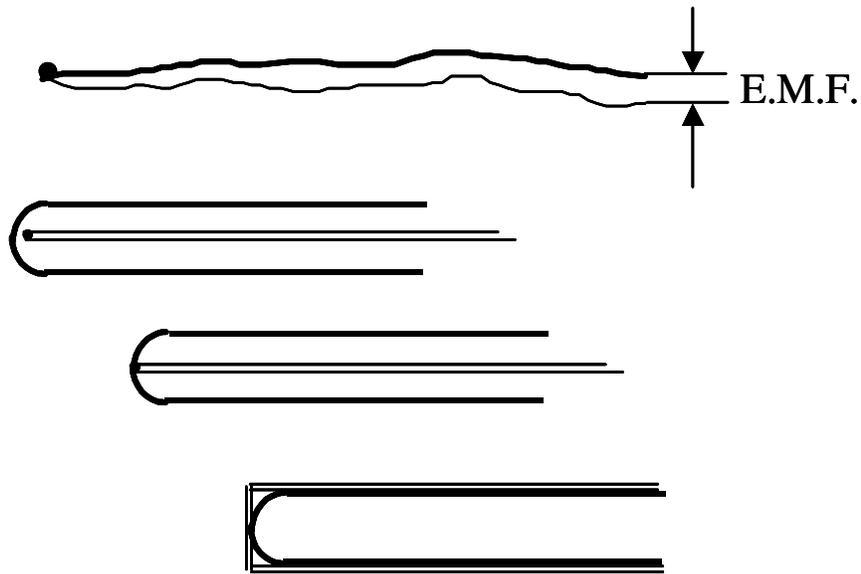
Figure 1.3.3 – Thermocouples, Sheathing and wells

Advantages of thermocouples are
- low cost and simplicity
- high temperature capability (750C, 1250C depending on the wire materials)
- potentially quick response

Disadvantages are
- only moderate accuracy e.g., +/- 2 degrees C)
- long-term instability
- dependence on measurement of small EMF
- susceptibility to extraneous electrical paths

A thermocouple can be used bare-ended but generally they are provided with a closed sheath to protect them from process fluids. The figure shows the thermocouple both insulated from the sheath and in contact with the sheath. The latter configuration gives faster response but requires that the sheath be insulated from the measurement circuitry. If the device is positioned in a thermo-well then it must make good contact with the end of the well.

Another electrically-based instrument is the *resistive temperature device* or *RTD*. The principle is that the resistance of a metal changes with temperature. By making a length of wire (wound into a coil) subject to the temperature to be measured and building it into an automated Wheatstone Bridge so that its resistance can be measured, the temperature is determined.

Resistance
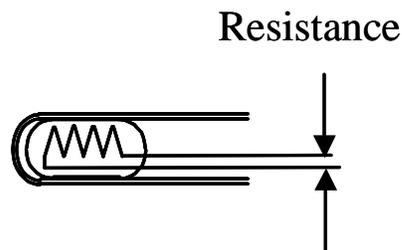
Figure 1.3.4 – Principle of the RTD

The wire coil occupies a finite space as opposed to the point junction of the thermocouple, so the measurement is not as localized.

Advantages are
- accuracy, e.g., +/- 0.15 degree C
- long-term stability

Disadvantages are
- cost
- fragility
- upper temperature limit of about 500 C

Whereas the thermocouple must make good contact with the end of a thermo-well, the RTD must have a tight fit along its sides.

Flow

One class of flow meter depends on the inter-convertibility of pressure and momentum. In the *orifice* and *venturi* meters the flow is accelerated by passing through a constriction, and the volumetric flow is inferred from the decrease in pressure. With the *pitot tube* the flow is brought to zero at a point (the 'dynamic hole') and the excess of pressure above that at the 'static hole' is a measure of velocity.
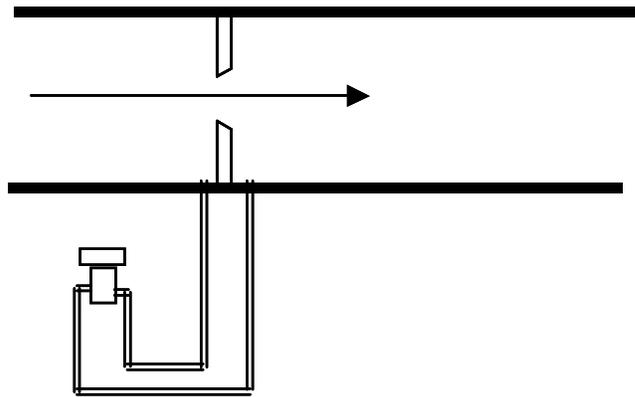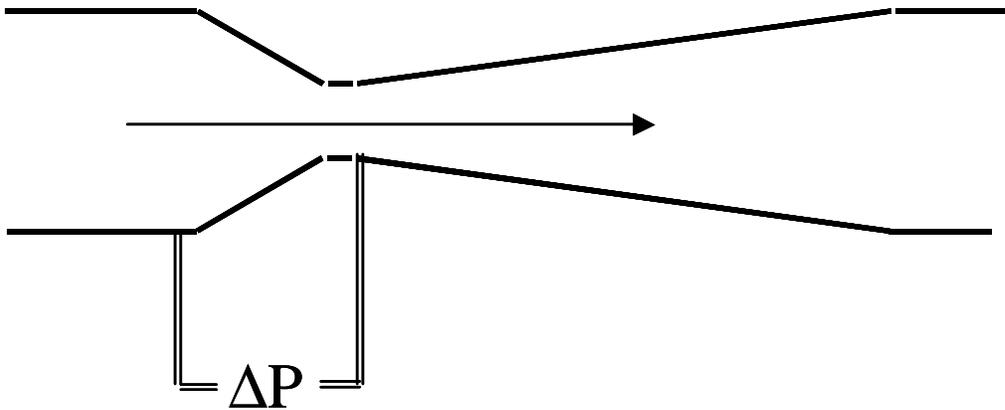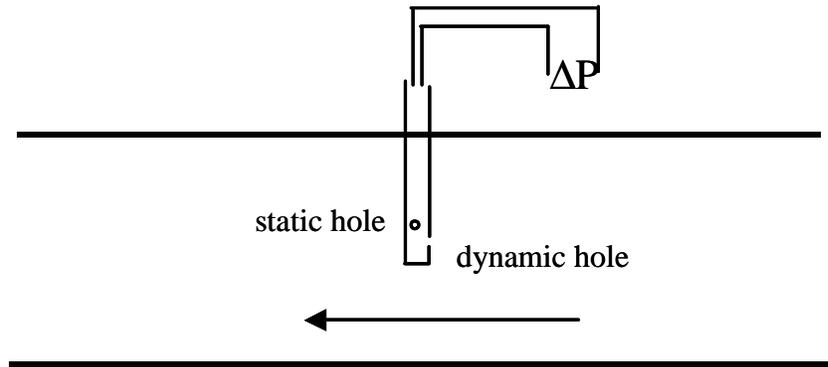
Figure 1.3.5 – Orifice Meter

Figure 1.3.6 – Venturi Meter

Figure 1.3.7 – Pitot Tube

An advantage of all these devices is their relative simplicity and low cost. A disadvantage is the fact that that the signal,

Pressure difference, ΔP is proportional to Velocity^2

This relationship makes the devices inaccurate at flow rates below about 25 percent of full scale.

