Plantwide Control of Chemical Processes
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Lecture - 39
C4 Isomerization Process Revisited

So, today we are going to look at the n C 4 Isomerization process, plant wide control of that process. Now let us first look at what the process is, or why the process is required. You see in refineries, isobutene is a more important feast, more valuable feed stock than normal butane, because branched isomer are more valuable, because isobutene can be used to make certain branched chain molecules, then become additives for increasing the increasing the octane number of gasoline, and I think isobutene is also the feed stock for some other downstream products. So, n C 4 is not as valuable, n C 4 is valuable for fuel, you know burn it like LPG. So, i C 4 is valuable as a feed stock, and, so what is done is, normal butane is converted to the more valuable isobutene, and for that you use the isomerization process.

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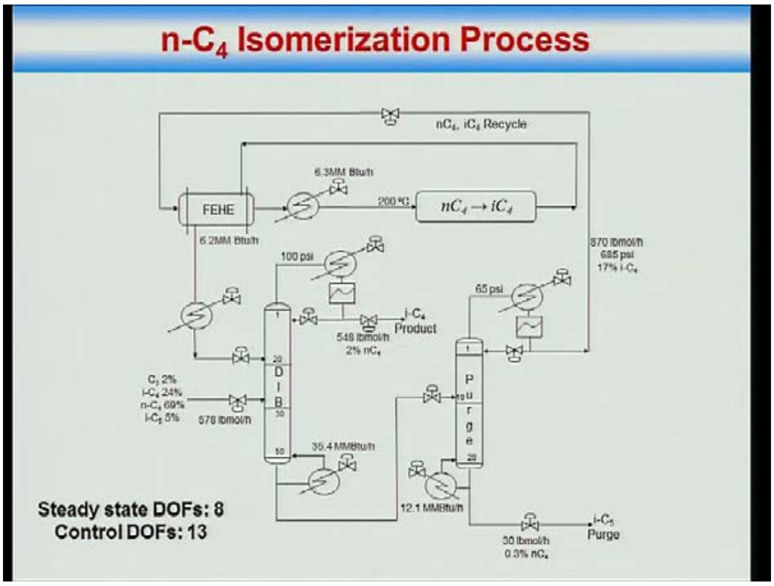

Now here is the process you have a C 4 feed stream, there is the C 4 feed stream, and this feed stream is containing C 3 propane, i C 4 and n C 4 isobutene and normal butane, and i C 5 iso pentane. So, the feed is mostly normal butane, little bit of C 3 and little bit of C 5 impurities. Now this feed is sent to a d isobutanizer DIB, de isobutanizer column, it is a tall tower you can see it is got 50 trace. And what the deisobutanizer does, is it takes out the more volatile i C four as the top product, and n C 4 drops down the bottoms. So, the light key, what is the light key. What is the light key for this column; i C 4 is the light key, and what is the heavy key; n C 4, n C 4 drops down. So, what will be the impurity in the product, any propane is bound to end up in the product, any propane in the feed, is bound to end up in the product; yes or no. So, the impurities are all the propane; that is coming in the fresh feed, plus any n C 4 that leaks up the top, in the bottoms what is the impurity, any i C 4 that leaks down the bottoms. So, the bottoms is fed to a purged column, what does the purged column do.

It removes i C 5 down the bottoms, which is heavy, and the mixture containing n C 4 and some i C 4, any i C 4 that might have leaked down the bottoms, is sent to the reaction section. How is the reaction done, you know you send the distillate through a feed affluent heat exchanger, what is the feed affluent heat exchanger do. It heats up the cold feed to the reactor, using the hot reactor affluent. And then you have a heater that heats the feed to the reactor temperature, and then you have an adiabatic pact bed reactor, in which the reaction n C 4, isomerization reaction n C 4 goes to i C 4 occurs. The reactor outlet is used to pre heat the reactor feed, and then this outlet from the feed affluent heat exchanger is cooled, condensed, and because n C 4 has been converted to i C 4 in the reactor, this feed is, you know this stream is richer in i C 4; therefore, it is fed slightly above, ten trace above I think, or 20 trace above, 20 trace above the fresh feed in the deisobutanizer column, so this is the process.

Now you are going to develop a. There is a base case that has been taken from one of Luyben’s case studies. So, without, let us see 17 percent i C 4. Degrees of freedom; steady state degrees of freedom, Batao Bhai how many steady state degrees of freedom two three four five; no, one for the fresh feed, two each for the two columns; that is five, one for the heater, one for the cooler, so how much is that; seven, you have to count the fresh feed too, so seven. I have also included the reactor pressure as a degree of freedom; you know, what is the pressure in the gas loop, that has also been included as a degree of freedom. So, including that there is eight degrees of freedom, controlled degrees of freedom is 13, you just count the number of valves; that is 13.

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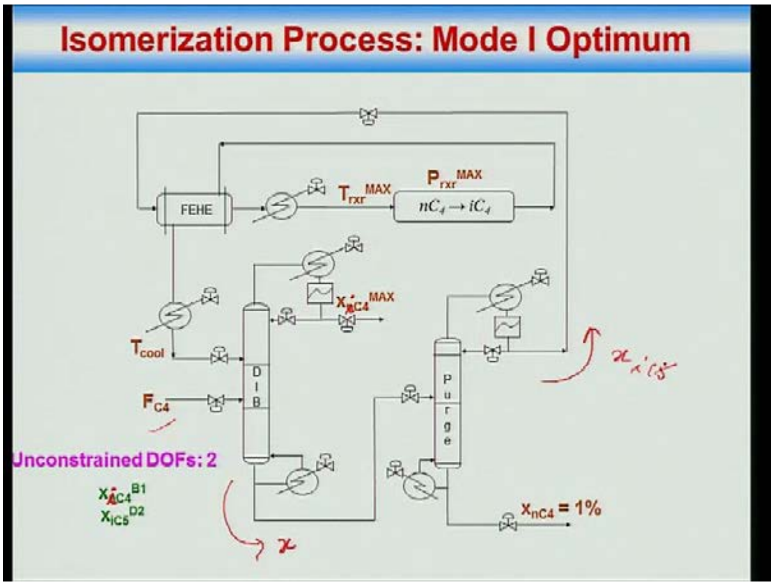

