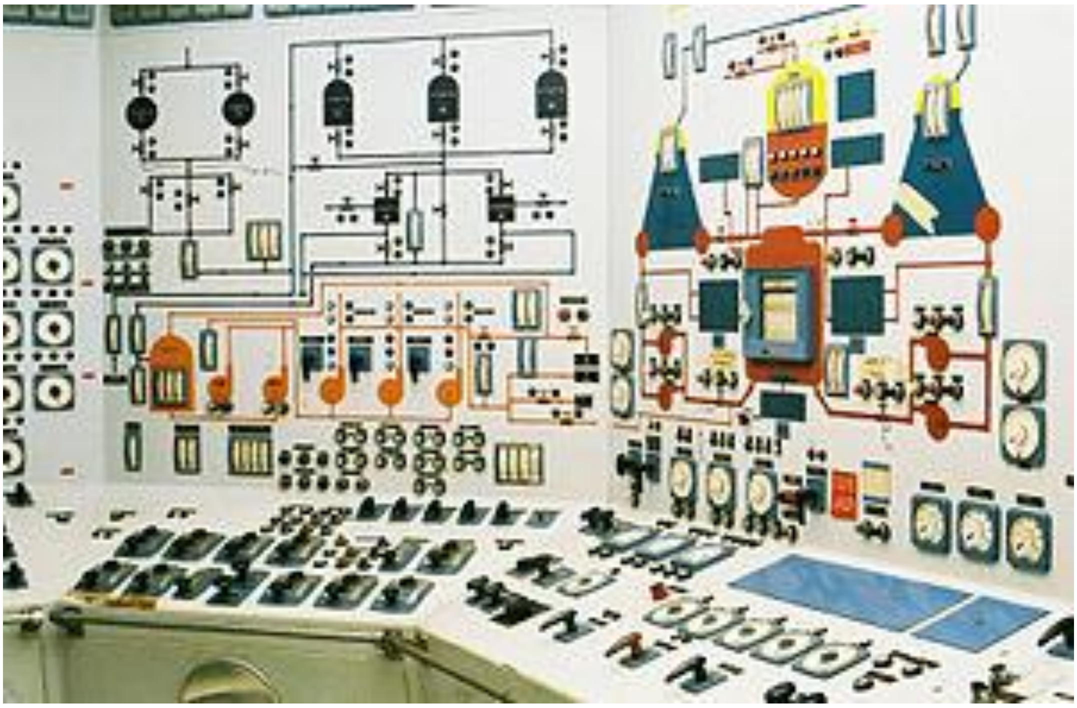
PROCESS CONTROL



PROCESS CONTROL.

10.1 Introduction.

As previously mentioned, for the successful operation of most chemical plant operations and manufacturing processes it is essential that the variables, referred to as process variables must be maintained within predetermined limits to ensure safe and economic operation of the plant.

In the early days of industrial development, successful operation was largely due to the skill of the process operator, however individual judgements varied, thus causing irregularity in product and unreliable operation of the plant hardware including control and monitoring instrumentation.

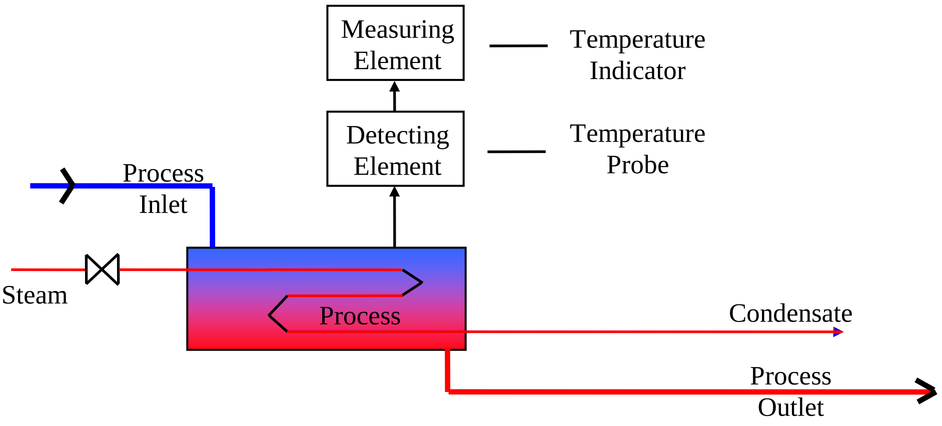
Due to the advances made in technology it is now possible for a large chemical plants to be completely automated, and the process operator need only make minor adjustments such as increasing or decreasing the plant load requirements and start -up or shutdown procedures. Instrumentation does the rest.

All control systems must conform to a standard. British Standard 1523 is the one which relates to process control, and therefore all definitions and examples which follow in this section conform to this standard, either in part or whole.

10.2 Open loop control systems.

An *open loop* control system is one where a measurement is taken of the process condition, but this has no direct effect on the control of the process.
A simple example of this would be a simple temperature gauge which is merely used to indicate the process temperature, any controlling measure would need to be carried out manually by a process operator.

A typical example is shown below:-



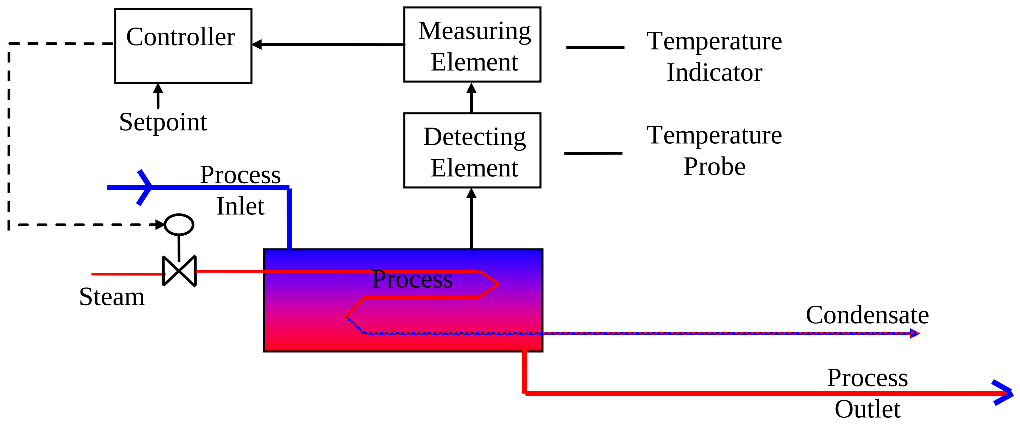
To change the temperature the process operator would have to manually adjust the position of the hand valve until the desired temperature was reached.

Another simple example of a open loop control system is a car wash. Once the cleaning cycle has started, the system merely carries out a series of actions until the end of the cycle, what the system can't detect is the final product, the cleanliness of the car. If the owner wasn't satisfied then they would decide whether to re wash the car.