Relative to venturi meters, orifices and pitot tubes are prone to pluggage

Relative to orifices, venturi meters have much better pressure recovery downstream.

Relative to venturis and orifices, pitot tubes provide only a point measurement.

With all three devices, care must be taken to ensure zero or equal liquid head in the lines leading to the differential pressure cell.

Two instruments that provide a signal that is proportional to the first power of velocity are the *turbine meter* and the *vortex-shedding meter*. Both of them require an external electronic 'pick-up' to count either turbine rotation or shedding frequency.
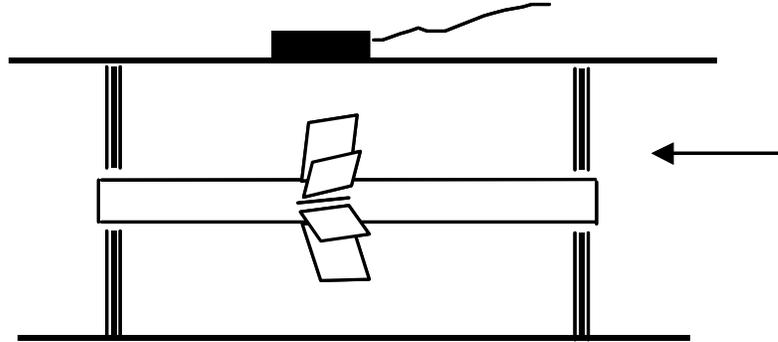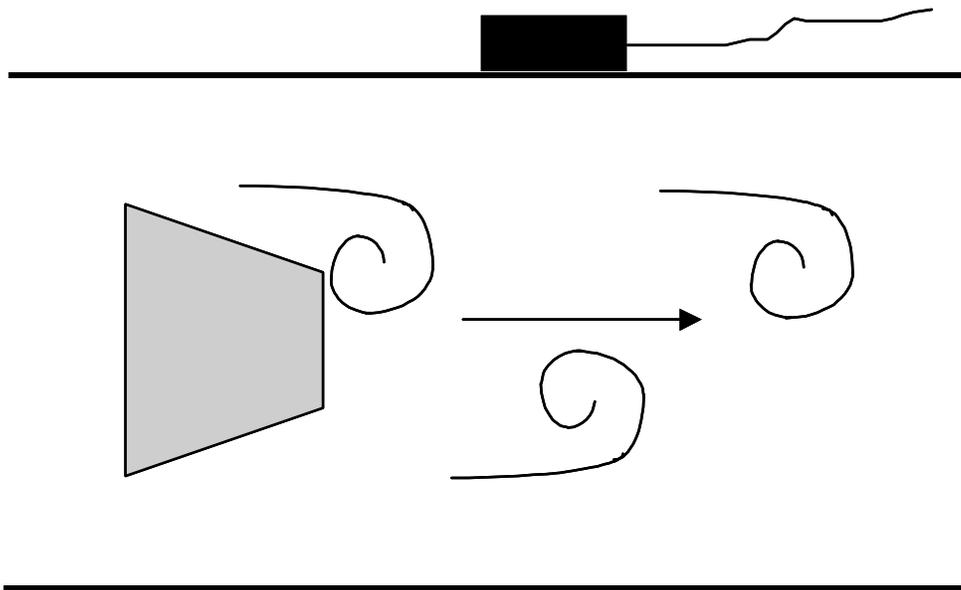
Figure 1.3.8 – Turbine Meter

Figure 1.3.9 – Vortex-Shedding Meter

The advantage of the turbine is good 'turn-down', about 10:1 and good accuracy, about 0.25%.

Disadvantages are susceptibility to fouling, especially of the bearings at each end of the turning shaft.

Advantages of the vortex-shedding meter are turn-down of 25:1, accuracy of 1%, no moving parts, resistance to fouling.

Disadvantages are susceptibility of the electronics to miscellaneous noise and also the requirement of adequate Reynolds number (high flow rate) to ensure that shedding frequency really proportional to velocity.

The *magnetic flow meter* or *magmeter* works on the principle that a voltage (EMF) gets induced in a conductor moving through a magnetic field. In this case the conductor is the flowing fluid, which must possess some conductivity and is therefore restricted to liquids. It is ineffective for de-mineralized water and non-conducting oils.

The advantages are turn-down of 30:1 and accuracy of about 0.75%.

The disadvantages are
- restriction to conductive liquids
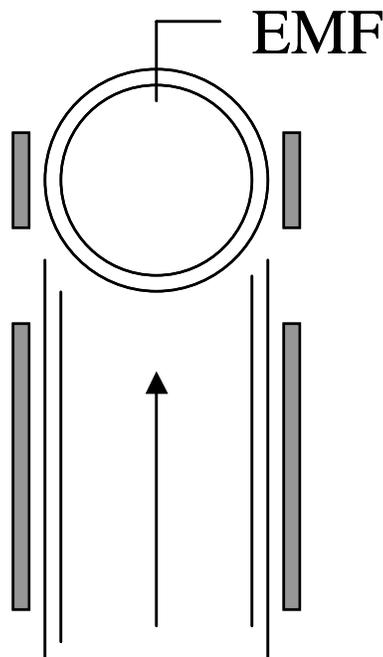- temperature limited to 200, 300 C