Now, mode one; given, so in mode one what do we have. In mode one this is given, the fresh feed rate to be processed, the processing rate is given. Also why do you have the cooler, basically to condense the vapor, to liquefy the vapor. So, basically the temperature at the outlet of the cooler is essentially fixed. So, we will say that these two things are fixed, F C 4 the amount the processing rate is given, and how much I need to cool down, so that everything condenses; that is also sort of fixed by the vapor liquid equilibrium; yes or no. So, t cool is fixed, F C 4 is fixed. The reactor should be operated at maximum inlet temperature why is that, and the pressure in the reactor should be as high it can be why is that. Maximum reactor temperature implies maximum conversion, maximum conversion means, less amount of n C 4 circulating around the plant. If less amount of n C 4 is circulating around the plant, then, if less amount of n C 4 is circulating around the plant, then you essentially have to boil less of in the purged column; yes or no. Also you will you will save energy in the DIB column, because this.

So, for the same reason, pressure at the reactor should be maximum. This is like you know in a liquid phase reactor the level of the reactor should be maximum, similarly pressure of the reactor should be maximum. Obviously you want the impurity that is allowed in your product, what is the impurity n C 4, that should be at its maximum limit; yes or no. Anything else that you can think of, that will have an economic impact. What about the purge, you know you are losing n C 5 here, what about this guy, what will be the impurity here. So, normal butane can go down here, normal butane is your raw material, which you are converting to i C 4, its valuable. If you are going to lose it, then you are going to lose it. So, you do not want too much n C 4 to leave down the bottoms of the purge column; yes or no. So, I am saying we will not allow n C 4 to go beyond one percent. So, we will just say fix it, because that is an indirect way of imposing an economic constraint, ensure that too much n C 4 does not leak down the bottoms.

So, how many things have been specified one two three four five six, what is left; two degrees of freedom are left right, and constraint degrees of freedom two. What are those two degrees of freedom. I m just saying the amount of n C 4 leaking down here, and the amount of i C 5 leaking up here, how does that affect your economics. If you allow more amount of n C 4 to leak down the bottoms of the first column, what happens to the boil up in the first column, it goes down, it will go down. If I allow more heavy key to leave down the bottoms boil up will go down; yes or no; so I save an energy. But then that whatever has leaked down, must be sent up in the purged column, so that will go up, so these are the two opposing.

Also there is another fact there is a another effect, the dilution effects, so if you allow too much n C 4 to leak down the bottoms, it will dilute the n C sorry this should be i. So, that is the this should be i x. So, the amount of isobutene and the amount of isopentene circulating around, these are your two unconstrained degrees of freedom. If you allow more to leak out a i C 4 to leak down the bottoms of the deisobutanizer, your deisobutanizer duty goes down, however the purge column duty will go up, because I has to be sent up. Also your feed to the reactor will be more dilute; yes or no. So, these are the opposing effects that give you an optimum, a similar argument also holds for x i C 5 in the distillate here. These are your two unconstrained degrees of freedom, how much i C 4 and i C 5 you are allowing to circulate around the plant, these are your two unconstrained degrees of freedom; yes or no, agreed.

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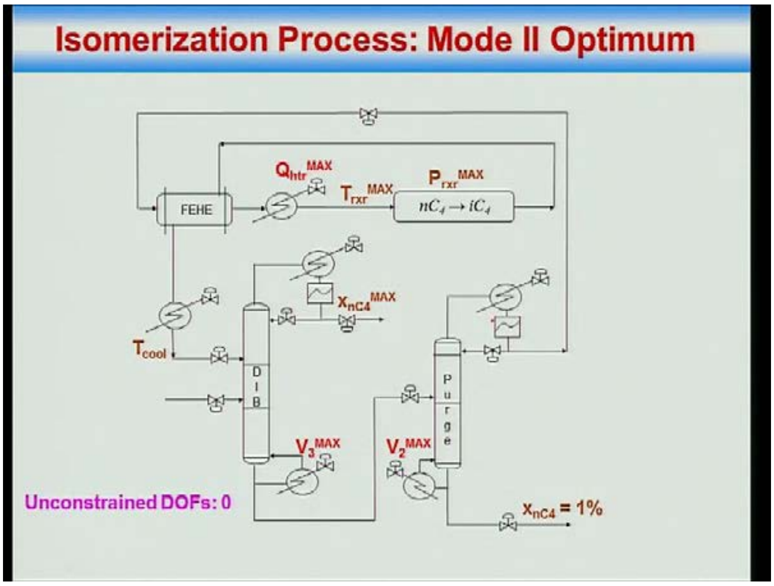

Mode two; I want to maximize throughput, then what do I do. Mode one constraints will anyway be active, what happens is, of course, now the throughput is not no longer fixed. It is the degree of freedom that will be optimized. So, one two three four five constraints are active, plus these three constraints becomes active, and so you will count 5 plus 3 is actually eight, all degrees of freedom have been used up, in maximizing your throughput. So, your steady state operating degrees of freedom is eight, eight constraints are active, equality or inequality; yes or no. So, now by the way does it make sense, I mean think about it, if I go back; F C 4 is fixed, let us say I am jacking up F C 4.

