Process Design Decisions and Project Economics

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Module - 6

Economic Design Making: Case study of a gas absorber

Lecture - 30

Energy Balance, Process Alternatives and Design of the Absorber

Welcome, in this module we are looking at a case study of design of a gas absorber. And evaluating the economic potential at the various stages of design and generating process alternatives and once again choosing between these alternatives based on the economics of the process. This is essentially, application of the principles of process design, flow sheet synthesis and project economics that we have learnt in the previous modules. The problem at hand is about recovery of acetone from a waste gas stream generated from a process plant.

The stream is dilute containing only 10.3 moles per hour of acetone in 687moles per hour of air. Then in the previous lecture, we saw how we started the flow sheet synthesis. The first stage was to decide whether we have to go for a batch or continuous process. But, absorber is always a continuous process. So, that was the first decision. The second decision was about choosing the process for the solvent recovery, the solute recovery, acetone recovery. We saw how we had different alternatives compression, condensation, absorption, adsorption, membrane, reaction separation.

Then, we saw how we made a decision among these alternatives. We saw as how the experience or heuristics from the previous studies helped us in deciding the process alternative for the solvent vapor recovery. The next stage was designing of a simple flow sheet. We saw how we can design a simple absorber followed by distillation column. For recovery of acetone, we chose water as the solvent, which is a cheapest. Then we generated a process alternative of recycling water. Then we saw what were the pros and cons or the merits and demerits of such an alternative. Then we also did material balance for absorber as well as distillation column. Now, in this module we go ahead with the energy balance.

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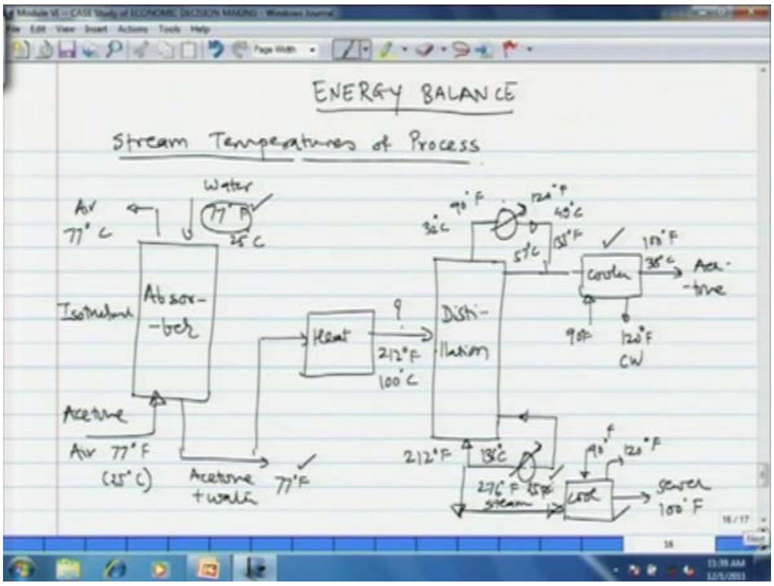

Now, let us see the stream temperatures of the process. We are looking at a process where we do not recycle water. Now, absorber is essentially an isothermal unit. We are looking at very dilute stream of waste gas containing only 10.3 moles per hour of acetone. So, the temperature does not change much. We are absorbing that acetone in 1943 moles per hour of water. So, the quantity of water is quite high. Therefore, we can reasonably assume that the temperature across absorber is same as that of water that is entering the process, which is 77 degrees Fahrenheit or 25 degree centigrade.

Air exits also at the same temperature and acetone water mixture also exits at the same temperature, 77 degrees Fahrenheit. Now, this acetone water mixture is then fed to a distillation column. Previously, it is heated. Now, what should be the temperature of the stream emerging from the heater? As we have seen in previous module as well as we have seen in mass transfer operation. In most of the cases, the feed to the distillation column is a saturated liquid story that is the liquid at its bubble point.

Now, we have a mixture of 1943moles of water with only; let us say, only 10.2moles of acetone. So, the bubble point of this particular mixture is going to be very close to the boiling point of water. So, we essentially assume the exit temperature of the stream from the heater to be 212 degrees Fahrenheit or 100 degree centigrade. At a distillation column, we are going to recover pure acetone, 99 percent acetone at the top as a distillate.

Once again, the quantity of water in it, as we calculated in previous module was very small; only 0.1 mole water in it, almost 10.2 moles of acetone. So, the dew point of that particular mixture is going to be very close to the condensation temperature of acetone. Therefore, we assume that the acetone that comes out from or that comes out of the condenser of the distillation column is about 135 degrees Fahrenheit or 57 degrees centigrade, which is the condensation temperature of acetone.

Now, this acetone has to be cooled before it is sent to further processing. Let us say, it is going to be a product of our process. Then we have to store it and we cannot store acetone at its boiling point. Therefore, we have to cool it further. Therefore, you can see here a second cooler after the condenser that cools the input stream of 135 degrees Fahrenheit or 57 degrees centigrade to almost room temperature that is 100 degrees Fahrenheit or 35 degrees centigrade.

Now, for this cooling, we use cooling water as the utility. That cooling water goes in the cooler at 90 degrees Fahrenheit and comes out at 128 degrees Fahrenheit. Let us say 32 degrees centigrade and 49 degrees centigrade. Now, these temperatures are going to be different in different seasons. For example, in winter these temperatures could be lower. This is water that is emerging from the humidification towers or cooling towers. Therefore, in winter it is likely that the cooling the water that is coming out from cooling tower will be at lower temperature may be 25 degree centigrade.

However, while designing particular exchanger, we have to consider the upper bound so that the design does not fail in case of temperatures raising than the design temperatures. The distillation column bottom, we have a re-boiler there. Once again, you see the material balance that was done in previous lecture. The amount of acetone that is present in the bottom is also very small, hardly 0.05moles per hour of acetone. So, the bubble point of the mixture in re-boiler of distillation column is going to be very close to 100 degrees centigrade or 212 degrees Fahrenheit; that is the boiling point of water.