Figure 1.3.10 – Magnetic Flowmeter

All of the above devices measure velocity or volumetric flow. The *Coriolis Meter* or *Mass Flowmeter* measures mass. The principle is that of Coriolis which is the rotational form of Newton's second law, Force = Mass x Acceleration. The fluid flows through a U-bend, the center of which is vibrated up and down. The result is a twisting of the U-tube, by an amount proportional to the mass flow rate.
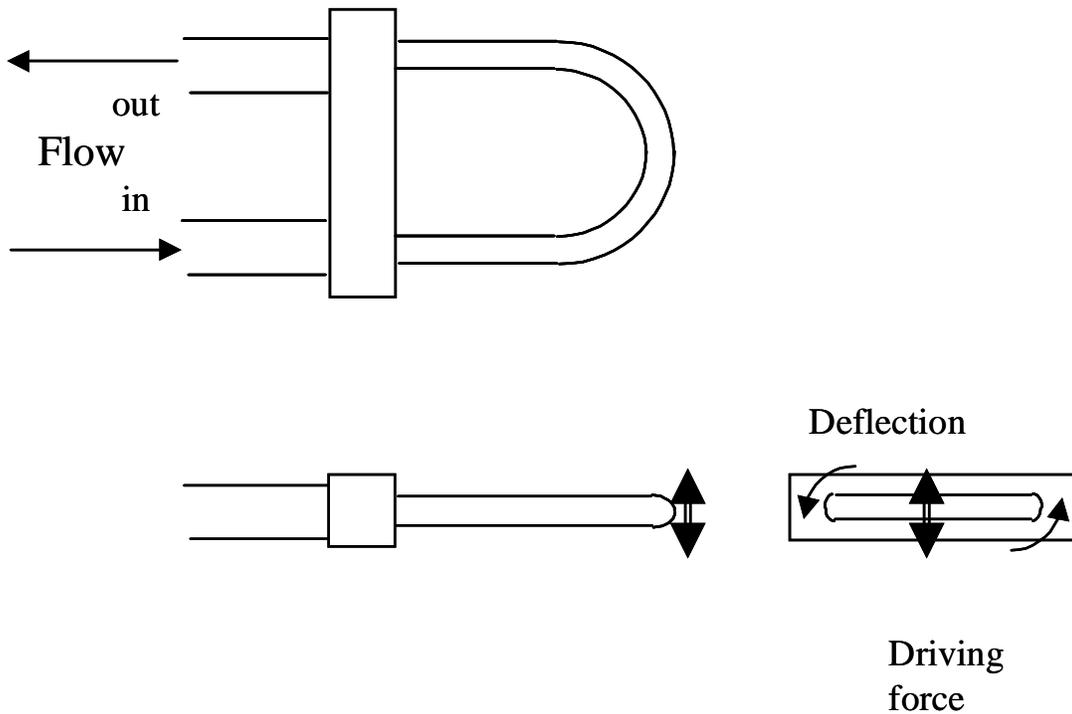
Figure 1.3.11 – Mass Flow Meter
The advantages are
- high accuracy, 0.1%
- high turn-down, 100:1
- applicable to both gases and liquids
- sensitivity to mass rather than volumetric flow
- pressure range up to 1500 psi
- temperature range –240 to + 400 C

Disadvantage: cost

Level

Two of the most common methods of measuring level make use of the static-head pressure that a liquid exerts. In one case the differential pressure between the vapour space above the liquid and a point below the surface is interpreted to determine the level. In the second case the pressure required to maintain a stream of gas into the lower levels of the liquid is measured.
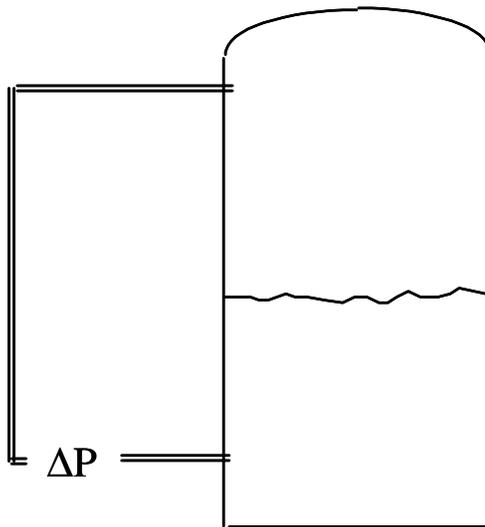

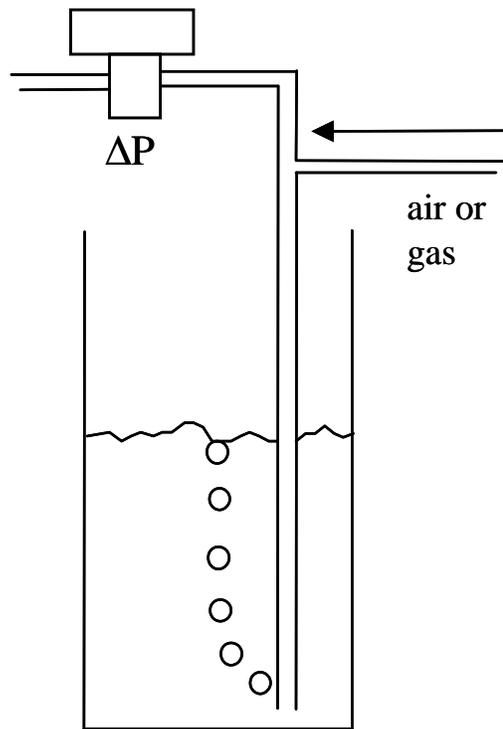Figure 1.3.12 – Liquid Level by Static Head Differential Pressure

Figure 1.3.13 – Liquid Level by Bubble-Tube Pressure

The advantage of these two methods is simplicity and relatively low cost.

The disadvantages are
- direct contact with process liquid
- dependence on knowledge of density
- restriction of bubbler system to vented systems
- necessity to keep lines free of condensate, e.g., condensate
- possibility of plugging

A better indication of the situation in the vessel is obtained by mounting a second measuring point at a different level in the liquid. The difference in static head between the two liquid-immersed points gives a measure of density, which can then be used in interpreting the height signal.

When it is desired to avoid intrusion into a vessel, for instance if it is under high pressure or high vacuum or if the contents are hazardous, it is possible to use nuclear radiation to 'see' what is inside. An expanding beam from a source on one side of the vessel is

partially intercepted and attenuated by liquid in the vessel. The receiver on the other side picks up the residual radiation.
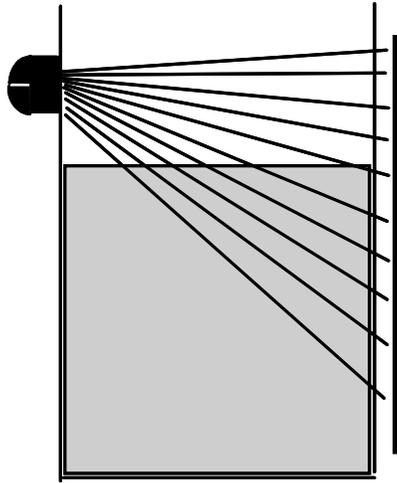

Figure 1.3.14 – Nuclear Level Gauge

The nuclear source is usually the isotope Cesium 137 or Cobalt 60.