What am I basically saying is, as I am jacking up F C 4, Q heater max becomes active, and if I want to increase F C 4 further I should also maximize my boil up here in both the columns. Why is maximizing boil up causing rise in throughput, how will you maximize boil up. Why will it suck up more. Sure I agree. What is boil up doing, V 3 max is essentially preventing i C 4 from leaking down the bottoms. If I want to drive it up, what would I. If I drive to max what will happen to the i C 4 leaking down the bottoms, it will become less. Similarly if I drive V 2 to max; how see n C 4 is fixed, that constraint has to be active. So, if I am maximizing the boil up what is happening is, I am sending i C 5 up. Ah man am I sending i C 5 up; no, if I increase the. Well maybe we will talk about it; once the control system is there I think it will be very clear.

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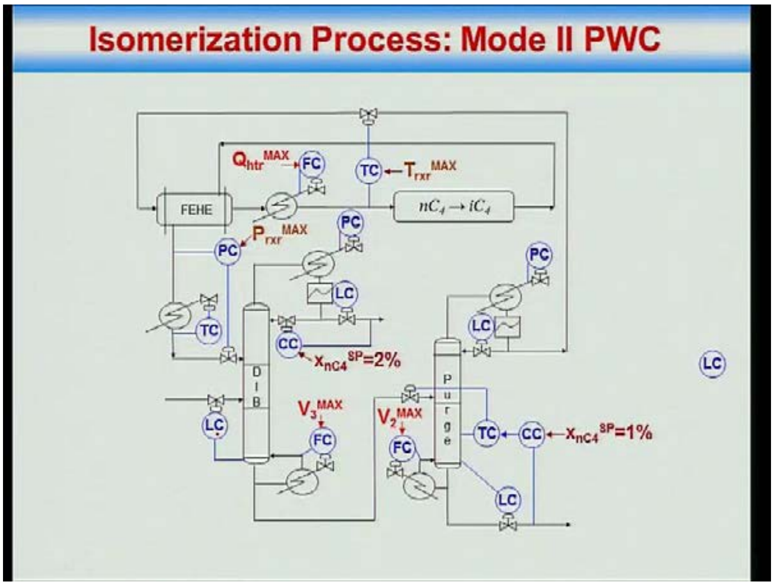

So, now let us just unconstrained degrees of freedom is 0, now let us just that level controller is, there is in case I need to put in some controller, you know copy paste. Now because these three constants are active, these are flow control, not used for controlling anything. Then I need to control the temperature in, going into the reactor. The only way I can do that, is by adjusting the flow through the reactor. So, therefore, that is this; see this stream, is what flows through the reactor. So, by adjusting this flow, I am controlling the temperature inside the reactor; yes or no. It is kind of like adjusting hot and conventionally what you would do is, you would adjust the heater duty to control the reactor temperature. FEHE is actually working like a heat exchanger, which is depending upon the flow of the. FEHE; there is no control degree of freedom there, whatever heat gets transferred there gets transferred. But the flow that is coming into the FEHE is (()) the reason that if, you know the lesser the amount of m n, the input formed a purge coloumn, the more it will get (()). But I want that input to be as high as possible, because I want to maximize throughput more through.

So, that is why we need to control the flow in a way, so that there is you know limited amount of. No, none of that. I cannot use the heater duty, because it is active, conventionally I would use the heater duty, but because that is active I do not use it. Next handle that is available to me is how much it is flowing through the system. So, I am adjusting the flow through that, through the reactor. And that flow through the reactor, happens to flow through the FEHE, happens to flow through the heater, happens to be the distillate of the purge column, but it has got nothing to do with, but please understand that the that the degree of tightness of control this way, or alternatively this way. Suppose I was doing it this way, I could have done it this way too, suppose I would do. The degree of tightness of control would be comparable, do you see what I am saying. I make an adjustment in the furnace fuel flow rate, firing, greater firing happens, the tubes heat up, and as the tube heat up, you know my feed to the reactor is hotter, that takes may be a couple of minutes, may be three four minutes depending on how big the furnace is, similarly I change the flow rate and then, because flow rate is essentially in compressible flow.

So, I make a change in flow here, there is immediate change in flow there, and therefore, things heat up or cool down immediately. So, I make a change in the furnace duty, things heat up or cool down almost immediately. I make a change in the flow through the system; things heat up or cool down almost immediately. So, the degree of tightness of control of temperature will be about the same; agreed. Whether I am the controlling the temperature using the furnace, or I am controlling the temperature using the flow through the reactor; yes or no. Once we agree on that, then, temperature control is this way, I also need to control the pressure. So, I control the pressure by adjusting how much I am withdrawing from here. So, that gives me pressure control; that is it, these are my. Just a second, I also need to control the purities.

So, I am saying composition control, but this will typically be done using, for example a delta t measurement, what is the delta t measurement. If you have a separation; that is easy in a column, then you will have a large temperature drop across the column, then temperature is a good indicator of composition; tray temperature. But if you have closed boiling components, which is the case here i C 4 n C 4, they will be close boiling. So, over 50 trays your temperature may be changing by only 15 20 degree 15 degree Celsius. So, therefore, temperature is not a good indicator of composition why, because change in temperature from one tray to the next is very small. That small change in temperature could be due to change in composition, or it could also be due to small fluctuation in pressure across the column.

So, whether the change in temperature is due to composition, or the change in temperature is due to slight change, small changes in the pressure, of the column, you really cannot tell, and therefore, temperature is not a good indicator of what, composition. So, in such columns which are called super fractionators, tall towers 50 trays here. In such columns what is typically done is, you instead of controlling t you control a delta t. Delta t means, you know you will take five trays one tray here, may be five trays above or below. Take the difference in these 2 tray temperatures, when you take the difference in these two tray temperatures the pressure effect gets cancelled out, because you are taking the difference, the pressure effect, if the pressure of this tray is rising, pressure of this tray also must be rising.