This type of system therefore is impractical when it comes to accurately controlling the process, a method is therefore required to not only sense the conditions but also to act upon them automatically and this is known as a closed loop system. This type of system could also be referred to as a *manual control* system.

10.3 Closed loop control systems.

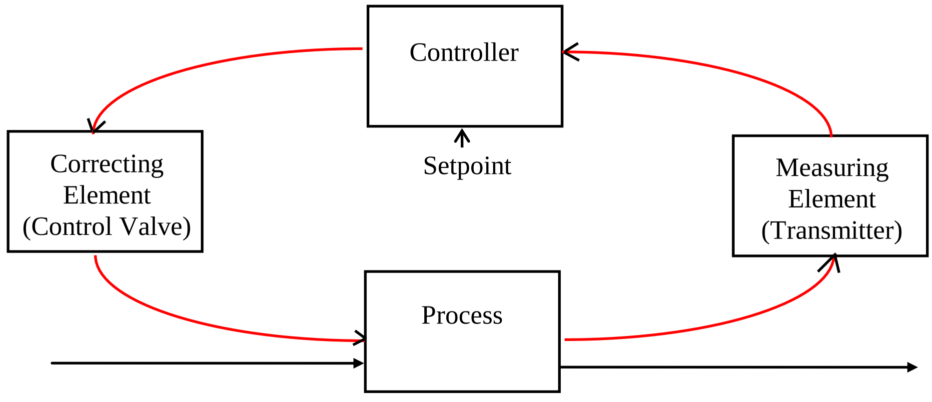
The closed loop differs from an open loop system in that has the addition of a device known as a controller. The measured process condition information (measured variable) is fed into the controller which acts as a comparator, if there is a difference between what is desired (setpoint) and the measured variable, the controller will then create a corrective output which is fed to the control valve directly, and thus automatically respond to changes in process condition. Now all the process operator has to do is tell (programme) the controller with the desired value. As the next diagram shows:-



This type of system is similar in operation to the autopilot in an aircraft. The pilot merely tells the aircraft where to fly, how high and how fast, instrumentation does the rest.

Another point to notice is that the measured variable is taken after the control valve, therefore the information is being fed backwards and so this type of system is often referred to as a *feedback system*. This system could also be referred to as an *automatic control* system, because it is capable of controlling and correcting itself automatically. The next block diagram shows this:-

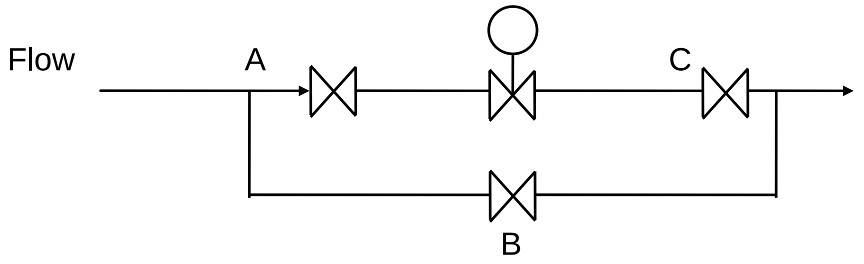
A typical feedback control system:-



The next point to consider is what would happen if the measuring element developed a fault. Quite simply the controller would respond to the error as if it where the actual condition and this is impractical as will be explained in later sections.

10.4 Hand Bypass Control.

This mode of control is required when maintenance work is to be carried out on the control valve, where disturbances would causes problems. A bypass system would then be put in place around the valve to avoid shutting that section of plant or pipe down merely to carry out maintenance checks. The next diagram shows a bypass network:-

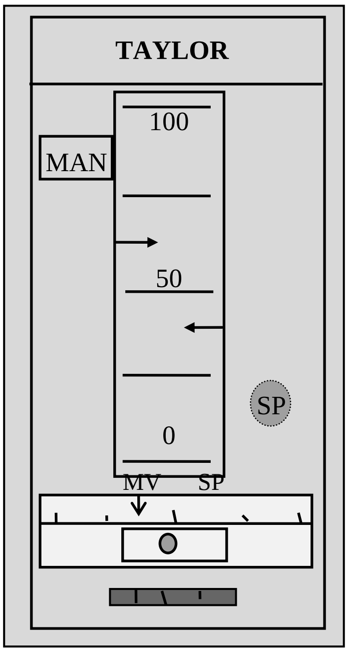


To operate the bypass, valve B should be opened first. Then isolate valves A & C. Any pressure needs to be carefully vented and the control valve can either be removed or stroke checks carried out. Returning the system to normal is the reverse of the isolation procedure. *As a guide, remember Bypass first and last*.

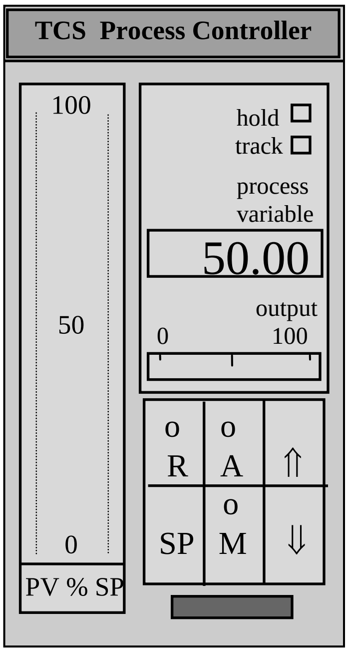
10.5 The Controller.

The *Controller* is described as a device which *measures* an input (measured variable), *compares* this to a setpoint, *computes* the difference, and generates a *corrective* output.

The diagrams below show what a typical Pneumatic and Electronic controller would look like:-



PNEUMATIC



ELECTRONIC

Each of the above controllers have a facility to indicate the measured variable, as either MV or PV, there is also an indication of the setpoint SP, this may be adjusted by either turning the Knurled adjustment knob on the pneumatic, or by pressing the SP button and the raise or lower arrow. The manual output may by adjusted by putting the controller into manual using the switch labelled MAN or M, and then turning the adjuster or pressing the raise lower buttons. Auto may be selected by pressing the A button, or moving the MAN switch to AUTO.

An indication of the output can be seen on the output indicator.

The controller has 2 function areas, the first is the *manual* station in which the controller ignores the input and acts similar to a regulator merely providing a manually adjustable output signal.

The *Automatic* station is the section which acts upon the input to provide a continuous corrective output. There are a number of ways in which the controller responds to the input and these are often known as the modes of control. There are 3 modes of control, these are Proportional Action, Integral

Action and Derivative Action. The following are examples of the controller modes combined together:-

P	Proportional only
P + I	Proportional + Integral
P + D	Proportional + Derivative
P + I + D	Proportional + Integral + Derivative

Notice that all options have proportional control action.