Now, that water is part of water is boiled and sent back as the vapor reflux and part is taken out and sent to the sewer. But before we send it to the effluent treatment plant and alter to sewer, we have to cool that particular water. Once again, we have a heat exchanger here, cooler. That uses cooling water as the utility going in at 90 degrees Fahrenheit, coming out at 120 degrees Fahrenheit.

Now, for boiling the water in a distillation column, we can make use of a low pressure steam may be 25 PSI steam or a 1.5 bar steam 2 bar steam, which is at about 136 degrees centigrade or 276 degrees Fahrenheit. That is sufficient to boil the liquid mixture in the re-boiler. So, that is how we can make the balance or this we can determine the stream temperatures of the process. The next stage is to determine the heat load.

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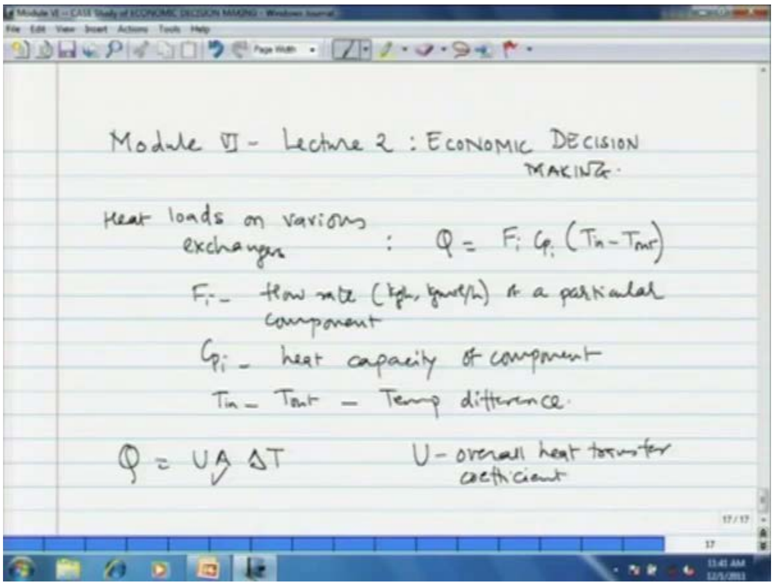

Module VI - Lecture 2 : ECONOMIC DECISION MAKING.

Heat loads on various exchangers : $Q = F_i C_{p_i} (T_{in} - T_{out})$

F_i – flow rate (kg/h, kgmol/h) of a particular component

C_{p_i} – heat capacity of component

$T_{in} - T_{out}$ – Temp difference.

$Q = UA \Delta T$ U – overall heat transfer coefficient

That heat load Q can be calculated from a simple legislation where F i is the flow rate either in kg or kg mole, whatever you choose or kg mole per hour. In case of continuous process, kg per hour flow rate of all components, a flow rate of a particular component. Then CP i is the heat capacity of that component. T in minus, T out is a temperature differences of the inlet and outlet stream. With this, we can calculate the areas of different heat exchangers of our process; using the relation Q is equal to UA delta T.

For determining U, the overall heat transfer coefficient, we again make use of heuristics depending on the type of fluid that is inside the tube or inside the shell. We can have a range of overall heat transfer coefficients available. Well, you can get these from any standard text like Perry’s chemical engineering hand book or the book of DQ Kern and process heat transfer, the range of overall heat transfer coefficients for a particular combination of streams.

Then, from delta T, we can calculate area and from area you can get the cost capital cost of the exchanger. Then you can also determine the cost of water that is used for operating that exchanger. In this case, we can have some process alternatives.

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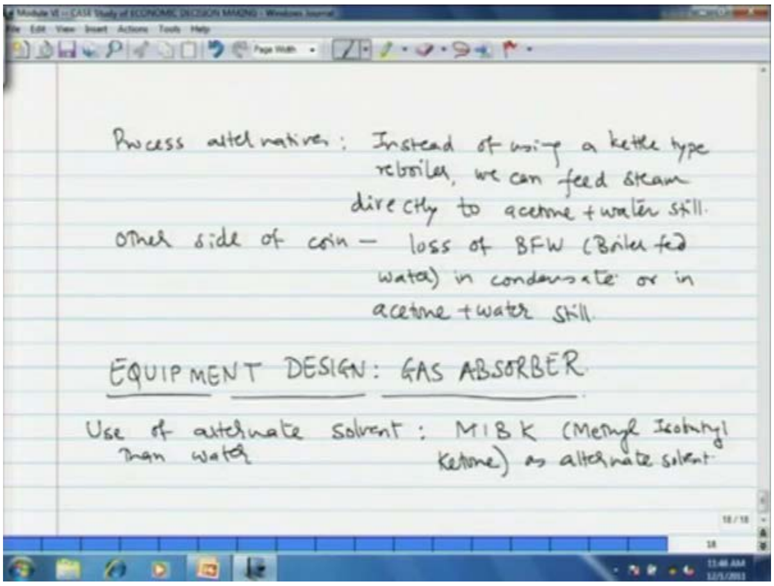

Process alternatives: Instead of using a kettle type reboiler, we can feed steam directly to acetone + water still.

Other side of coin — loss of BFW (Boiler fed water) in condensate or in acetone + water still.

EQUIPMENT DESIGN: GAS ABSORBER

Use of alternate solvent than water : MIBK (Methyl Isobutyl Ketone) as alternate solvent

Instead of using a re-boiler, let us say kettle type re-boiler, we can feed steam directly to acetone water still. This will increase the heat transfer tremendously as steam is directly condensing into the liquid, which it is suppose to boil. So, it will eliminate all kinds of resistances. However, the other side of the coin in this case is that, the water that will form after condensation of steam gets mixed with the acetone water still.