So, the temperature of both the trays, boiling temperature of both the trays have increased by about the same amount, when you subtract the two that pressure effect gets canceled out. Then the difference in the tray temperature is essentially due to change in composition as you move from here to here, or from here to here. So, then delta t is a better indicator of what; composition. So, typically what will be done is, you will have a delta t controller that is adjusting the reflux, and the delta t set point is being adjusted by the composition controller, but for ease of drawing I have just shown the single composition controller there. Similarly in the second column, you have. You see, now the boil up is at max, since the boil up is at max, you cannot use it for temperature control.

Therefore, you control the temperature, or the amount of i, the amount n C 4 leaking down the bottoms, you control that by adjusting the feed. And of course, in order to keep the composition fixed at one percent, you have a cascade composition controller. So, I have put in place loops for controlling the active constraint, is there any other active constraint. Let us see one two three 4, here you can have one two three four, you know six, there should be eight something is missing; no pressure is controlled, temperature is controlled composition; no. I said there were eight active constraints, are there eight active constraints here, one two three four five six seven, where is the 8th one. Cooler temperature, t cool is the one that I missed, forget that.

All constraints have been controlled, except that t cool will be controlled by the cooler duty, that loop has not been shown, so that makes it eight. So, now I have put in the inventory loops, so t cool comes in there. Level control, you see now the level controller in the purge column, because the distillate is taken for temperature control to the reactor. Level control has to be this way, there is no other option. Level control in the bottom can be this way, so we do it that way. Level control in the deisobutanizer has to be using the fresh feed; there is no other option, because there is no other valve left. So, I have put in the level controllers; of course, pressure controllers will be as before. This is my control system for processed operation at max throughput.

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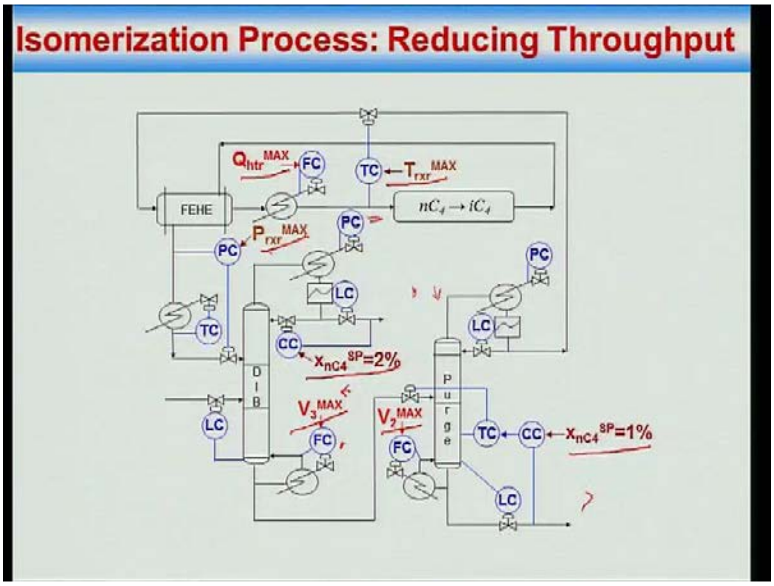

Now let us say I want to reduce throughput, how will I reduce throughput. Well let me just I will keep the boil up in the first column, in ratio with the total feed to that column, and I will adjust, you see I will reduce that, I will reduce or I will increase. No I will reduce the ratio set point, we will repeat it again. Then I will do the same thing here, I will keep the boil up in ratio with the fresh feed, and I will reduce that set point. And finally, what I will do is, I will use this Q max, by heater duty as my throughput manipulator, to reduce the throughput further. Let us go in the reverse direction, I am at low throughputs I want to increase throughput, what do I do, I start jacking up my heater duty, reactor heater duty; the throughput manipulator, as I jack it up, it is the first constraint to become active, so this becomes max, this max is out. Then what do I do, I start increasing this ratio set point; this guy.

As I am increasing this ratio set point, the boil up is, as the boil up is increasing what is happening, more material is sent up, so now what I will get is, I am increasing the boil up, the amount of n C 4 that is leaking down the bottoms is kept fixed, by the controller here, this is fixed, this keeps it fixed whatever is leaking down. So, then what happens is more material accumulates inside the reflux drum; therefore, the level controller starts putting more reflux back into the column. Now, because the reflux has increased into the column; the amount of i C 5 leaking up the top, will go down; yes or no. So, an increase in boil up, translates to an increase in reflux, and because the reflux has increased, now the i C 5 impurity up the top goes down. So, let us say earlier I was having 10 percent i C 5 leaking up the top, boil up is increasing, and therefore, this goes down to let us say one percent or less than one percent, does that make sense. Now what is happening is, the amount of i C 5 going up the top has reduced.

So, if I look at the stream, which is going to the reactor, it is got mostly n C 4, a little bit of i C 5, and a little bit of i C 4. By increasing the boil up what I have done is, I have made this go to 0 or close to 0. Therefore, the flow to the reactor has now become, this is gone, the flow to the reactor is essentially reduced, because the flow to the reactor is has reduced; Q heater is at max, therefore this stream will be hotter, because this stream will be hotter, I will start sucking more stuff here. I will say this is hotter, give me send me more stuff. So, as I start send sending more stuff, essentially I will suck in more feed, through the action of the inventory controllers; yes or no, does that make sense or no. By driving the amount of i C 4 to 0, I can essentially suck in more feed; yes or no.