The advantages of this method are

- non-contact with process materials
- effectiveness even through heavy-walled vessels
- for existing vessels, no modification required
- low maintenance
- The disadvantages are

- stringent regulation
- potential hazard to health if mismanaged
- relatively narrow range of height
- confounding effect of foam and density

The property of liquid capacitance can be used to detect level. A probe is inserted into the liquid and the capacitance is measured between an outer sheath and an inner core. The signal depends on how much of the surfaces are in contact with liquid.
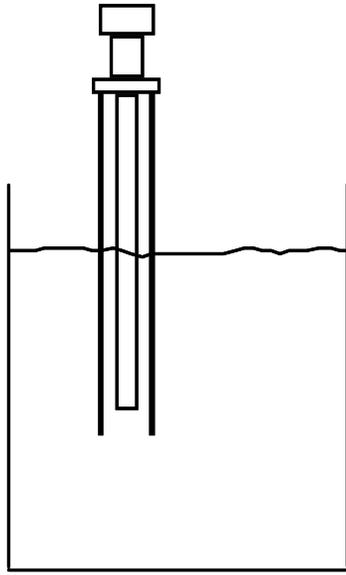
Figure 1.3.15 – Capacitance Level probe

The advantage is that such probes are in common use and therefore economical and well developed.

The disadvantages are

- full contact with process liquid
- capacitance sensitivity to the nature of the liquid
- possibility of surface fouling and therefore change of response

An ultrasonic source and detector mounted in the top of a vessel provides a measurement of level by measuring the time it takes for sound waves to travel down to the surface and reflect back.
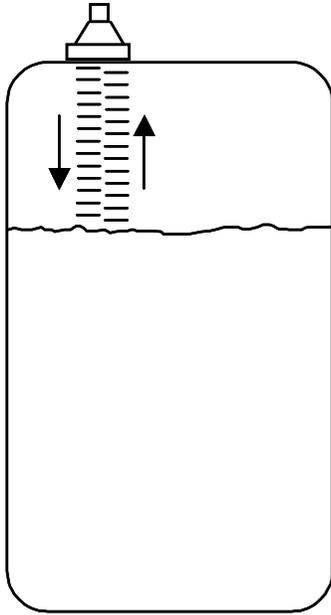
Figure 1.3.16 – Ultrasonic Level detector

The advantages are

- the intrusion into the vessel is at the top, where leakage and contamination are least likely
- there are no moving parts

The disadvantages are

- the signal can be affected by dust, foam, waves and noise
- the upper temperature limit is low
- the instrument must be calibrated against an empty tank.

Pressure

Pressure is measured for its own sake and also as part of the measurement of flow and of liquid level. In many cases the measurement is of the difference of pressure between two points. The instrumentation is based on the movement of a flexible element, for instance a bellows or a diaphragm, which is translated into a gauge reading and/or into a pneumatic or electrical signal for transmission to a controller.
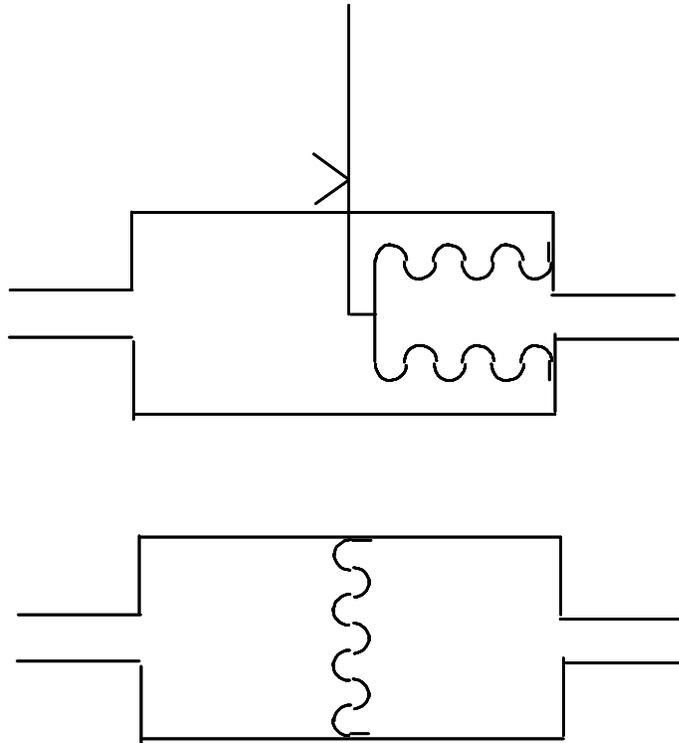

Figure 1.3.17 – Bellows and Diaphragms In Pressure Measurement

If a differential pressure, say between 'A' and 'B', is being measured, the two pressures are connected to opposite sides of the instrument. If a single-point gauge pressure is desired, the instrument is put in 'vented' mode, i.e., one side is left open to atmosphere. If single-point absolute pressure is desired, then one side of the instrument is plugged and evacuated.

Press. 'A'   Press. 'B'

Atmosphere   Pressure
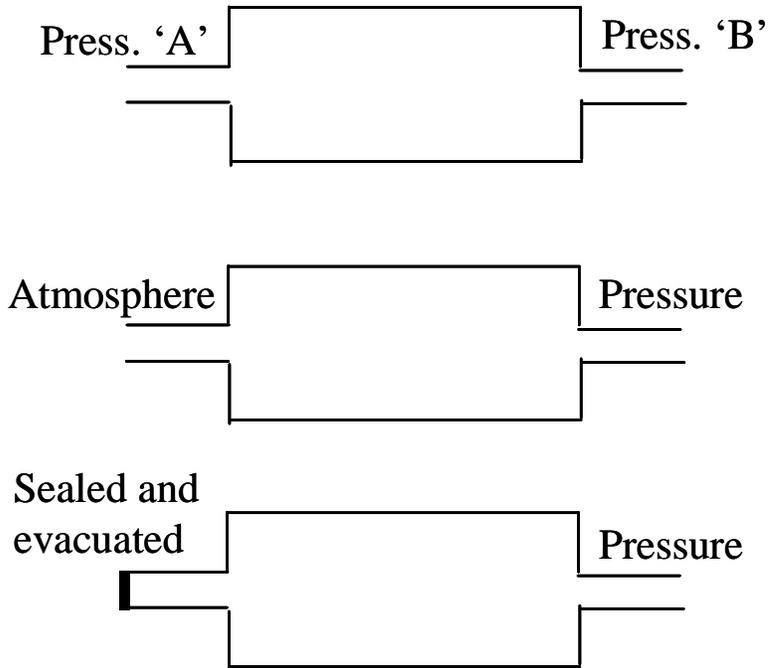
Sealed and
evacuated   Pressure

Figure 1.3.18 – Pressure Device Configured for Differential, Gauge, and Absolute
Measurement