Then, it cannot be separated. Now, the water that is used for generating steam which also known as boiler fed water, is special type of water which is completely demineralized using iron exchange rising columns treatment. So, that water is rather expensive. You cannot afford wastage of that water. It has to be recycled almost to completion. Now, in case you are feeding steam directly to acetone water still, the water condenses in the still and it cannot be separated. So, that is a loss to that point we note other side of the coin loss of BFW boiler fed water in condensate or in the acetone water still.

Now, depending on the number of streams that you have in the process you can also go for heat integration of the process. But that is a topic that we are going to separately cover. In the present situation we have a very simple system. So, there is no much scope of heat integration. Now, the next stage of design is the actual design of gas absorber but before we do that, let us see some other process alternatives, like in previous lecture we talked of using of an alternative solvent than water. We can use some organic solvent, let us say, we try to use MIBK that is methyl isobutyl ketone as alternate solvent.

I mentioned to you in previous lecture that while doing the material balance we have not accounted for the loss of water that occurs through the exiting air. The air that is exiting from absorber is likely to be saturated with water vapor. Therefore, some, not all of the water that is fed to the absorber comes out of the absorber. There is always some evaporation loss.

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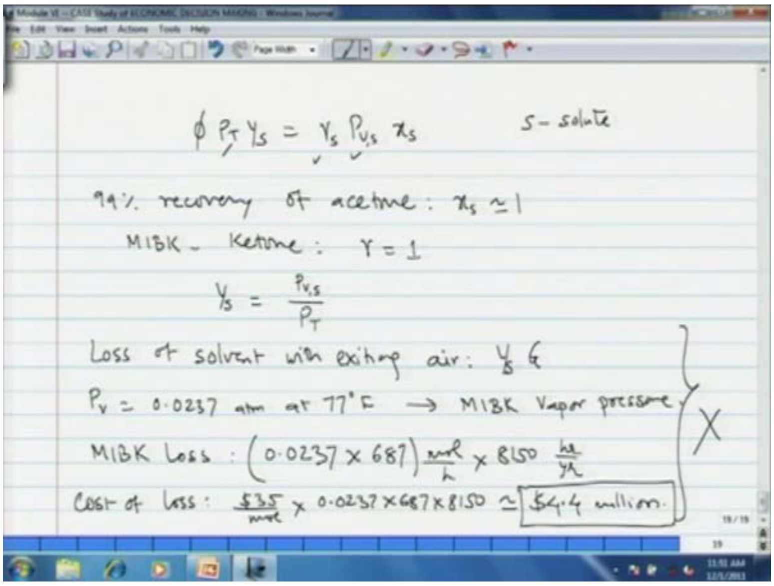

$\phi P_T y_s = \gamma_s P_{v,s} x_s$ S – solute

99% recovery of acetone : $x_s \simeq 1$

MIBK – Ketone : $\gamma = 1$

$y_s = \frac{P_{v,s}}{P_T}$

Loss of solvent with exiting air: $y_s G$

$P_v = 0.0237$ atm at 77°F → MIBK Vapor pressure

MIBK Loss : $(0.0237 \times 687) \frac{mol}{h} \times 8150 \frac{hr}{yr}$

Cost of loss : $\frac{\$35}{mol} \times 0.0237 \times 687 \times 8150 \simeq \boxed{\$4.4 \text{ million.}}$

Now, that loss again can be estimated from vapor liquid equilibrium like of the basic relation phi PTYS equal to gamma SPVS into x S where S denote solute which is presently the acetone. PT is the total pressure system. PV is the vapor pressure. Gamma is activity coefficient. x and y are mole fraction in liquid and gas respectively. Now, if we are talking of 99 percent recovery of acetone then excess is close to being 1. That is the solvent composition of the top tray. Now, if we use MIBK then it is essentially a ketone.

Therefore, the activity coefficient is likely to be close to 1. Therefore, what we have is the YS, that is the mole fraction of solvent exiting from or exiting with the exit gas or the air will be YS equal to, sorry, PV by or PVS by PT. Then the loss of solvent with exiting air is going to be YS into G, where G is the flow rate molar flow rate of air.

Now, in the present situation, let us say we talk of using MIBK. Now, MIBK has vapor pressure of 0.0237 atmospheres at 77 degrees Fahrenheit. We are taking of operation at atmospheric pressure so, the MIBK loss will be 0.0237 moles per hour into 687, sorry, 0.0237 into 687 moles per hour into 8150 hour per year.

Then, the cost of this loss will be the cost of MIBK 35 dollars per mole into 0.0237 into 687 into 8150. So, that comes out to be something like 4.6 million or 4.4 million dollars. Now, if you compare the loss of solvent with exiting air, which cannot be reduced. It can be reduced only by decreasing the vapor pressure of solvent or reducing the temperature of absorber, which is of course, quiet expensive. So, the loss itself is about 4.4 million dollars. We estimated the economic potential of the process in previous lecture as something like 1.4 million dollars if we recover all of the acetone that is exiting.

Now, comparison of these two figures would immediately tell you that the use of MIBK as solvent is going to kill the economy of the process because the solvent loss will be three times almost than the cost of the recovered acetone. Therefore, we cannot afford use of MIBK as a solvent for the process. It is a good idea to see the cost of other losses from the process.