Now let us go the, so even though Q heater has reached max, by adjusting the amount of i C 5 leaking up the top of the second column, I can still jack up my throughput a little bit. I can do the same thing here, I can do the same thing here, see what happens is, as I am jacking up this throughput manipulator, as I am jacking up this ratio set point this guy, V 2 reaches max, the column reaches its flooding limit. So, then what do I do, I say well, now I am going to decrease the amount of i C 4 that is circulating around. Just like I decrease this, I am going to decrease this too, how do I decrease this, by cranking up this boil up, by increasing the ratio set point boil up to feed ratio set point. As I am increasing this ratio point what happens is, what will happen, boil up increases.

If boil up increases the amount of i C 4 that is leaking down decreases, and therefore in the same way as before I will be sucking in more fresh feed; yes or no. Even though the increase in throughput by these is very small; two percent or three percent of the order of three to four percent; nevertheless three to four percent integrated over years of operation, translates to you know a heck of a lot of yearly revenue, so that is that. So, you can look at it two ways, I am at a given low throughput, I want to jack up throughput, or I am at max throughput, and I want to jack down throughput. So, if I am at max throughput, which is what this is. How do I, what do what do I take up, I take up control of what, as I am reducing throughput. Well I take up ratio control, I take up ratio control here, and then finally I reduce the throughput using this.

So, boil up is taking up ratio control, with respect to the feed to the column, both cases. Both boil ups are maintained in ratio with the feed to the column, that make sure essentially that you are not unnecessarily consuming too much over boiling, consuming too much energy. Looked at the other way I am at a low throughput, when boil up when a column gets flooded, I give up that ratio control. And when I give up that ratio control what is being given up, what is essentially being given up. When this hits max, the amount of i C 5; that is going up floats, it is no longer under control, it will float to whatever it has to float to. Similarly when this hits max, when this chap hits max, which is, well when do the, I do not know. So, when this hits max, the amount of i C 4 leaking down the bottoms is not in my hands; does that make sense or no.

So, when a constraint gets heat you lose something, what is it that you are losing, that must be very clear. So, in this case, when V 3 hits max, you lose control on, the amount of i C 4 leaking down the bottoms. Similarly when V 2 hits max, you lose control on the amount of i C 4 5 leaking up the top. Is that acceptable; that is the question. Yes the answer your control system must be structured, that the answer is yes; is that acceptable, does that make a, you know will the plant still be operable. Let us say I was using V 3 max or V 2 max to control impurity in the bottoms, you know let us say I was using, you know I was doing the conventional stuff, I was doing this. Let us say I was doing this, would this be acceptable when V 3 hits max. When V 3 hits max you essentially lose control of how much n C 4 is dropping down the bottoms.

That has an economic penalty, because n C 4 by converting it to i C 4 I am making money. Now I am just letting n C 4 go down, there’s a severe economic penalty associated with that. Similarly on the other hand, by losing control of, because it is all in the recycle loop , how much i C 4 is leaking down, how much i C 5 is leaking up that is all inside the plant. There may be a penalty, but it is not very severe, why is it not severe, because energy is much cheaper than raw material and products; yes or no. Sir, what the recycle cost increase, but that does not hit me as much, as for example what I told you, it is all a matter of. So, having variability in recycle is ok, having variability in the fresh feeds, or in the products, or in the purge stream, I would say there will, you know you better think twice about it, before you say that is ok.

You will have to put your variability some place, where would you rather have it, in the product stream; hell no, that is what I am going to sell, in the raw material stream; hell no, it costs in the purge stream; hell no, you know my productivity goes down, because instead of making more i C 4, I am just dumping it down the purge, do you see what I am saying. So, where would I rather put it, in the levels; yes, in the heater duty boil up etcetera; yes, because those are all utilities. Next in the recycles, only in a worst case situation I would say, well there is no option, this is not working, that level control is not robust etcetera, I have to take some variability in one of the inlet or outputs to the plant, material in inputs or material outputs from the plant.

Then you will take that material output which costs the least, and will say let the variability go there. Do you see what I am saying, ultimately it is all about managing variability, where would you rather have the variability. In this case it turns out I need not have variability in the purge stream; I can manage my plant perfectly, without what. You know without compromising on any of my active constraints, you see in this scheme, this active constraint is tightly controlled; yes or no. This active constraint is tightly controlled, this active constraint is at max, this active constraint is at max, this active constraint is at max, this active constraint is tightly controlled. You tell me one thing that is not tightly controlled. Everything is tightly controlled, and the inventory management scheme is also robust, so I would rather put this.

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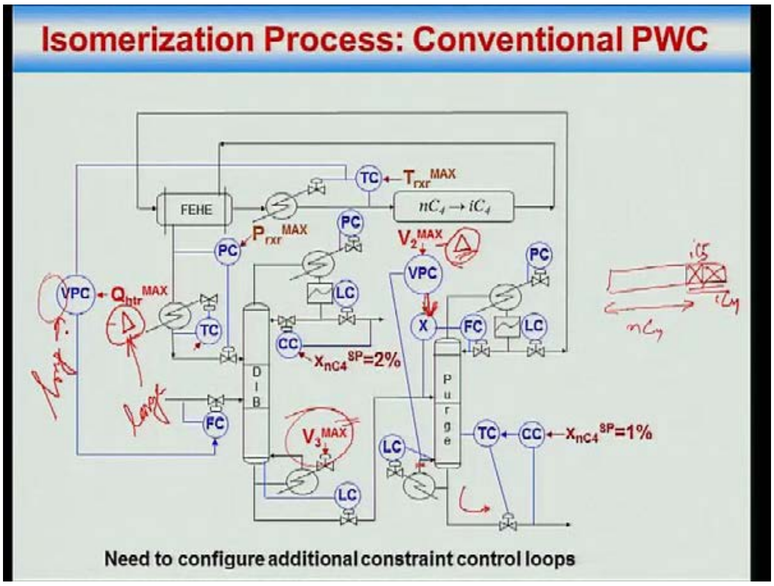