If the measured variable is removed the controller can not act in its automatic mode, thus return it to manual or it will give a constant extreme output by responding to the lack of input as though it was zero. Some electronic controllers go into what is known as forced manual.

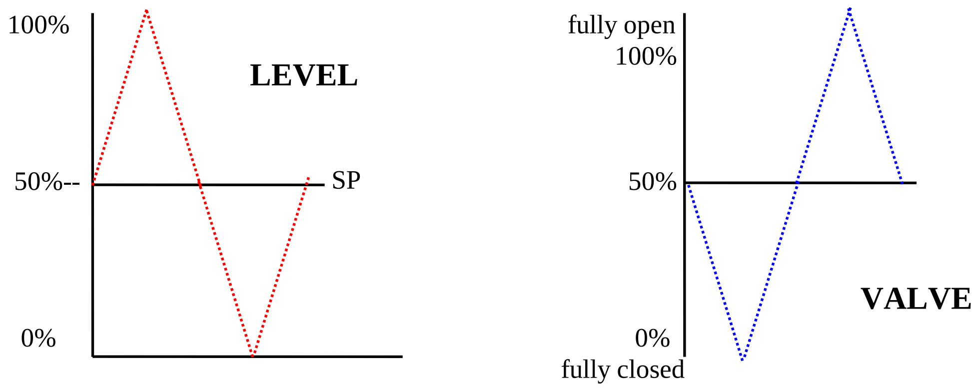
10.6 Bumpless transfer.

Bumpless transfer is the name given to the desired effect when switching from auto to manual. With pneumatic controllers, the output of each station may be different. If there is an in-balance when switch-over takes place, a bump is put onto the output that would cause a process disturbance. The size of the in-balance would determine the size of the effect, and in some circumstances may make regaining control difficult.

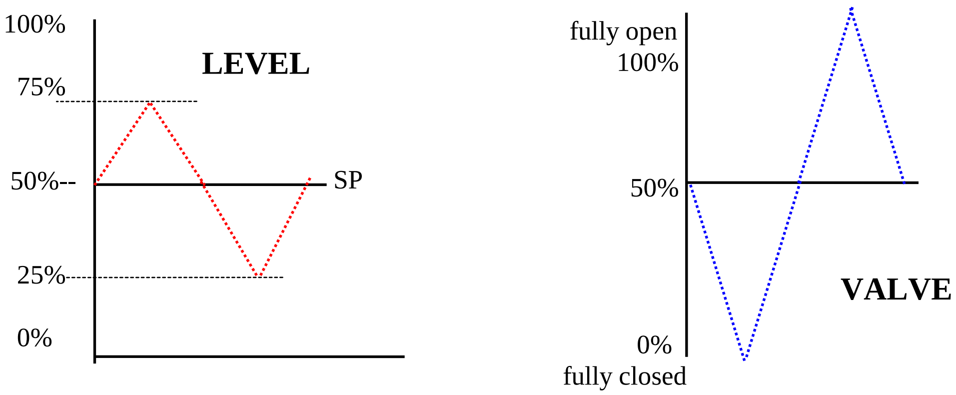
It is important therefore to ensure that the output of each station is matched before switching. With electronic controllers this normally happens automatically however with pneumatic controllers the outputs need to be balanced before switch over.

10.7 Proportional Control.

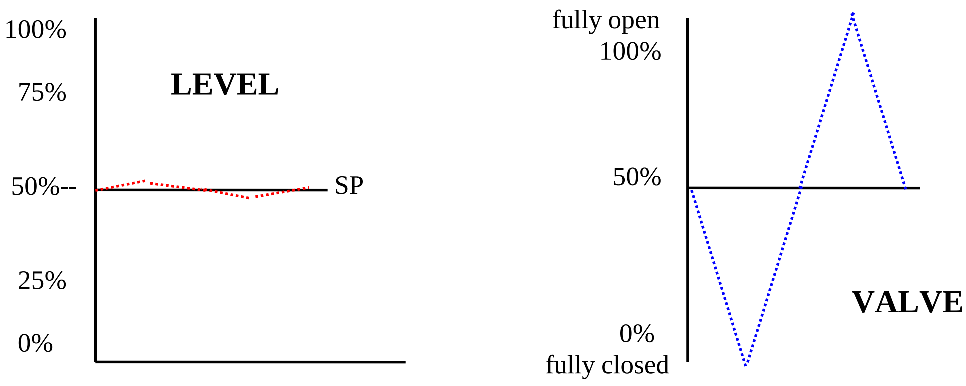
This is where the output of the controller is proportional to the input, ie:- 1 unit in gives 1unit of change out, so for example if the level in a tank rose by 10%, the control valve would close by 10%. An example of this could be when the driver of a car goes into a skid and has to apply what is commonly called opposite lock, the driver does this until the car travels in the way they want to go. The next example shows a graph of a level in a tank with the level being expressed as a percentage, the setpoint is set at 50%:-



From the diagram above you can see that as the level rises above the setpoint the control valve closes by an equal amount. This system is said to have a *proportional band* of 100%, because the level has to go from empty (0%) to full (100%) to get the valve to move by the same percentage amount. If the supply to the tank was quite fast and therefore the changes were rapid, we would need the system to respond differently, as the next diagram shows:-



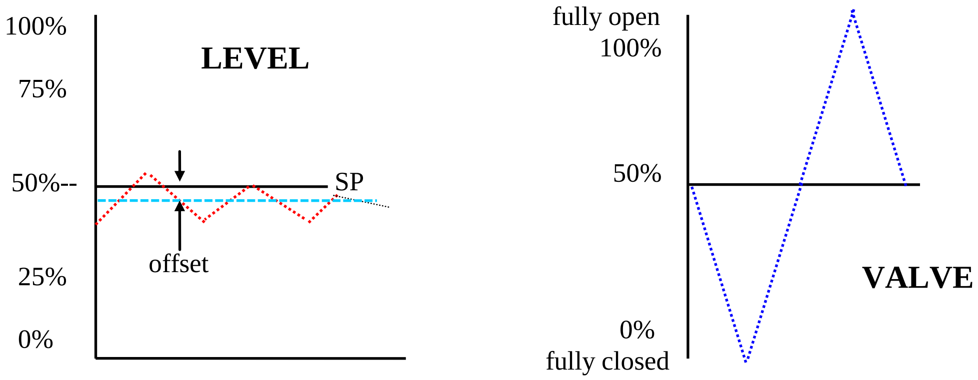
The diagram shows that when the level in the tank reached 75%, the control valve had fully closed, therefore showing that in input change of 50%, created an output change of 100%, or an input change of 1 unit gives an output change of 2 units. This system is said to have a proportional band of 50% or alternatively a *gain* of 2. The narrower the band is made the more reactive the system becomes until a point is reached were the slightest change from setpoint creates a 100% change in valve position, this is known as *On/ Off*, or 2 position control where the proportional band is 0. the next diagram will show this:-



This time it can be seen that the proportional band is only about 4%, so now 1 unit input creates an output of 100%, a gain of 25.