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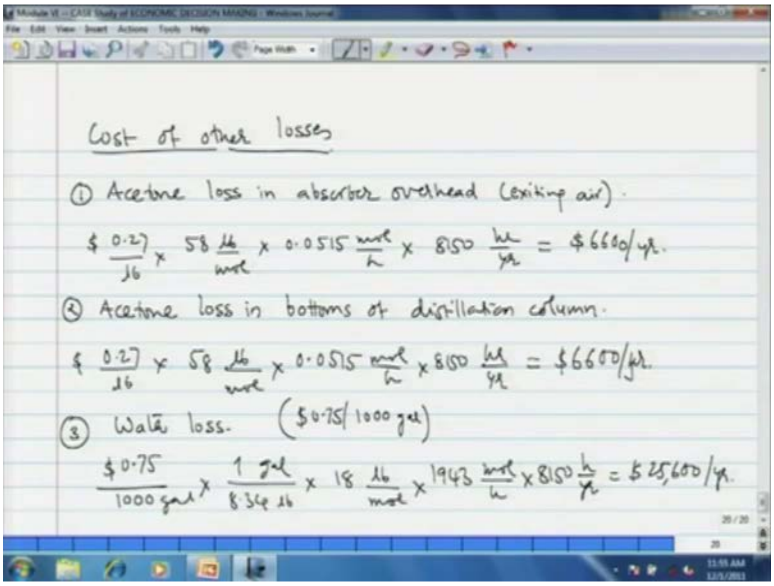

Cost of other losses

① Acetone loss in absorber overhead (exiting air).

$$\frac{\$\,0.27}{lb} \times 58\,\frac{lb}{mol} \times 0.0515\,\frac{mol}{h} \times 8150\,\frac{hr}{yr} = \$\,6600/yr.$$

② Acetone loss in bottoms of distillation column.

$$\frac{\$\,0.27}{lb} \times 58\,\frac{lb}{mol} \times 0.0515\,\frac{mol}{h} \times 8150\,\frac{hr}{yr} = \$\,6600/yr.$$

③ Water loss. ($0.75/1000 gal)

$$\frac{\$\,0.75}{1000\ gal} \times \frac{1\ gal}{8.34\ lb} \times 18\,\frac{lb}{mol} \times 1943\,\frac{mol}{h} \times 8150\,\frac{h}{yr} = \$\,25{,}600/yr.$$

Like as I mentioned to you in previous lecture that acetone exits at three locations. First is the acetone trace that are left with exiting air, acetone traces that are left with the water that is coming from distillation column, which is then fed to sever and then the pollution treatment cost. So, let us see what those costs, total cost that are there are, and first acetone loss in absorber over head or with the exiting air.

The cost of acetone we assume to be 27 cents per pound, molecular weight 58 pound per mole. Then the amount of acetone in exiting air was 0.51, 0.0515 mole per hour and then the number of operating hours per year. So, this comes out to be close to 66,00 or 6,600 dollars per year. Now, acetone loss in bottoms of distillation column. Again, we use the same number 27 cents per pound as cost of acetone into molecular weight. Then we have already determine that amount of acetone that will be present in bottoms will be 0.0515 mole per hour and 8150 hour per year of operation and thus loss also comes out to be about 6,600 dollars per year.

Then, the third loss is that of solvent water itself. Cost of water is much smaller. We can, I take a cost of about 75 cents per thousand gallons. Then the total loss annual will be this much. 1943 mole per hour into 8150 hour per year of operation into the molecular weight, 18 pound per mole into; now, we have to determine the volume of this water now, 1 gallon is typically 8.348 pounds. Then these many gallons and the cost of 75 cents per 1000 gallons; so this loss comes out to be something like 25,600 dollars per year. In addition to this, we have this sewer chargers and the pollution treatment cost. So, those are also not going to be much that those are likely to be of the order of about 6500 dollars per year.

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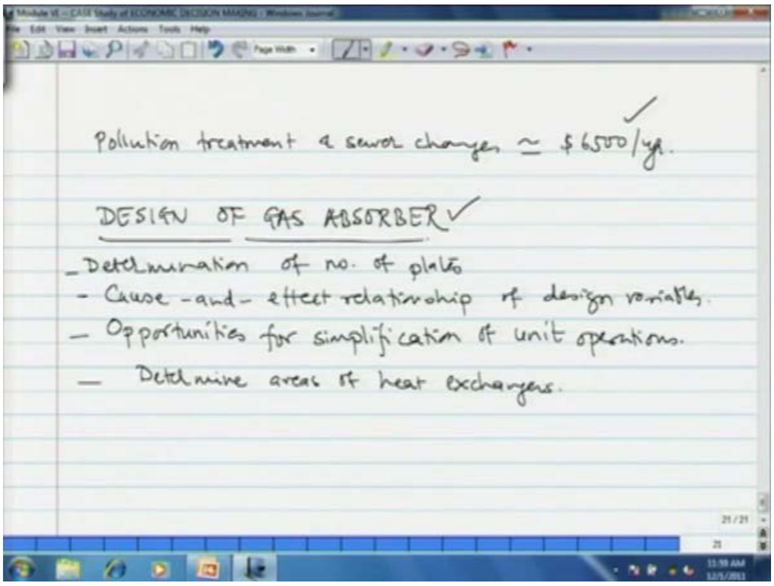

Pollution treatment & sewer charges ≃ \$6500/yr. ✓

DESIGN OF GAS ABSORBER ✓

- Determination of no. of plates
- Cause-and-effect relationship of design variables.
- Opportunities for simplification of unit operations.
- Determine areas of heat exchangers.

That point we note, pollution treatment and sewer chargers both are likely to be in the range of 6,500 dollars per year. So, what we can see from this analysis is that, all of these figures are much smaller than the cost of acetone that is recovered from the process which was estimated at about 1.4 million dollars. Therefore, we can go ahead with the design. So, that is how. Now, having said this, we go for the design of the equipment that is the main absorption column.

Now, the first step would be to determine the number of plates in the column design of gas absorber. We also need to know the cause and effect relationship of design variables and opportunities for simplification of unit operations. Of course, in addition to this, we have to also determine the areas of heat exchanger that we are going to use and also the distillation column. But, the principle or major unit of this particular mode is module is design of the gas absorber of this particular unit and then we start with it.