In the case of differential measurement, the actual difference to be measured may be much smaller than the overall pressure of the system. In this case the two sides of the device must be equipped with a balance line which is open when the device is put in service, to avoid having high un-balanced pressure exerted on one side of a delicate diaphragm or bellows. Valve 'C' is then closed
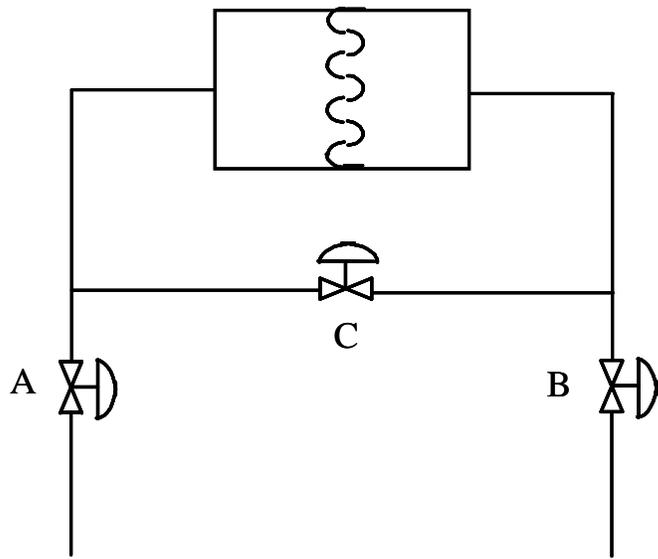After both valves 'A' and 'B' are open..

Figure 1.3.19 – Differential Pressure Device with Balance Line

1.3.3. <u>Interaction of Controls</u>

The simplest configuration in process control is the Single-Input/Single Output loop, as described in Figure 1.3.2. The following illustrations expand on the possibilities.

The two basic PMC units are *indicators* and *controllers* (and combined *indicator-controllers).*
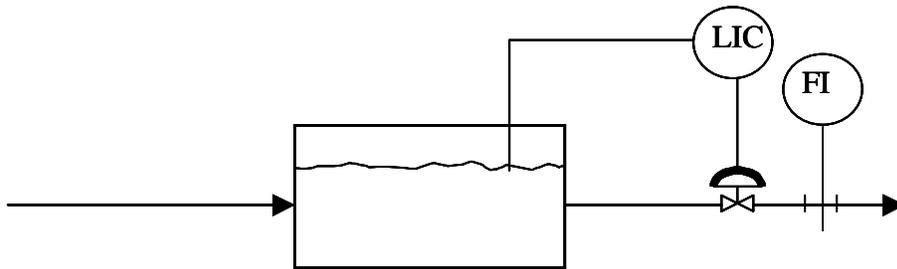

Figure 1.3.20 – Level Control and Flow Indication

Without stating any preference for how the level or flow are to be measured, we are conveying the specification that the level in the tank is to be controlled and we would like to know what it actually is at any time, and that the flow rate out of the tank may allowed to vary but that we would also like to know what it is.

If on the other hand we required a constant out-flow (e.g., as a supply to another step in the process) and were using the tank only as a supply vessel, then the PMC arrangement would change.
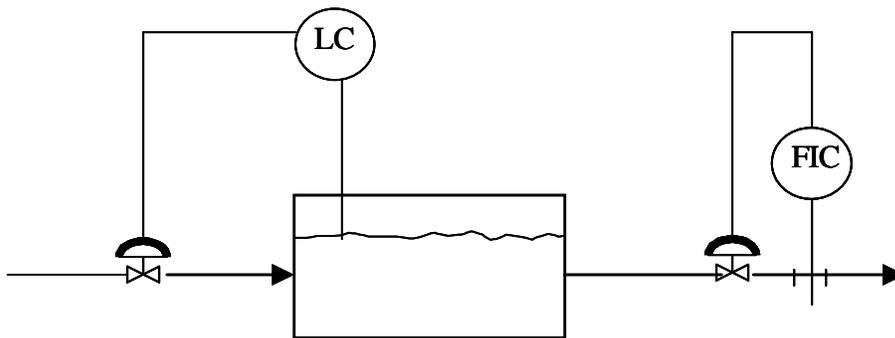

Figure 1.3.21 – Level Control and Flow control

In this specification we are requiring the level to be controlled but are not asking for the

level to be indicated (probably a bad move).

The following schemes show some ways in which control loops are configured to work in tandem with one another.
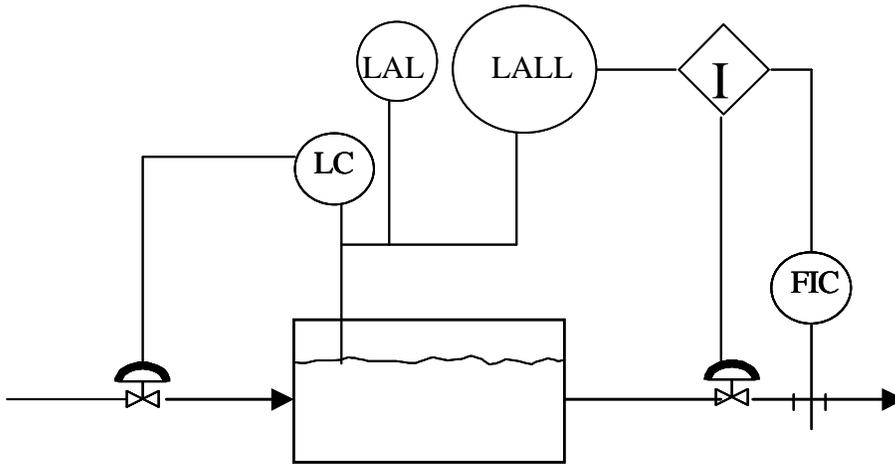
Figure 1.3.22 – Alarm and Interlock

An *alarm* is given if the contents of the tank drop below a prescribed level. An alarm is given and the outflow is shut off via an interlock if the tank level drops further to a second prescribed value.
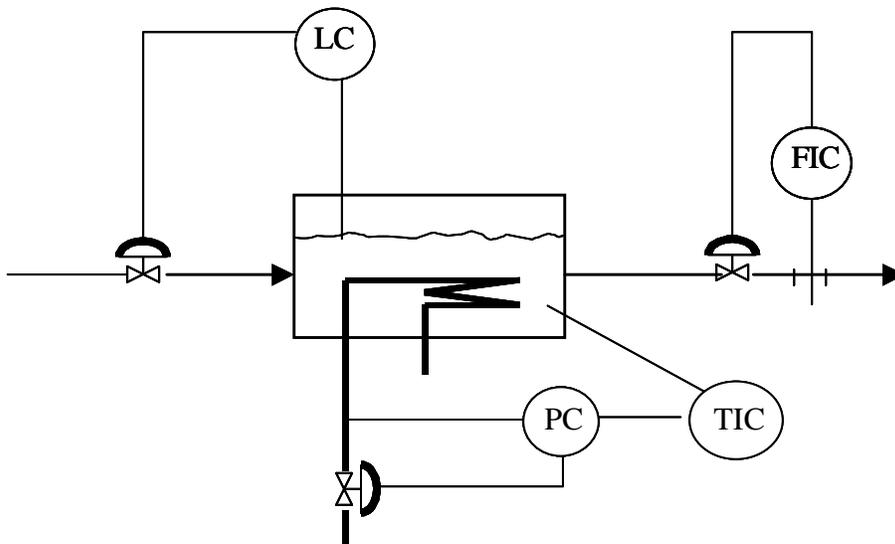
Figure 1.3.23 – Cascade Control