Other typical examples of where 2 step control is used is in batch process operations, and a simple household example is the thermostatically controlled central heating system.

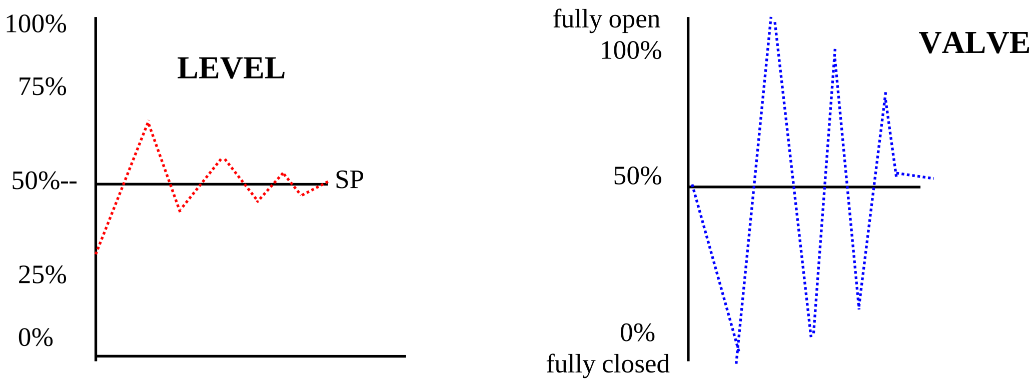
Unfortunately the figures above would only appear in an ideal system. Because the plants are large the distance between the measuring element, controller and valve can be quite large as well time delay can affect the figures, also the resistance (pipesize) and capacitance (volume) of the pipelines and vessels is not always consistent. As a result the measured value may be slightly away from the setpoint when the system steadies itself, or alternatively at the centre of the oscillations. This deviance is known as *offset*, a sustained error/ position away from setpoint.



In some systems a small degree of offset may be acceptable. However in others it may not and therefore needs to be reduced or eliminated. This is done by adding a second mode of control to proportional. The mode to be added is *Integral Action.*

10.8 Integral Action.

Offset could be likened to the driver of a car approaching a bend without breaking, the driver is able to negotiate the bend but unfortunately has to use the other side of the road to do so. Therefore in order to remove the offset, Integral action is added to proportional action. By an internal process of repeating the control actions the offset is removed and the measured variable returns to setpoint. Hence on a controller the integral action time is usually expressed as repeats/ minute. The next diagram shows the effect of integral action:-



Unfortunately what this system does not do is take into account the rate of process change. For example temperature processes tend to be slow, whereas pressure, level and flow tend to change more rapidly. Because of this it is possible to lose control quite quickly with a fast reacting process. To overcome this problem the final mode of control is added , this is called *Derivative action.*

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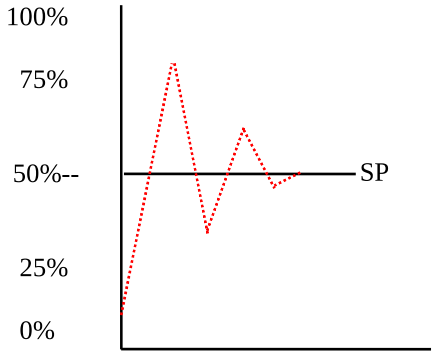
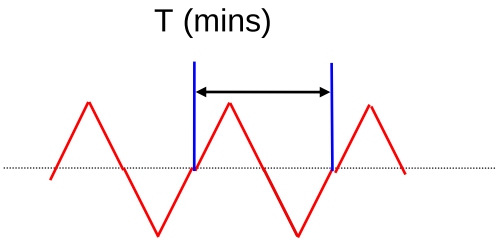
10.9 Derivative action.

This could be likened to heavy breaking when a driver approaches a corner far to quickly and has to "slam- on", failing to do so would result in not negotiating the corner. Derivative action responds to the rate of change by acting faster than it, by first producing a large overshoot. Derivative is often referred to as *rate* or *preact* control. If used correctly, combined with the other 2 modes of control, derivative should be able to return the process to a steady state at setpoint within 4 half cycles. This is known as *1/4 decay*. The next diagram shows the effect of all the control actions working together in response to a step change:-



The above diagram shows the combined effect of PID acting together in response to a process change, and represents the ideal response.
A mathematical way of obtaining this response is known as the Ziegler - Nichols method of loop tuning. The following procedure shows how this may be achieved:-

Set gain to minimum (maximum proportional band).
Integral action time to 1.
Derivative action time to 1.
Introduce a small disturbance by moving the setpoint.
Increase the gain in small steps(ie:- reduce P.B by 10% each time) until the measured value oscillates evenly at a constant amplitude. As shown below:-



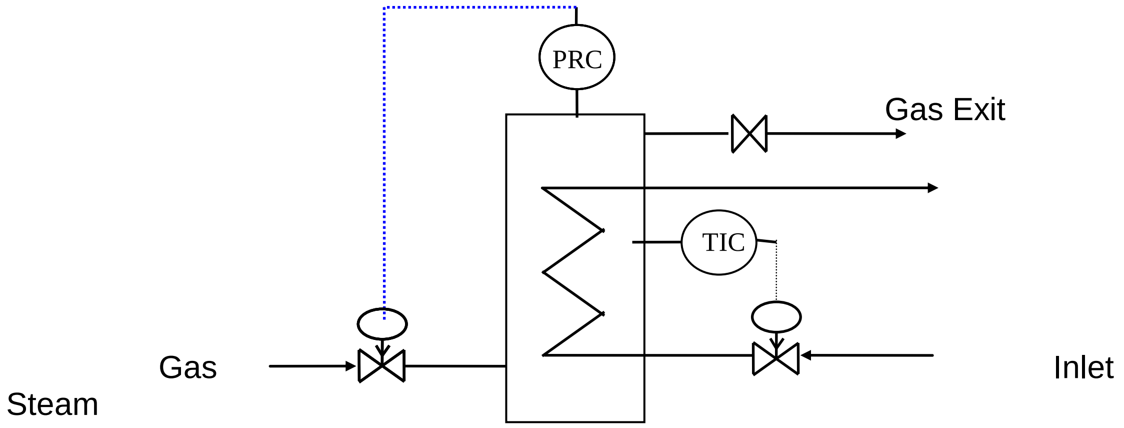
Once the oscillations have stabilised the following settings may be applied dependent upon the number of modes of control required or available:-

Type of Control	PB	I.A.T	D.A.T
P	2PB	0	0
P.I	2.2PB	0.8T	0
P.I.D	1.67PB	0.5T	0.12T