Now let us compare this with for example, bla bla bla conventional, in a conventional scheme I will have my throughput manipulator at the feed. Then I do the usual stuff, level controllers in the direction of flow, in the purge column, because this is a very small stream, you know because this is a very small stream, it is just a leak, I control the level using the boil up, everything else, everything is conventional reactor temperature control is using heater duty; yes or no. Then you know you will have this; yes or no, that is your control system. Mode one operation; no problem, no constraints are active, you are away from the constraint limit, this will work just fine. Now I start jacking up the throughput, by the way the two set points in green are unconstraint degrees of freedom, mode one unconstraint degrees of freedom what two, so what are the two set points that the operator can set.

So, these are the two set points that the operator is setting. Does that make sense or no. I am jacking up throughput, Q heater approaches maximum, but I cannot have a situation, where reactor temperature control is lost; yes or no. So, what do I do, I look at the position of the reactor heater valve, and adjust the throughput, to maintain Q heater at close to max, but not at not at max, because I cannot afford to lose reactor temperature control. So, I have to put in this loop. Then what do I do, I want to jack up my throughput further, what do I do. Well I will say this is my throughput manipulator, I will increase the ratio, by increasing the ratio I am reducing the re increasing the reflux, by increasing the reflux I am essentially reducing the amount of i C 5 leaking up the top, I am doing the same thing.

But then, this boil up approaches max. Once this boil up approaches max, I cannot afford to allow, I cannot afford to lose control of what is leaking down the bottoms, of this purge column. Therefore, what will I do I will adjust the ratio set point, so that this boil up is maintained close to max, but not max, because at max the column will flood, I will have to basically, probably take a shut down; yes or no. So this thing comes into being, and then what do I do. Then this is my throughput manipulator, I keep on reducing the composition of i C 4 leaking down the bottoms of the. Sir in this column when you increase the reflux, the amount of i C 4 going up the top is reduced, so how does that increase the boil up.

Well if you are increasing the reflux, you are putting more cold stuff in, but then i C 5 has to leak down the bottom, i C 5 will leak down the bottoms. So, how does that increase the boil up. That is what I am saying. So, when you are increasing the reflux, temperatures inside the column will go down, temperature in the stripping section will also go down. When temperature in the stripping section is going down, what does that mean. That means n C 4, which was the impurity, is dropping down; it is the light guy, that is dropping down. Now to maintain that at one percent, I will have to increase the boil up. So, as long as this loop is working, there will be an increase in boil up; yes or no. So, as I increase the reflux, boil up will also increase; yes or no. But I do not want the boil up to hit max, because I do not want to lose control of what is leaking down, because there’s an economic penalty associated with that.

So, what do I do, I will say I will measure the boil up, and I will adjust the ratio set point, so that the boil up is close to max, but not max. Note here that there will have to be, some amount of back of in the set point, minus delta. I have put in here max, but this will also have to have some back of minus delta, because I cannot afford to lose temperature control, reactor temperature control. So, now my throughput manipulator is this chap, again this guy becomes max, and then when this thing becomes max, so it is max, basically the amount of i C 4 that is leaking down floats. So, in this case, I am taking a back of here, and I am taking a back of here; yes or no, because these are hard constraints. Also notice that, when a constraint becomes active, I need to install a loop, that valve positioning controller has to come in, and when the constraint becomes inactive, that controller has to be taken offline; yes or no.

So, there is the additional complexity associated with configuring loops, when a constraint becomes active. The other option is, do not configure the loop, and then whatever is leaking down the bottoms in the purge, well you live with that, it may be more, it may be less I do not give a damn. I do not give a damn about the economic penalty associated with that; that is fine, but if do care about the economic penalty; you will have to configure that loop; yes or no. In the previous case, there were no loops, that ratio controller was always on, when boil up became active, ratio control was lost. When this boil up became active, ratio control was lost; yes or no. But as you are reducing throughput, whenever the boil up reduces sufficiently, ratio control will be taken up; do you see what I am saying.

So, that controller is always on, or is there, you are just loosing, you know it is heating a max limit, and then whenever it comes below the max limit, control will be taken up, whenever it goes to the max limit, control is given up. There is no question of configuring the loop, when a constraint becomes active; yes or no. So, this is the most, the structure that I drew before, even though all steady state degrees of freedom were lost, in driving as many active constraints. Yet you could manage it, in a way that there is acceptable inventory control, acceptable pressure control, while getting tight control of all active constraints. Here if you are not hitting constraint, if you are away from the constraint limits, there’s no problem, this will run. The moment you get to a constraint, you have to do things right; that is the problem with this structure. And as a consequence of that, so need to configure additional constraint control loops, in this system.

Sir in this, when we are controlling, for example when the heater duty reaches near max, we are controlling it with the fresh flow. That is what an operator will do, if you are the operator you are increasing the throughput manipulator, his duty is operating max, he will say do not jack it up further, because then I will lose temperature control. Sir that is means, when we start controlling we are essentially limiting the amount of throughput. Exactly, my heater max is limiting throughput, but note that, this loop is a very long one, I will make a change in the boil up, or in the fresh feed, and its effect on the heater duty will take, may be half n hour to reach. So, this will be a very lose loop, because this loop, this VPC will be very loose long loop, this is a long loop; therefore, this delta will be large.