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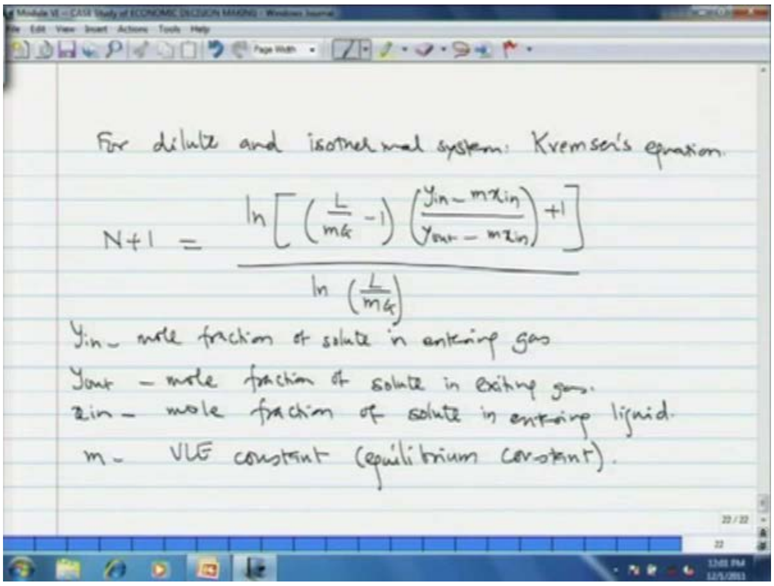

For dilute and isothermal system: Kremser's equation.

$$N+1 = \frac{\ln\left[\left(\frac{L}{mG}-1\right)\left(\frac{y_{in}-mx_{in}}{y_{out}-mx_{in}}\right)+1\right]}{\ln\left(\frac{L}{mG}\right)}$$

y_{in} – mole fraction of solute in entering gas

y_{out} – mole fraction of solute in exiting gas.

x_{in} – mole fraction of solute in entering liquid.

m – VLE constant (equilibrium constant).

Now, for a dilute and isothermal system, we can make use of Kremser’s equation to determine the number of plates of a distillation column number of theoretical plates. Kremser’s equation can be written as follows. This is you have seen in mass transfer too. So, we can take it directly the derivation is covered in the other course. Now, y in is the mole fraction of solute in entering gas. y out is the mole fraction of solute in exiting gas. x in is the mole fraction of solute that is acetone in the present case in entering liquid,

liquid entering the actually absorption column. m is the VLE constant or equilibrium constant.

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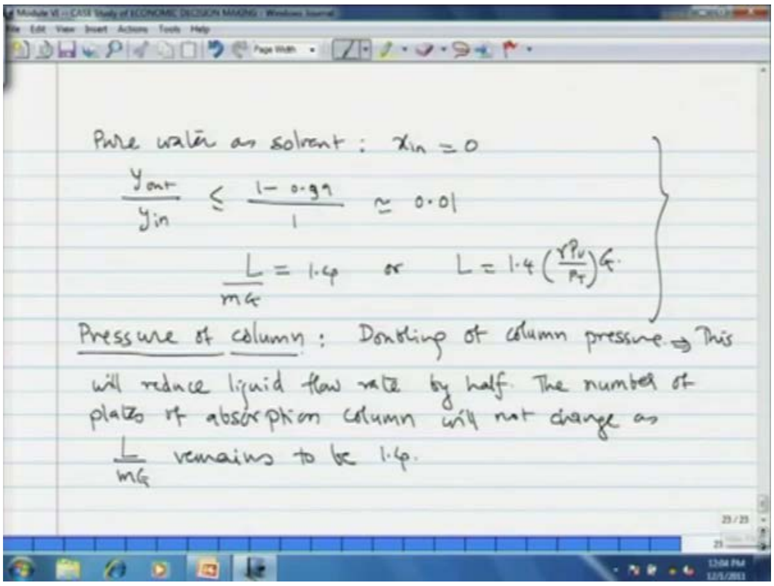

Pure water as solvent : $x_{in} = 0$

$$\frac{y_{out}}{y_{in}} \leq \frac{1-0.99}{1} \simeq 0.01$$

$$\frac{L}{mG} = 1.4 \quad \text{or} \quad L = 1.4\left(\frac{\gamma P_v}{P_T}\right)G.$$

<u>Pressure of column</u> : Doubling of column pressure. ⇒ This will reduce liquid flow rate by half. The number of plates of absorption column will not change as $\frac{L}{mG}$ remains to be 1.4.

Now, we have already chosen to use pure water as solvent and that particular choice immediately puts x in equal to 0. We have also seen recovery of 99 percent of acetone. So then we can immediately write y out by y in is either less than or equal to 1 minus 0.99 by 1 that is 0.01 or lesser. The thumb rule of having l by m G equal to 0.14 or l is equal to 1.4 into gamma PV by PT into G. Now, with this information we will try to design the column.

Now, another information that is not available with us right now is the pressure of the column. We have assumed in our previous calculation that the pressure of the column would be close to atmospheric. However, let us see what consequences the design will have if we change the column pressure. Let us say we increase the column pressure twice, that means let us say that we make PT as 2 atmospheres. This will reduce the liquid flow rate by half. Liquid flow rate will reduce two times by factor of 2.

However, L by m G remains to be 1.4. Therefore, the number of plates in the column does not change. That point we note the number of plates of absorption column will not change as L by m G remains to be 1.4. Now, what further consequences column pressure will have?