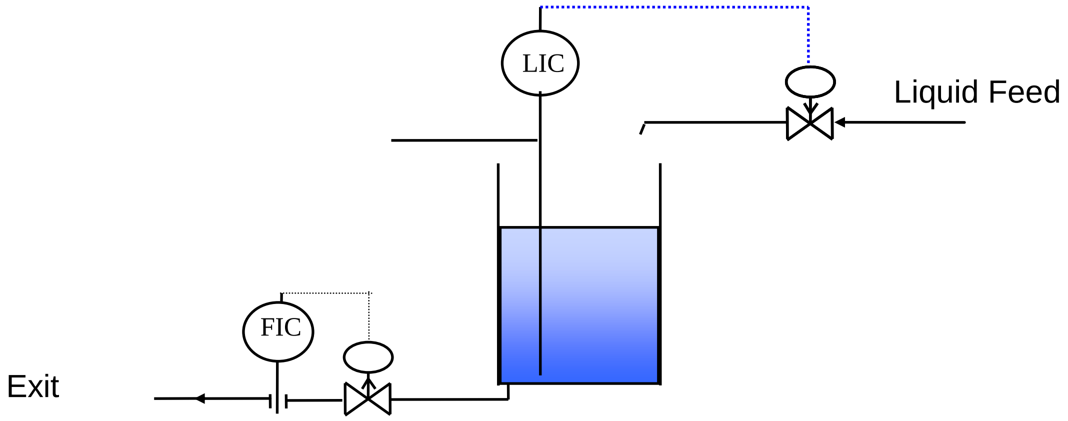
A well tuned process can substantially improve plant and process reliability and help to produce more consistent products.

So far we have looked at single loop control systems, with single loop controllers. The following examples show this:-

Pressure and Temperature Control:



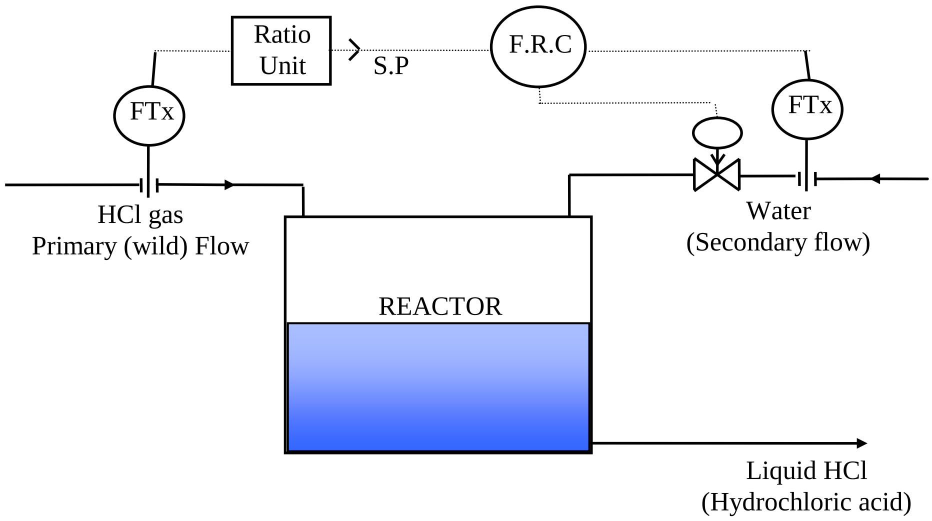
Level and Flow Control:



10.10 Multi Loop Control systems.

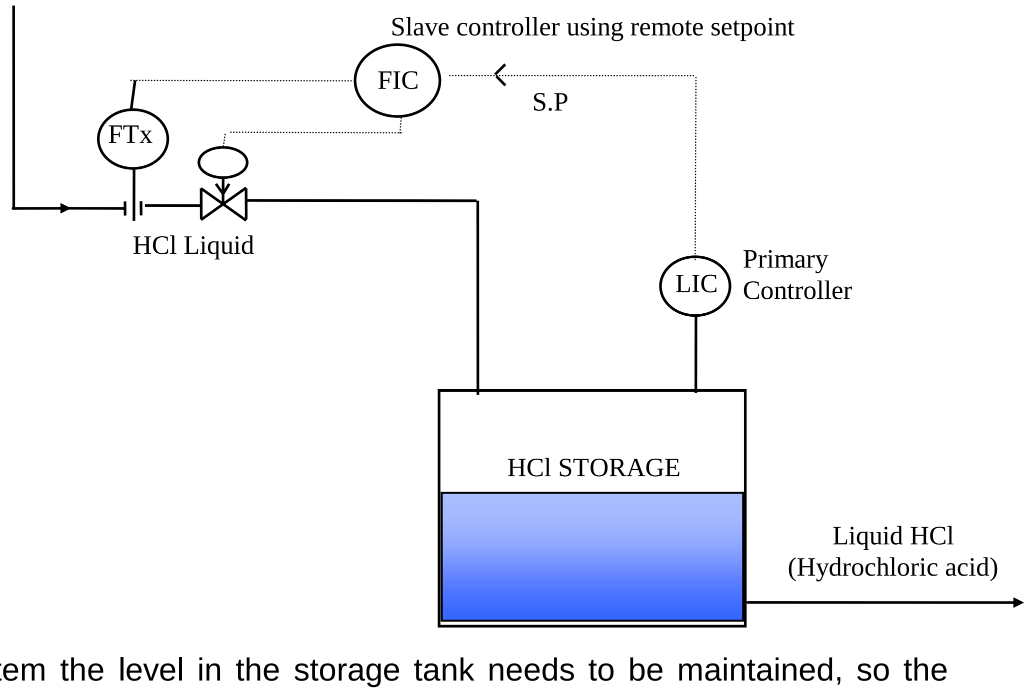
The are occasions when a control loop may require more than 1 controller, and therefore more than 1 input. These are known as ratio or cascade control systems.

The diagram below shows a simple example of a *Ratio* control system:-



In the above example, as the HCl gas flow changes the water flow is ratio'd to it to maintain a percentage balance in the reactor so that the make-up of the Liquid HCl out remains the same.