Now we want to control tank temperature as well using a steam heater. Instead of manipulating the steam valve directly to control vessel temperature, we use *Cascade control* and control temperature by manipulating the set point of the steam pressure controller. The benefit is that the control is less sensitive to any fluctuations in steam upstream supply pressure.
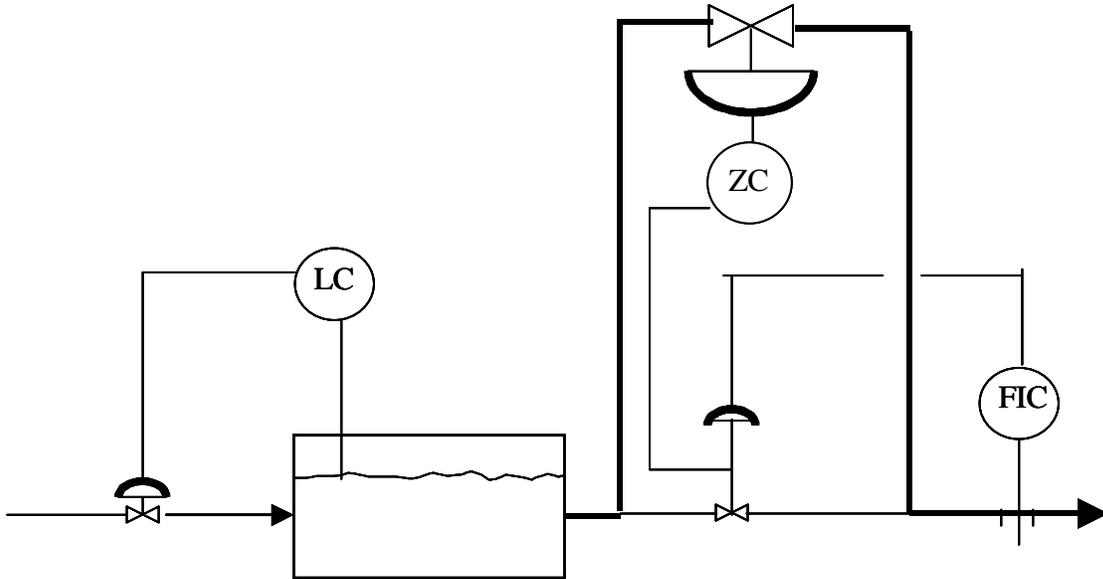

Figure 1.3.24 – Balancing Control

To obtain more precise control of the outflow we provide parallel flows and do the immediate manipulation on the smaller flow. But the range of this valve is small, so we add a slower loop to adjust the large flow to keep the small valve near the centre of its range. This is called *Balancing Control* (in this case *Valve-Position Control)* and we are using the second loop to control small-valve position, denoted by 'Z'.
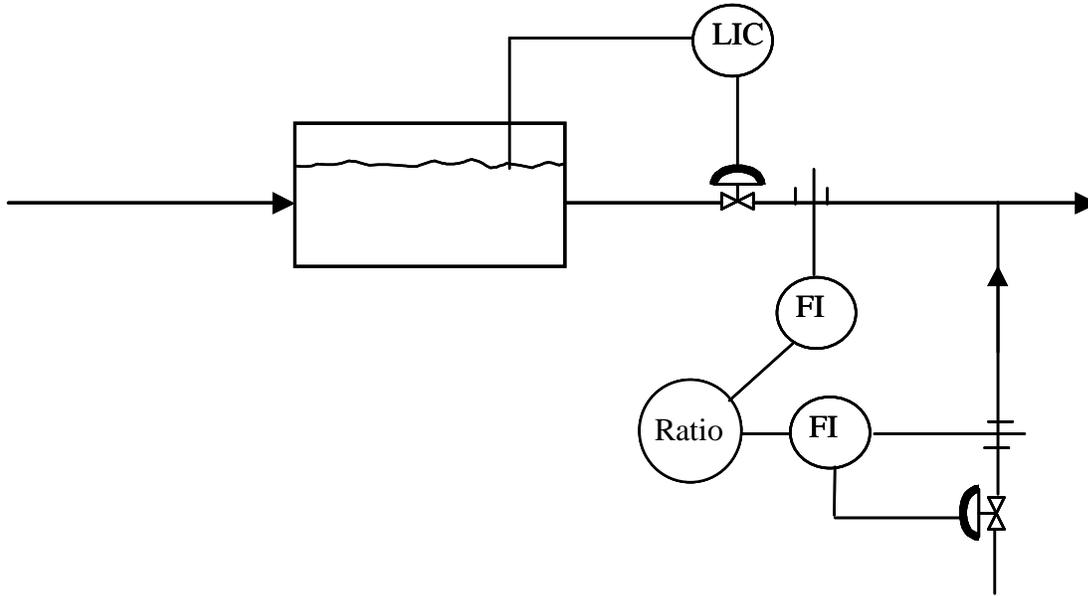
Figure 1.3.25 – Ratio Control

It is desired to keep a constant composition of flow feeding a subsequent unit, so the side-stream flow is kept in constant ratio to the main flow. The technique is called *Ratio Control*. As explained in Section 1.2.1 (and illustrated in Figure 1.2.1) ratio control is a form of feed-forward control.
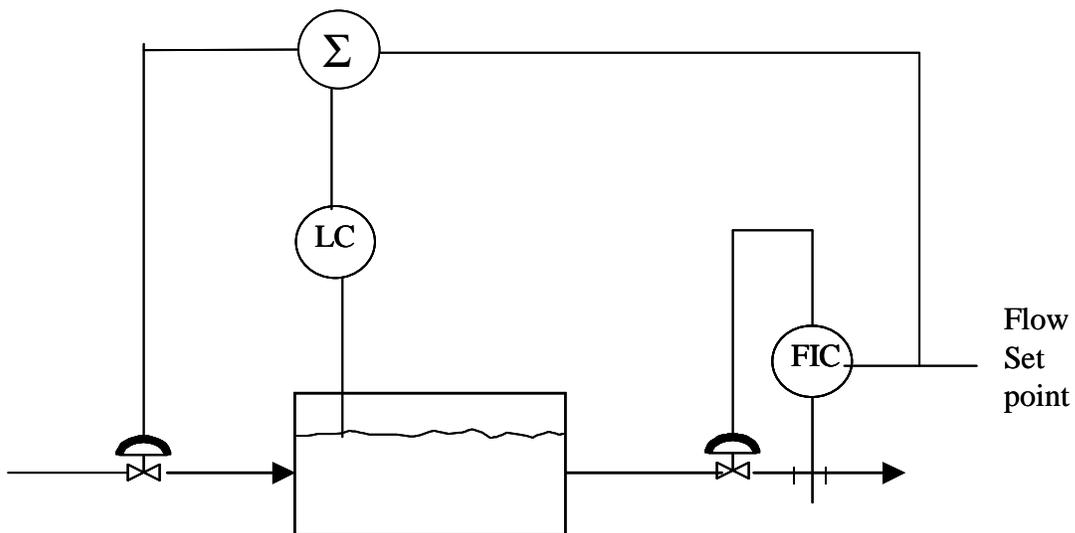