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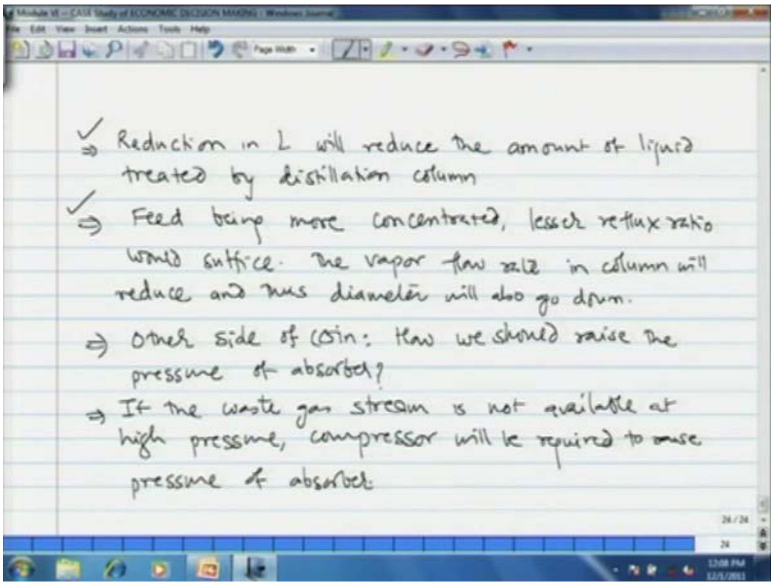

✓ ⇒ Reduction in L will reduce the amount of liquid treated by distillation column

✓ ⇒ Feed being more concentrated, lesser reflux ratio would suffice. The vapor flow rate in column will reduce and thus diameter will also go down.

⇒ Other side of coin: How we should raise the pressure of absorber?

⇒ If the waste gas stream is not available at high pressure, compressor will be required to raise pressure of absorber.

Reduction in L, the liquid flow rate will reduce the amount of liquid treated by distillation column. So, the load on distillation column will reduce. The feed will be more concentrated. Therefore, you can have a lesser reflux ratio. Feed being more concentrated; lesser reflux ratio would suffice. The vapor flow rate in column will also reduce. Thus, the diameter of the column will also go down. Thus, you can see that the operating cost is likely to be reduced significantly if P doubles the column operation.

However, we have to check the other side of coin. Other side of coin is how we should raise the pressure of operation or the pressure of absorber. If the waste gas stream is already available from the process at sufficiently high pressure, we can go for this option because we do not need additional effort to raise the pressure. However, if the pressure of the exiting gas the waste gas is not much is let us say only few thousand Pascal's above atmospheric.

Then, we need a compressor to compress the stream further and 687 moles per hour is a quiet huge volume of gas that is to be compressed. Therefore, the cost of compressor will be high. Then it will reduce all the, you know, it will reduce all the merits that are available with that particular option, that is increase in the column pressure. So, that point we note if the waste stream waste gas stream is not available at high pressure. Compressor will be required to raise pressure of absorber.

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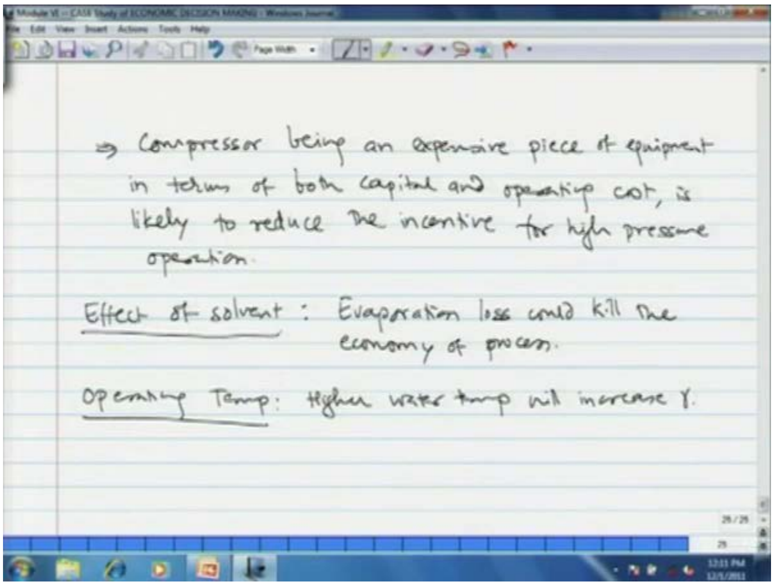

⇒ Compressor being an expensive piece of equipment in terms of both capital and operating cost, is likely to reduce the incentive for high pressure operation.

Effect of solvent : Evaporation loss could kill the economy of process.

Operating Temp: Higher water temp will increase γ.

Compressor being an expensive piece of equipment both in terms of capital and operating cost is likely to reduce the incentive for raising the column pressure. Let us see once again the effect of solvent. If you use MIBK instead of instead of water we are likely to require much low quantities of MIBK. However, the evaporation loss itself will be so high that it will kill the economy. So, that point we need to note. Evaporation loss could kill the economy of the process. Now, operating temperature that is another design variable.

Operating temperature will increase gamma the activity coefficient. That means for the same extent of absorption, you will need more water. However, L by m G being equal to 1.4, the number of plates will not reduce. So, these are some of the design variables that are associated with the Kremser’s equation or the basic like design absorber. Now, let us try to determine the number of plates of the absorber with Kremser’s equation.

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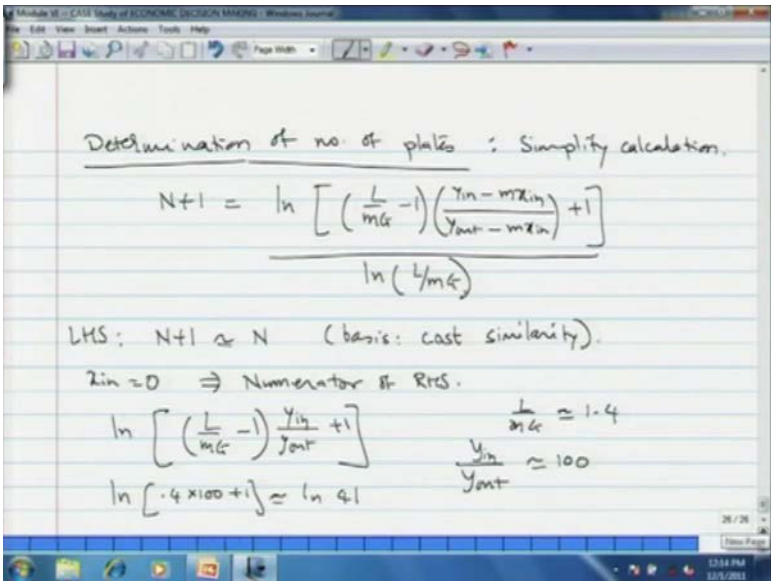

Determination of no. of plates : Simplify calculation.

$$N+1 = \frac{\ln\left[\left(\frac{L}{mG}-1\right)\left(\frac{y_{in}-mx_{in}}{y_{out}-mx_{in}}\right)+1\right]}{\ln\left(L/mG\right)}$$

LHS : $N+1 \simeq N$ (basis: cost similarity).