The next diagram shows a *Cascade* control system:-



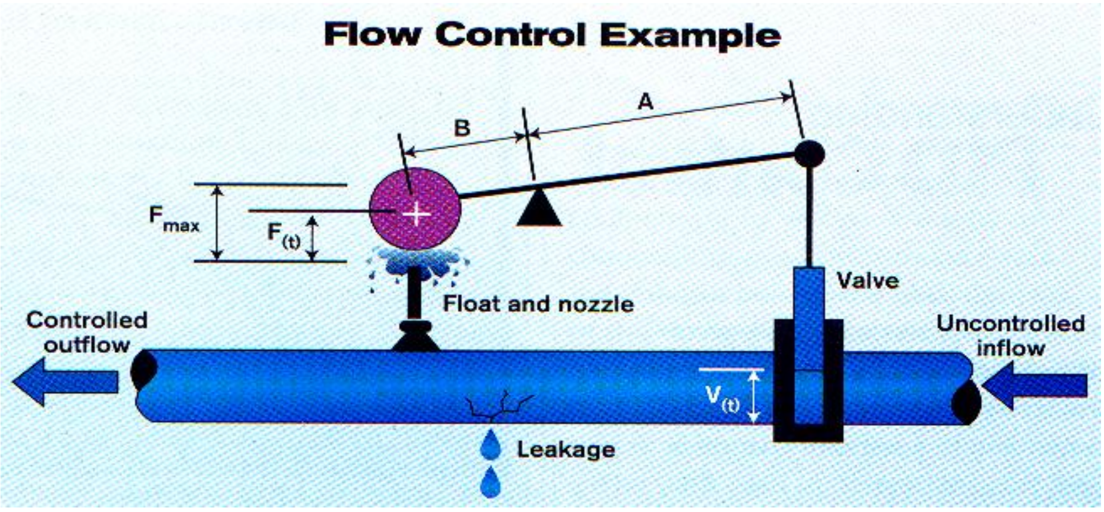
In this system the level in the storage tank needs to be maintained, so the introduction of the cascade system means that the flow into it may be kept at a steady flowrate rather than having the control valve going fully open to fully closed to maintain the level at setpoint. In simple form the output of the first controller *(master)* is used as the *(remote)* setpoint for the second controller *(slave).*

10.11 Tuning Controllers

Understanding P.I.D Control

A feedback controller is designed to generate an output that causes some corrective effort to be applied to a process so as to drive a measurable process variable towards a desired value known as the setpoint. The controller uses an actuator to affect the process and a sensor to measure the results.

Virtually all feedback controllers determine their output by observing the error between the setpoint and a measurement of the process variable. Errors occur when an operator changes the setpoint intentionally or when a disturbance or a load on the process changes the process variable accidentally. The controller's mission is to eliminate the error automatically.



A mechanical flow controller manipulates the valve to maintain the downstream flow rate in spite of the leakage. The size of the valve opening at time t is V(t). The flow rate is measured by the vertical position of the float F(t). The gain of the controller is A/B. This arrangement would be entirely impractical for a modern flow control application.

Consider for example, the mechanical flow controller depicted above. A portion of the water flowing through the tube is bled off through the nozzle on the left, driving the spherical float upwards in proportion to the flow rate. if the flow rate slows because of a disturbance such as leakage, the float fails and the valve opens until the desired flow rate is restored.

In this example, the water flowing through the tube is the process, and its flow rate is the process variable that is to be measured and controlled. The lever arm serves as the controller, taking the process variable measured by the float's position and generating an output that moves the valve's piston.

Adjusting the length of the piston rod sets the desired flow rate; a longer rod corresponds to a lower setpoint and vice versa.

Suppose that at time t the valve opening is V(t) centimetres and the resulting flow rate is sufficient to push the float to a height of F(t) centimetres. This process is said to have a gain of Gp = F(t)IV(t). The gain of a process shows how much the process variable changes when the controller output changes. In this case,

[1] F(t) = Gp V(t)

Equation [1] is an example of a process model that quantifies the relationships between the controller's efforts and its effects on the process variable.

The controller also has a gain Ge, which determines the controller's output at time t according to

[2] V(t) = Ge (Fmax - F(t))

The constant Fmax is the highest possible float position, achieved when the valve's piston is completely depressed. The geometry of the lever arm shows that Ge = A/B, since the valve's piston will move A centimeters for every B centimeters that the float moves. in other words, the quantity (Fmax - F(t)) that enters the controller as an input "gains" strength by a factor of A/B before it is output to the process as a control effort v(t).

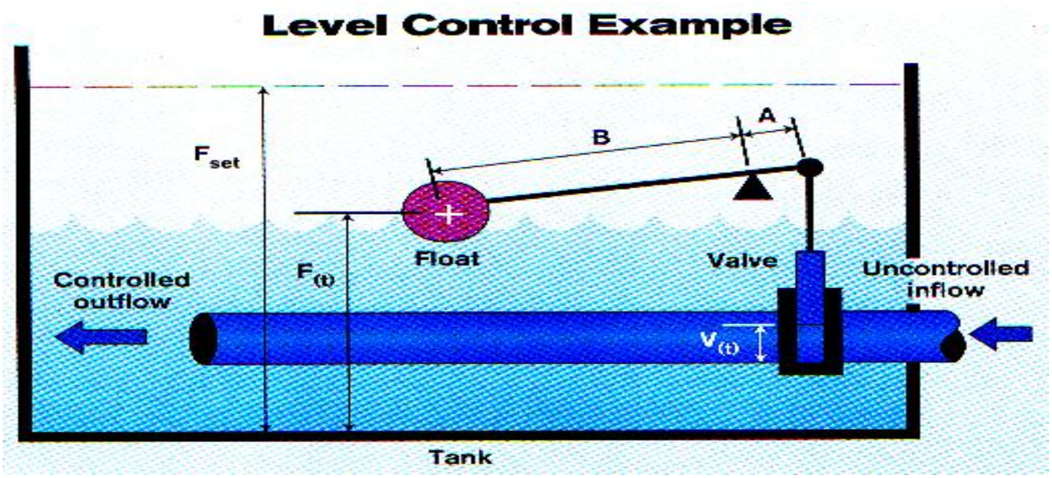
Note that controller equation [2] can also be expressed as

[3] V(t) = Gc (Fset - F(t)) + VB

where Fset is the desired float position (achieved when the flow rate equals the setpoint) and VB = Gc (Fmax - Fset) is a constant known as the bias. A controller's bias represents the control effort required to maintain the process variable at its setpoint in the absence of a load.

Proportional control

Equation [3] shows how this simple mechanical controller computes its output as a result of the error between the process variable and the setpoint. It is a proportional controller because its output changes in proportion to a change in the measured error. The greater the error, the greater the control effort,. and as long as the error remains, the controller will continue to try to generate a corrective effort.



The same mechanical controller now manipulates the valve to shut off the flow once the tank has filled to the desired level Fset. The controller' gain of A/B has been set much lower, since the float position now spans a much greater range.

So why would a feedback controller have to be any more sophisticated than that? The problem is, a proportional controller tends to settle on the wrong corrective effort. As a result, it will generally leave a steady state error (offset) between the setpoint and the process variable after it has finished responding to a setpoint change or a load.

This phenomenon puzzled early control engineers, but it can be seen in the flow control example above. Suppose the process gain Gp is 1 so that any valve position V(t) will cause an identical float position F(t). Suppose also the controller gain Ge is 1 and the controller's bias VB is 1. if the flow rate's setpoint requires Fset to be 3 centimetres and the actual float position is only 2 centimetres, there will be an error of (Fset - F(t)) = 1 centimetre. The controller will amplify that 1 centimetre error to a 2 centimetre valve opening according to equation [3].