You will have to back of big time, because the swings will be large, and over long durations. Sir once we have fixed the flow, then there's no reason why the boil up in the first column should reach max. Which flow, the main flow, flow to d I b; fresh feed. No, see agreed, as long as you do not change the ratio set point, there is no need to do anything, but you are saying that; no, I want to maximize production, so the figure that I drew, the feed to the reactor is mostly n C 4, a little bit i C 5, and a little bit i C 4. If I drive this to 0, and if I drive this to 0 then I can suck in more, that remains. How do I drive these to 0, by basically driving the boil ups to maximum. How do I drive the boil up to maximum, in this case I will say, jack up this ratio set point. As I am jacking up this ratio set point, boil up will approach max, but then I do not want to lose control of what is dropping down.

Therefore, what I will say is, adjust this ratio set point to keep the boil up close to max, but there will have to be a back of, you cannot keep it at max, same logic of you know, similarly here understood. So, the point is, you need to configure additional loops; that is a complexity no one wants. Operators in particular would say, this is a lousy loop I can do it myself, he will turn it off, and he will do the same thing, but by himself. Like this VPC, what is this doing? If the heater duty is approaching max, do not increase the throughput, do not increase the fresh feed, operator can do that himself. So, the he is acting as the VPC, you could have an automatic controller, or you could have the operator do it, but it is drawn to show that, some, that it needs to be done. In the previous case, nothing needed to be done, take up give up, take up give up, regardless of throughput.

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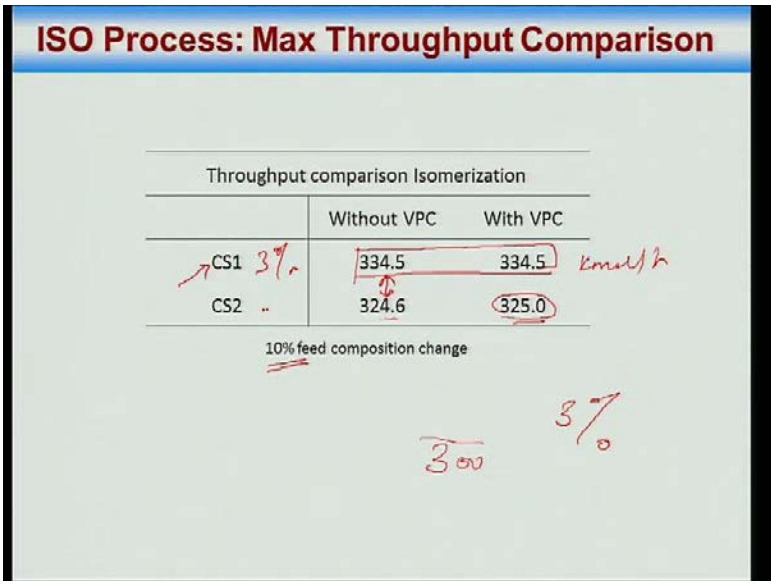

So, I now just took a 10 percent feed composition change, and what this is showing is that, maximize you will throughput is 334.5 I do not know, kilo moles an hour probably. In CS 1, and it remains. What the hell is without VPC. CS 1 is CS 1 there is no VPC there. CS 2 with VPC, you get a slight improvement, and it is very slight, because the loops are long. This almost no improvement between this and this, but you can compare this and this, how much is this; 10 kilo moles an hour on 300; so, 300 pe 10 Kithna Hua about three percent. Three percent extra throughputs you are getting using that control structure, using CS 1. This is giving three percent extra throughput, compare to this; yes or no. If I need to operate my plant at maximum throughput, well I will be making three percent extra material, integrated over the whole year; that is crores of extra revenue, millions of dollars of extra revenue. This is the same plant, run this way, same plant run that way. One control scheme happens to be better than the other, because it has been designed, taking into consideration the constraints that have going to become active; yes or no.

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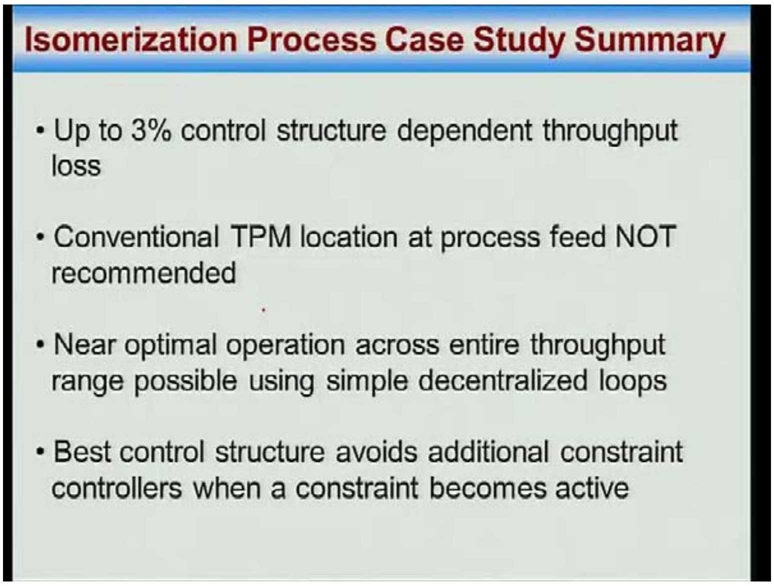

So, up to three percent control structure dependent throughput loss, conventional throughput manipulator location at process feed, not recommended; that is because all your active constraints are inside, they are not at the feed. Boil up in column one, boil up in column two, maximum heater duty. These are all inside the process, what is the point in having the throughput manipulator away. Also you get near optimal operation across the entire throughput range, using simple p I control loops, no nothing fancy, no fancy constraints managers, or MPC, or this or that. By just simply by addressing which constraint is going to become active, you are getting optimal operation over the entire throughput range. The best control structure avoids additional constraint controllers, when a constraint becomes active. You do not need to configure additional controllers; it is natural to that process. Of course, the key is what are the active constraints, my feeling is that, the set of active constraints does not change for a process, or even if it changes, it could be for example, well when the equipment is clean, these are your active constraints, then you know the furnace fowls up, six months or one year down the, that becomes the active constraint.