$x_{in}=0 \Rightarrow$ Numerator of RHS.

$$\ln\left[\left(\frac{L}{mG}-1\right)\frac{y_{in}}{y_{out}}+1\right]$$

$$\frac{L}{mG} \simeq 1.4$$

$$\frac{y_{in}}{y_{out}} \simeq 100$$

$$\ln\left[.4 \times 100 + 1\right] \simeq \ln 41$$

Our attempt will be to simplify calculation. Can we simplify Kremser’s equation, so that we can very quickly get an estimate of the number of plates required for a particular recovery? I will write once again the complete equation. Let us see whether we can simplify this equation with certain assumptions, reasonable assumptions. Now, let us look at the left hand side of this equation. We have N plus 1on the left hand side. However, if you see the cost of the column which contains, let us say, between 10 to 20 trays, the cost of column containing 10 trays or 11 trays or 20 trays or 21 trays is not going to be much different.

It will be only difference of only few percent. Therefore, we can approximate N plus 1 as N. But, the basis for this is that the cost similarity. We have already seen that x in is equal to 0. Therefore, the numerator of right hand side will immediately reduce to this, L by m G minus 1 in y in by y out plus 1. Now, let us see what numbers we have. From thumb rule, we have said L by m G is equal to 1.4 and by 99 percent recovery or even higher, we have said that y in by y out will be 100 or even more. Now, if you put these numbers here then what you get is l n 0.4 into 100 plus 1which is l n 41.

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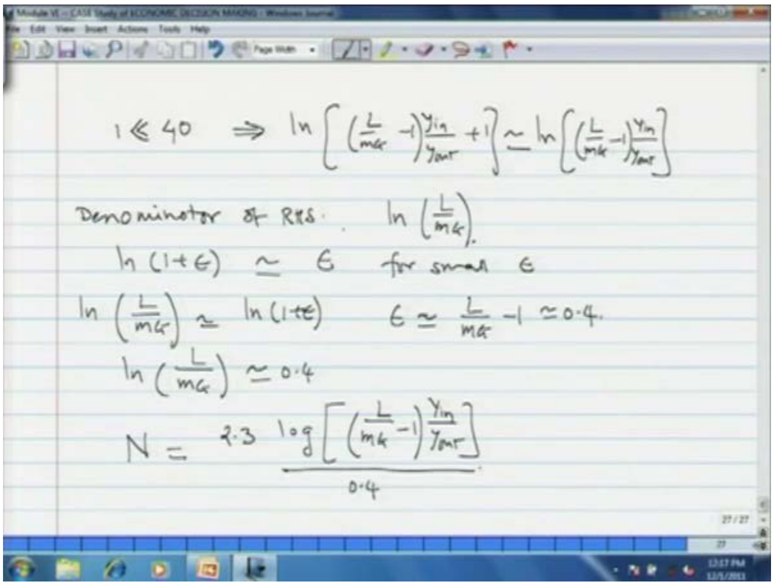

$$1 \ll 40 \Rightarrow \ln\left[\left(\frac{L}{mG}-1\right)\frac{y_{in}}{y_{out}}+1\right] \simeq \ln\left[\left(\frac{L}{mG}-1\right)\frac{y_{in}}{y_{out}}\right]$$

Denominator of RHS. $\ln\left(\frac{L}{mG}\right)$.

$$\ln(1+\epsilon) \simeq \epsilon \quad \text{for small } \epsilon$$

$$\ln\left(\frac{L}{mG}\right) \simeq \ln(1+\epsilon) \qquad \epsilon \simeq \frac{L}{mG}-1 \simeq 0.4.$$

$$\ln\left(\frac{L}{mG}\right) \simeq 0.4$$

$$N = \frac{2.3 \log\left[\left(\frac{L}{mG}-1\right)\frac{y_{in}}{y_{out}}\right]}{0.4}$$

Now, since 1 is much smaller than 40, we can approximate l n L by m G minus 1 by Y in by Y out plus 1 as l n L by m G minus 1 into Y in by Y out. We drop this plus 1. Now, let us see the denominator of right hand side which is l n L by m G. We know that l n L plus epsilon will be approximately equal to epsilon for small epsilon. As epsilon become smaller, this becomes valid.

Now, we have L by m G equal to 1.4. So, we can write an equivalent as l by l n L by m G equal to l N 1 plus epsilon, where epsilon is equal to l by m G minus 1 which is equal to 0.4. Then l n L by m G is approximately equal to 0.4. Now, with these simplifications, we can now write the new Kremser’s equation as N is equal to 2.3 log. Now, we are replacing l n by log. So, we have to multiply by 2.3. log of l by m G minus 1 into Y in by Y out. Divide by 0.4.