However, since that 2 centimetre valve opening will in turn cause the float position to remain at 2 centimetres, the controller will make no further change to its output and the error will remain at 1 centimetre.

Integral control

Even bias-free proportional controllers can cause steady-state errors (try the previous exercise again with Gp = 1, Gc = 2, and VB = 0). One of the first solutions to overcome this problem was the introduction of integral control. An integral controller generates a corrective effort proportional not to the present error, but to the sum of all previous errors.

The level controller depicted above illustrates this point. It is essentially the same float-and-lever mechanism from the flow control example except that it is now surrounded by a tank, and the float no longer hovers over a nozzle but rests on the surface of the water. This arrangement should look familiar to anyone who has inspected the workings of a common household toilet.
As in the first example, the controller uses the valve to control the flow rate of the water. However, its new objective is to refill the tank to a specified level whenever a load (ie., a flush) empties the tank. The float position F(t) still serves as the process variable, but it represents the level of the water in the tank, rather than the water's flow rate. The setpoint Fset is the level at which the tank is full.

The process model is no longer a simple gain equation like [1], since the water level is proportional to the accumulated volume of water that has passed through the valve. That is

[4] $F(t) = \int Gp\ V(t)\ DT$

Equation [4] shows that tank level F(t) depends not only on the size of the valve opening V(t) but also on how long the valve has been open.

The controller itself is the same, but the addition of the integral action in the process makes the controller more effective. Specifically, a controller that contains its own integral action or acts on a process with inherent integral action will generally not permit a steady-state error.

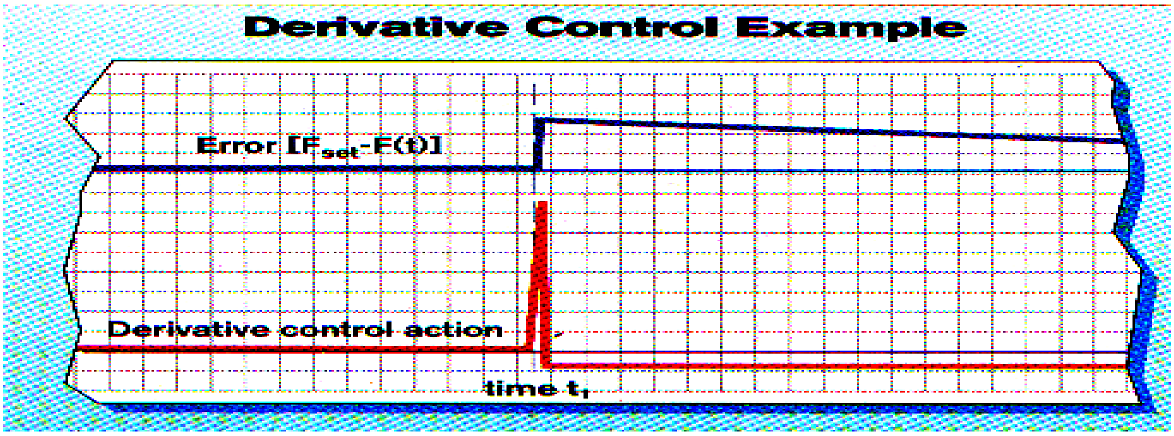
That phenomenon becomes apparent in this example. The water level in the tank will continue to rise until the tank is full and the valve shuts off. On the other hand, if both the controller and the process happened to be pure integrators as in equation [4], the tank would overflow because back-to-back integrators in a closed loop cause the steady-state error to grow without bound!

Derivative control

Proportional (P) and integral (I) controllers still weren't good enough for early control engineers. Combining the two operations into a single "PI' controller helped, but in many cases a PI controller still takes too long to compensate for a load or a setpoint change. Improved performance was the impetus behind the development of the derivative controller (D) that generates a control action proportional to the time derivative of the error signal.

The basic idea of derivative control is to generate one large corrective effort immediately after a load change in order to begin eliminating the error as quickly as possible. The strip chart in the derivative control example shows how a derivative controller achieves this. At time t1, the error, shown in blue, has increased abruptly because a load on the process has dramatically changed the process variable (such as when the toilet is flushed in the level control example).

The derivative of the error signal is shown in red. Note the spike at time t1.



The blue trace on the chart shows the error between the process variable F(t) and its desired value Fset. The derivative control action in red is the time derivative of the difference. Derivative control action is zero when the error is constant and spikes dramatically when the error changes abruptly.

This happens because the derivative of a rapidly increasing step-like function is itself an even more rapidly increasing impulse function. However, since the error signal is much more level after time t1, the derivative of the error returns to roughly zero thereafter.

In many cases, adding this 'kick" to the controller's output solves the performance problem nicely. The derivative action doesn't produce a particularly precise corrective effort, but it generally gets the process moving in the right direction much faster than a PI controller would.

Combined PID control

Fortunately, the proportional and integral actions of a full 'PID" controller tend to make up for the derivative action's lack of finesse. After the initial kick has passed, derivative action generally dies out while the integral and proportional actions take over to eliminate the remaining error with more precise corrective efforts. As it happens, derivative-only controllers are very difficult to implement anyway.

On the other hand, the addition of integral and derivative action to a proportional-only controller has several potential drawbacks. The most serious of these is the possibility of closed-loop instability. If the integral action is too aggressive, the controller may over- correct for an error and create a new one of even greater magnitude in the opposite direction. When that happens, the controller will eventually start driving its output back and forth between fully on and fully off, often described as 'hunting'. Proportional-only controllers are incapable of such behaviour.

Another problem with the PID controller is its complexity. Although the basic operations of its three actions are simple enough when taken individually, predicting just exactly how well they will work together for a particular application can be difficult. The stability issue is a prime example. Whereas adding integral action to a proportional-only controller can cause closed-loop instability, adding proportional action to an integral-only controller can prevent it.

PID in action

Revisiting the flow control example, suppose an electronic PID controller capable of generating integral and derivative action as well as proportional control has replaced the simple lever arm controller. Suppose too a viscous slurry has replaced the water so the flow rate changes gradually when the valve is opened or closed.

Since this viscous process tends to respond slowly to the controller's efforts-when the process variable suddenly differs from the setpoint because of a load or setpoint change -the controller's immediate reaction will be determined primarily by the derivative action, as shown on the derivative control example. This causes the controller to initiate a burst of corrective efforts the instant the error moves away from zero. The change in the process variable will also initiate the proportional action that keeps the controller's output going until the error is eliminated.

After a while, the integral action will begin to contribute to the controller's output as the error accumulates over time. In fact, the integral action will eventually dominate the controller's output, since the error decreases so slowly in a sluggish process. Even after the error has been eliminated, the controller will continue to generate an output based on the accumulation of errors remaining in the controller's integrator. The process variable may then overshoot the setpoint, causing an error in the opposite direction, or perhaps closed-loop instability.