Figure 1.3.26 – Control Using a Calculational Block

In order to maintain closer control over tank level, provision is made to respond early to changes in outflow set-point setting. This is an example of *Feed-Forward Control* even though, in this example, the direction of information flow is backwards. The use of a *Calculational Block* is also illustrated, combining the flow signal with the level signal to determine the correct position of the inflow valve.
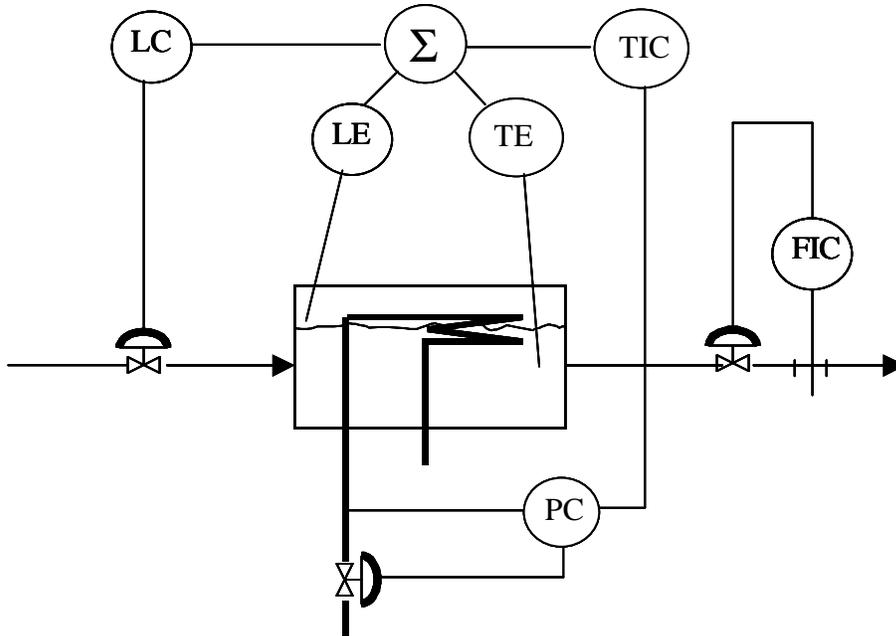

Figure 1.3.27- Multiple -Input/Multiple-Output (MIMO) Control

Sometimes there is an interaction between control loops: in this case the heater may create bubbles which affect level and, at the same time, changes in level may uncover part of the heater and affect the ability to transfer heat. To deal with the interaction, *Multiple-Input, Multiple-Output Control (MIMO)* may be appropriate. This is not a common practice: 95 percent of control loops are said to be *Single-Input, Single-Output (SISO)*.

As we see, there are many avenues for reaching the accuracy and tightness of control that is required. For the operator of the process, it is important to know what is controlling what.

1.3.4. Good Control and Bad Control

Good and bad control is a very simple concept: how close does the *controlled variable* stay to the *set point*. It is understood that *CV* will always deviate somewhat because this is the nature of feedback control: a deviation is required before control action is taken. Only if disturbances are absent and if the set-point is constant will the *CV* stay exactly at the set-point.

The key questions are

- how close is it really necessary for *CV* to be to its *SP*?
- how reliable is the measurement of *CV*?
- how fast and with how much cycling does *CV* return to its set-point after a disturbance or a set-point change?

The answer to the first question depends on the nature of the process. Sometimes the limits are wide: for instance the level in a holding tank or surge tank might satisfactorily be allowed to range between almost empty and almost full although half-full may be the preferred level, i.e., the set-point.

If the operator sees or suspects that the answer to the second or third question is not satisfactory then there are four areas to examine.

The process itself may be difficult to control. It may be inherently unstable, in which case the control system has to deal with this fact. On the other hand it may be suffering from disturbances that are correctable. For instance, it may be hard to control temperature in a vessel if a near-by door is opened from time to time to a winter wind. Sometimes control loops 'fight' each other: in the example of Figure 1.3.26 showing Multiple-Input/Multiple-Output control, if the level control and temperature control had been done separately (in two SISO loops), the interaction would have made one or both controls less than optimum. In Figure 1.3.27, below, if level control in the downstream vessel is aggressive, i.e., calling for large fluctuations in inflow, then it may be difficult to control level in the upstream vessel.
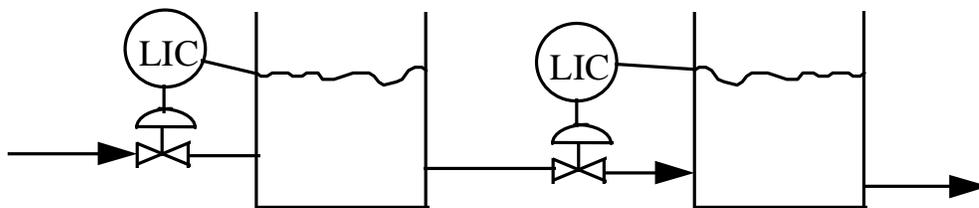

Figure 1.3.27 - Interacting Level Controls

Another situation where control may be difficult is for a loop that receives its set-point as the output from another controller. For instance, it may be the 'inner' loop in a cascade or balancing configuration. If the outer loop is tightly tuned, with high gain and rapid reset, it will cause the inner loop to work excessively.

The sensor, i.e., the measurement of the controlled variable, may be the source of trouble. There are several possibilities.

- It may be broken. This condition can manifest itself as a constant signal or as a rapidly fluctuating signal and is usually quite obvious.
- Its accuracy may be compromised: a thermocouple may be improperly installed in its thermowell or a differential-pressure cell may have legs unevenly filled with liquid or a flow meter may be operating at the low end of its range.
- It may be improperly calibrated or it may have been replaced by a unit with different calibration.