At worst you will have a situation, where an additional constraint becomes active, because there is deterioration in equipment quality. So, as far as I can tell, if you understand your process well enough, if your module of the process is good enough, then you will not have a situation that today this is the active constraint set, tomorrow something totally different is the active constraint set; do you see what I am saying. The active constraint set will be, because your capacity constraints are, where they are; that does not change, do you see what I am saying. So, what I think is as long as you know, what are your active constraints? Even if all you know your steady state degrees of freedom is ten, number of active constraints is ten, even then you can design a control system, that will give you safe stable and highly economical operation, economically efficient operation; yes or no. So, that is that, and I think.

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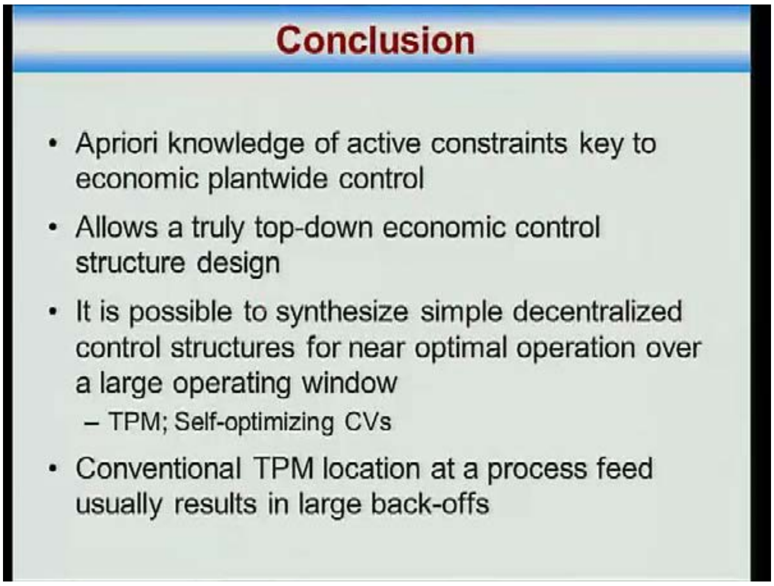

Well, so in conclusion a Apriori knowledge of active constraint is the key to economic plantwide control, and this active constraint set, is very unlikely to change, for a given process. Even if it changes, it will be changed by one, you know this equipment has degraded enough, so that now this is the bottleneck. It allows a truly top down economic control structure design; top down meaning, active constraints, I am not going to use them for control, put in the other loops, then put in the in your quality loops, and then put in your level and pressure controllers. It is possible to synthesize simple decentralized controlled structure for near optimal operation over the entire throughput, over a large operating window. You know you are not saying; this is my throughput plus 10 percent minus 10 percent, you are saying this is my throughput plus 50 percent minus 50 percent; that is a bloody large.

You know 330 was the maximum, I bet it was designed for about I think something like 250 or 260 kilo moles an hour, you can also go down to 200. So, 200 to 330 how much is that, that is a heck of a large change. You are not saying I am operating here, little bit here, and little bit there. No, you are operating over a large operating window, large throughput range. Well the other thing is conventional throughput manipulator location at a process feed, usually results in large back offs, large economic penalties, why is that. Because this is the calmest place in your plant, your active constraints are here, away from. Therefore, you cannot control those active constraints tightly, because you cannot control those active constraints tightly, you have to have a large back off, if you have a large back off, you take a large throughput penalty, or economic penalty, or whatever kind of penalty. And this will be typically the case, usually inside the recycle loop; that is where you will get active constraints; not at the feed or at the product. If you get at the feed, what do you need to do, just install another pump; big deal, but if you if you got a column that is flooded. You cannot do much about it, because it takes heck of a lot of money to install a new column; yes or no.

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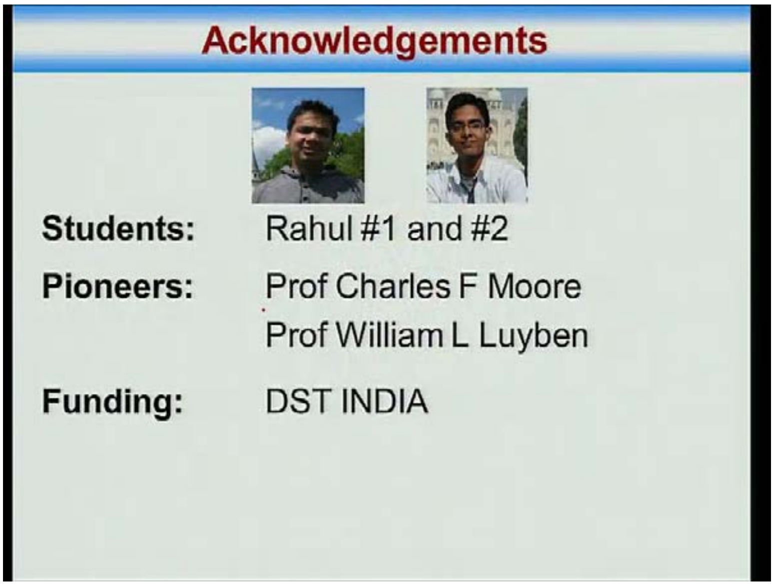

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