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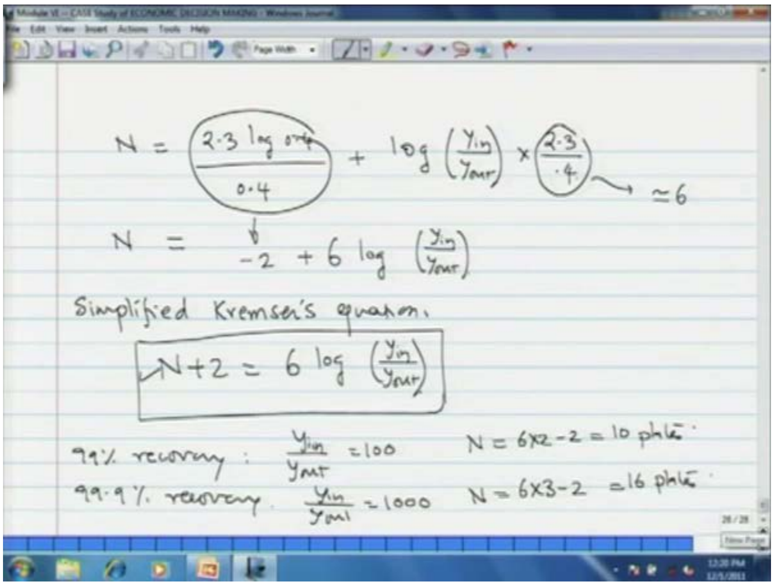

$$N = \underbrace{\frac{2.3 \log 0.4}{0.4}}_{-2} + \log\left(\frac{Y_{in}}{Y_{out}}\right) \times \frac{2.3}{.4} \rightarrow \simeq 6$$

$$N = -2 + 6 \log\left(\frac{y_{in}}{y_{out}}\right)$$

Simplified Kremser's equation.

$$\boxed{N+2 = 6 \log\left(\frac{y_{in}}{y_{out}}\right)}$$

99% recovery : $\frac{y_{in}}{y_{out}} = 100$ $N = 6 \times 2 - 2 = 10$ plate.

99.9% recovery $\frac{y_{in}}{y_{out}} = 1000$ $N = 6 \times 3 - 2 = 16$ plate.

Then, we can say that 2 point N is equal to 2.3 log of 0.4 divided by 0.4 plus log of Y in by Y out N is equal to further simplification. Now, here we have to multiply by 2.3 by 0.4 that we did not do it. Now, this particular fraction comes out to be minus 2, approximately plus 2.3 by 4. We can take approximately as 6. So, 6 log y in by y out. Therefore, we have simplified Kremser's equation as N plus 2 is equal to 6 times log y in by y out.

Of course, we have made several simplifications in this analysis. Now, you may ask me as how this simplification would deviate answer from accuracy. If you determine the number of plates for a certain gas input mole faction, gas output mole fraction x in m with all parameters, what will be the difference in number of plates determined by the simplified equation and the complete equation? I will give you some figures for 99 percent recovery.

If you use the simplified equation, you can very easily calculate the number of plates, that for 99 percent recovery, y in by y out will be equal to 100. So, N will be equal to 6 into 2 minus 2 that is 10 plates. For 99.9 percent recovery, y in by y out will be equal to 1000. So, N will be equal to 6 into 3 minus 2 that is 16 trays or 16 plates.

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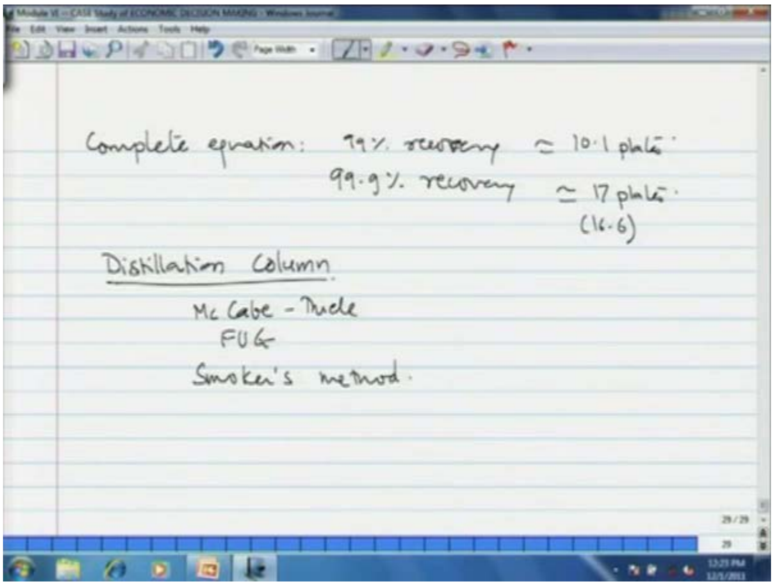

Complete equation: 99% recovery $\simeq$ 10.1 plates.
99.9% recovery $\simeq$ 17 plates.
(16.6)

Distillation Column

Mc Cabe - Thiele
FUG
Smoker's method.

Now, if you put the same numbers in complete equation, you will find that for 99 percent recovery, you will get 10.1 plates. For 99.9 percent recovery, you will get 16.6 or 17 plates. So, you can see very easily that thus, whatever simplifications we made in the analysis are not going to affect the accuracy much or I would say it is not going to affect accuracy at all. Therefore, such kind of simplifications is needed or we have to go for simplification of design equations, so as to design particular equipment to the point, where we can estimate its cost.

Now, in the present situation the cost of absorber will be determined mainly by the number of plates becomes the number of plates will determine the height of the absorber. The diameter will be determined as some other factor that we are about to see later in the short cut design procedure. So, this is how like we have designed a particular the main actual equipment of the process using a simplified expression. Now, we can use we can apply same procedure let us say, for distillation column.

We have already treated the subject of distillation column separately in another module of design of separation system. So, I am not going to elaborate here much on it. We have several equations for it. For example, the Mc Cabe Thiele diagram or the FUG method or the Smoker's method for determination of the number of plates. Now, having completed the basic design of the equipment, we shall now see how the rules of thumb are going to affect the economics of the whole process. That will be the topic treated in the next lecture.