It may be poorly located relative to the point at which the manipulated variable acts. In Figure 1.3.28 the presence of the vessel between the point of additive-addition and the point of measurement makes it difficult to achieve good control ('A' denotes an analytical measurement).
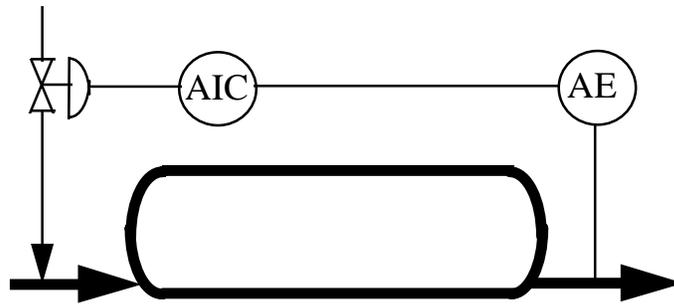

Figure 1.3.28 - Difficult Composition Control

The final control element may suffer from mal-function. The most common final element is a valve, which may suffer from

- stickiness, causing it to move in jumps rather than smoothly with its input signal
- hysteresis, causing it to have, for the same signal, a different opening depending on whether the opening is increasing or decreasing.
- saturation.

The first two conditions can be checked by taking the valve off-line and observing its position as the signal to it is changed manually. Proper maintenance can usually correct the problem.

Saturation is the condition where the element reaches the limit of its ability to affect the process. For instance, if the controller is calling for a flow that drives the valve wide open or to a state where further opening does not produce significantly more flow, then control is not achieved. This situation arises typically where, either the valve was under-sized in the first place, or an increase in production rate causes a valve to become under-sized.

The control system itself may be the cause of poor control. Control action is one of the three actions that affect the controlled variable, the other two being external disturbances and re-setting the set-point. Poor controller settings can make the controlled variable fluctuate and even make the system unstable. This happens when the process gain is set too high or when the reset (integral) action comes on too quickly. On the other hand, low gain and low integral action let the controlled variable wander loosely around the set point. Controller 'tuning' is the art of setting gain, reset and derivative action to the correct intensity for adequate control.

Figures 1.3.29 and 1.3.30 show the time-wise response of the *controlled variable* to a disturbance with fast-acting and slower-acting control. The firmer gets the *CV* back to its set-point more quickly but in an oscillatory fashion with some significant under-shoot. As long as the under-shoot can be tolerated this is a very satisfactory control action. If the oscillations persist and especially if they grow, then the control action is less ideal and possibly even destructive.
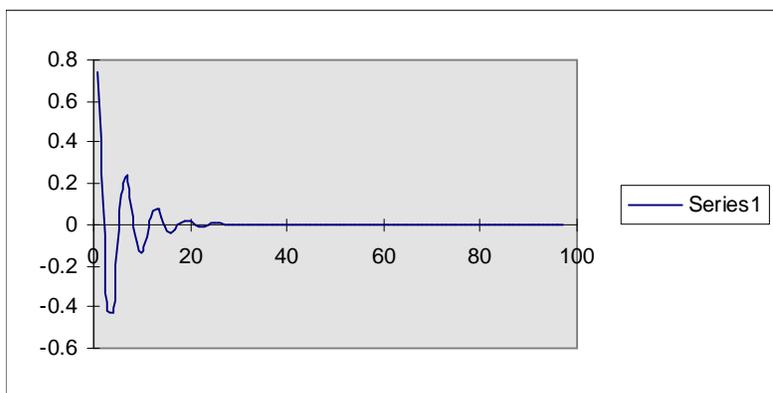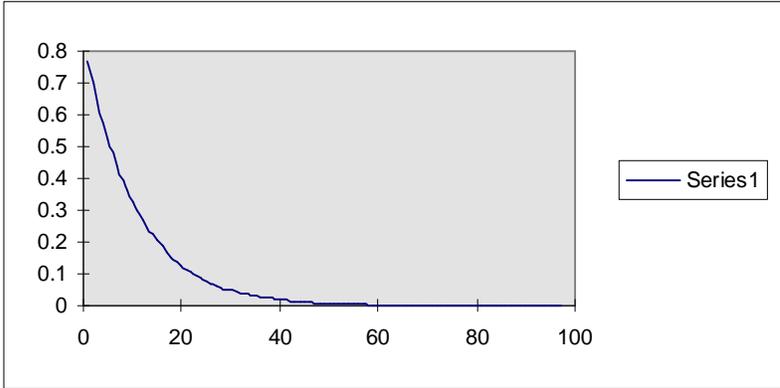

Figure 1.3.29 – Rapid-Response Control

Figure 1.3.30 – Slow-Response Control

1.3.5. <u>Distributed Control Systems</u>

A short history of the development of process control helps to put the current state in perspective.

The earliest process controllers were human beings, who read the gauges and turned the valves. For instance in the manufacture of nitroglycerine (Figure 1.1.9), in the 1950's the control was still carried out manually. To save manpower, some plants were configured to bring the indicators and control elements into a central location: doing this required routing the actual process piping through that location. As automatic controllers became available they were still often grouped together in panels and in control rooms but the measuring elements and final control elements stayed in the field. Rather than routing the process piping through the central location, it was necessary only to bring the information lines (pneumatic or electric) to the panel.

The earliest controllers operated in Single-Input/Single-Output mode (and most still do) but more sophisticated techniques like cascade were introduced. The progress in hardware also depended on mathematical development of control theory. Both the hardware and the math benefited greatly from the advent of computers.

One of the first computer-based concepts in process control was *DDC* or *Direct Digital Control*. Because the individual pneumatic or electronic controller is designed to operate according to a mathematical expression (i.e., the control equation), it seemed natural to turn this function over to a large central computer, since mathematics is a thing that computers do very well. The computer could handle all control functions in a plant and could be programmed with control algorithms much more sophisticated than those of a pneumatic device. But there was reluctance to put all the control 'eggs in one basket', since a computer sometimes fails and a whole plant might fail along with it.

A less daunting approach was *Supervisory Set-Point Control* in which traditional controllers were used but their set-points were set by a computer. This approach is commonly used today in labs, pilot plants and other small operations, with a Personal Computer (*PC*) supplying the calculational power.

In the mid-1970's *Distributed Control Systems* were introduced (i.e., Honeywell TDC2000), which were centralized but which had multiple calculational units rather than just one. If one unit failed there was a 'hot' backup to step into its place. *Programmable Logic Controllers* came along in the 1980's. Initially *PLC*'s were aimed at discrete control (controlling events rather than variables), but it is said that the *DCS* and *PLC* concepts have gradually merged.

At the same time 'smart' individual controllers have been developed with on-board calculational power: they work alone or in tandem with a *DCS* or *PLC*. But the traditional PID controller is still very much in use, with its simple capabilities of set-point setting, tuning adjustment, manual / auto switching, and display on request of input and output signals and error.

Computerized central systems like *DCS*, *PLC* and *PC*-based configurations provide a wealth of capability to help the operator run a good process:

- simple and sophisticated control algorithms
- reliability through redundancy
- programmability of varying difficulty
- graphic display of current information
- tabular and graphical records of past operation (trending)
- alarms
- logging of events and alarms
- recipe handling.

All that remains is for the operator to make use of these modern-day features.