If the integral action is not too aggressive, this subsequent error will be smaller than the original, and the integral action will begin to diminish as negative errors are added to the history of positive ones. This whole operation may then repeat several times until both the error and the accumulated error are eliminated. Meanwhile, the derivative term wiIl continue to add its share to the controller output based on the derivative of the oscillating error signal. The proportional action also will come and go as the error waxes and wanes.

Now replace the viscous slurry with water, causing the process to respond quickly to the controller's output changes. The integral action will not play as dominant a role in the controller's output, since the errors will be short lived. On the other hand, the derivative action will tend to be larger because the error changes rapidly when the process is highly responsive.

Clearly the possible effects of a PID controller are as varied as the processes to which they are applied. A PID controller can fulfil its mission to eliminate errors, but only if properly configured for each application.

10.12 Distributive Control Systems (D.C.S)

A D.C.S system is essentially a computer with a large memory bank, having masses of inputs and equally a mass number of outputs.

In a similar way that games are played on a computer, or games console, information or commands are sent to the processing unit from such examples as a handset, steering wheel, or joystick. In some cases the information is progressive eg: steering a car, or alternatively fireing a gun shot. In a similar fashion DCS can interpret variable information such as changing process variables through to plant alarm operations.

The inputs are converted from analogue to digital format, with each controller in the form of a software program, the configuration and link up to a signal processing unit allows appropriate signal information to be read, displayed and acted upon, that successfully combine to allow control of the process. Once acted upon information can be re-transmitted in the form of an output that can again be converted back from digital to analogue to enable operation of such devices as control valves etc.

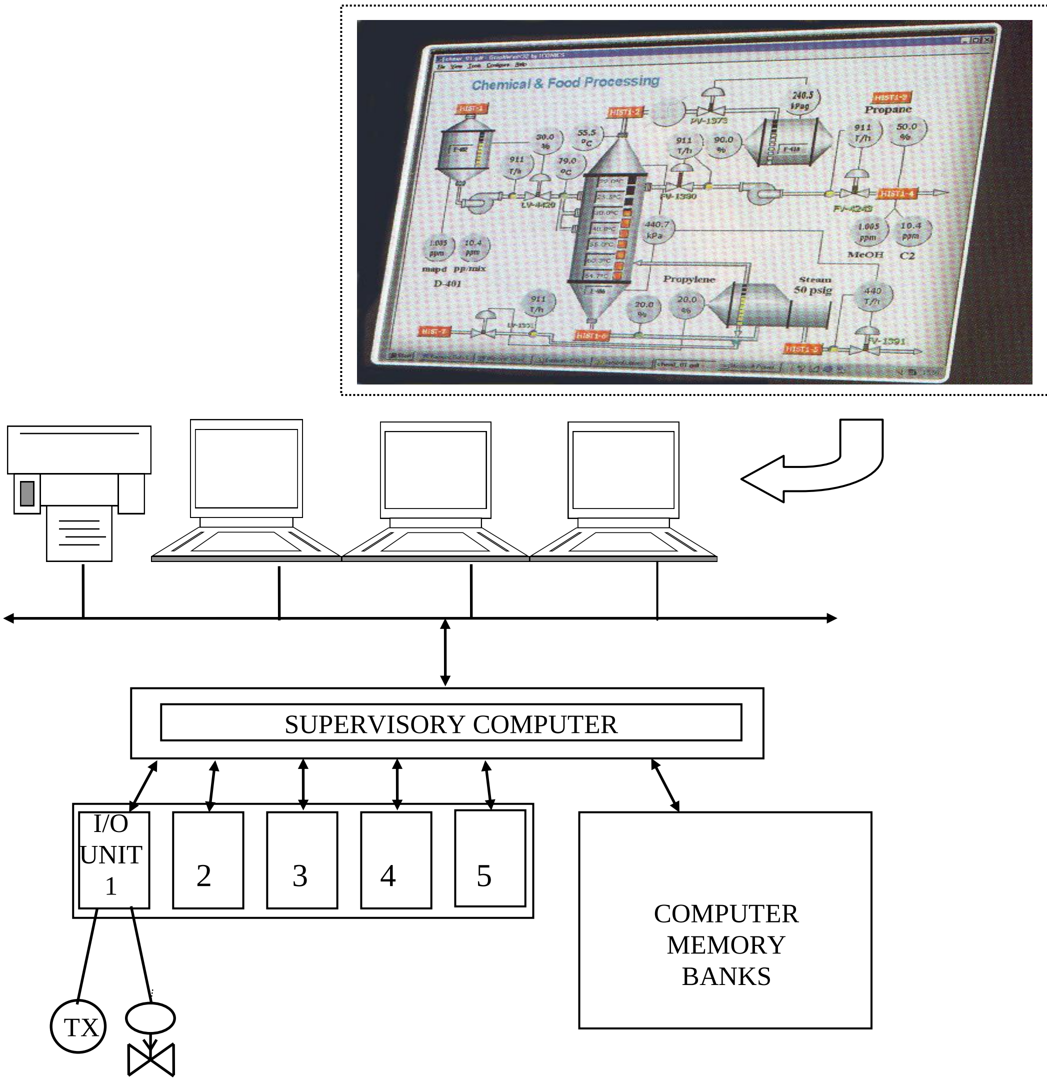
The same facilities exist in this form as in hardware form. The advantage of this system is the available peripherals such as graph trends, plant diagrams, touch screen control, and any of the other options available in the hardware format.

These can be a great space saver and have excellent controlling ability.

Most new plants and control systems use D.C.S, the negative is the immediate financial outlay which can be close to £.25 to £1million depending on size and complexity.

As can be seen from the photo the process operator can view the process plant performance from many different viewpoints. From these terminals control valves may be operated, whole plants run automatically, transmitters may be calibrated, predictive maintenance schedules can be created, process variables may be viewed, graphed, shown as trends….the options and use of DCS upgrade a quick as technology itself changes.

The following diagram below shows the typical components of a D.C.S:-



10.13 Process Control Into The Future

Just as technology continues to change, so to the advancements made in the field of process control technology. Process operators are gaining more responsibilities, some of which not only include operating the plant, but also carrying out maintenance.

The screens once operated by keyboards are now becoming ‘touch screens’ and instead of large tube monitors the advancement in silicon and colour LCD, display devices are now allowing flat screen displays no more than a few centimetre’s deep.

In the following example, the operator can view and operate the plant from a head mounted LCD display unit and arm mounted control pad. With this technology the operator can operate the plant from anywhere with communication range

In the final example, rather than being called in, why not ring in, with mobile phone access. Process variables can be viewed, monitored and even changed from your mobile phone.