

Lecture # 19

GAS SEPARATION TECHNOLOGIES

And their Application to CO₂ CAPTURE

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Classification of separation processes:

- Chemical Absorption (using liquid solvent)
- Physical Absorption (using liquid solvent)
- Physical Adsorption (using a solid solvent)
- Distillation (using refrigeration)
- Membranes (molecules moving across a filter)

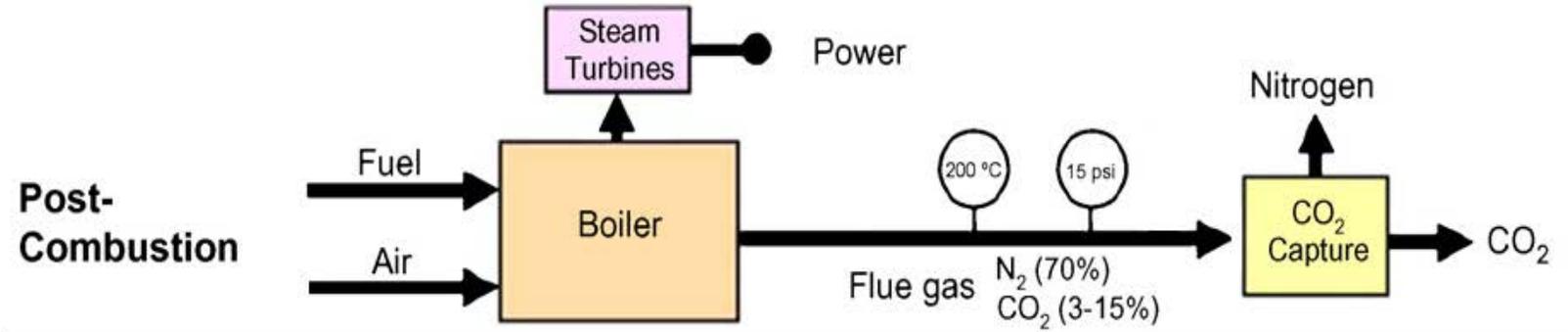
Average thermal efficiency and CO₂ production in (steam, pulverized) coal fired and (combined cycle) natural gas fired electricity generation plant, MJ_e and MW_h is the electricity generated.

	Pulverized coal (steam)	Natural gas (CC)
Efficiency	35-40%	55-60%
Heating value, MJ/kg _{fuel}	~30	~50
CO ₂ production, kg/MW _h	~1200	~400
CO ₂ specific energy, MJ _e /kgCO ₂	~3	~9

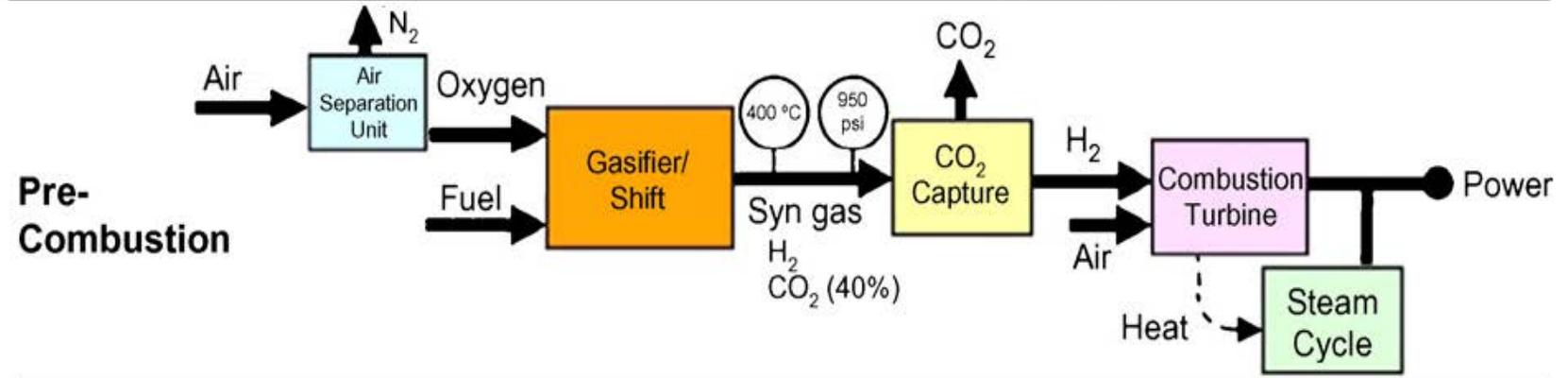
Approaches for CO₂ capture

(shown for coal used in steam cycle plants)

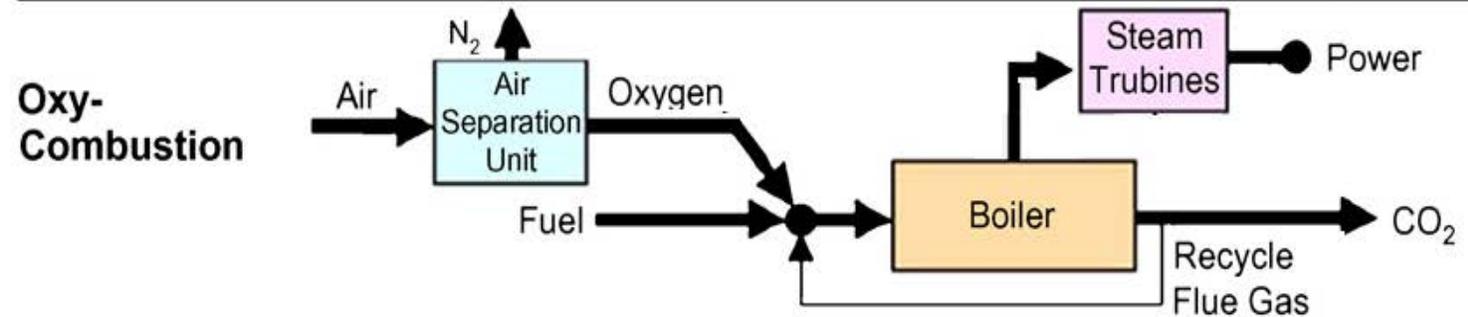
Separate CO₂ from CO₂+N₂



Separate CO₂ from CO₂+H₂

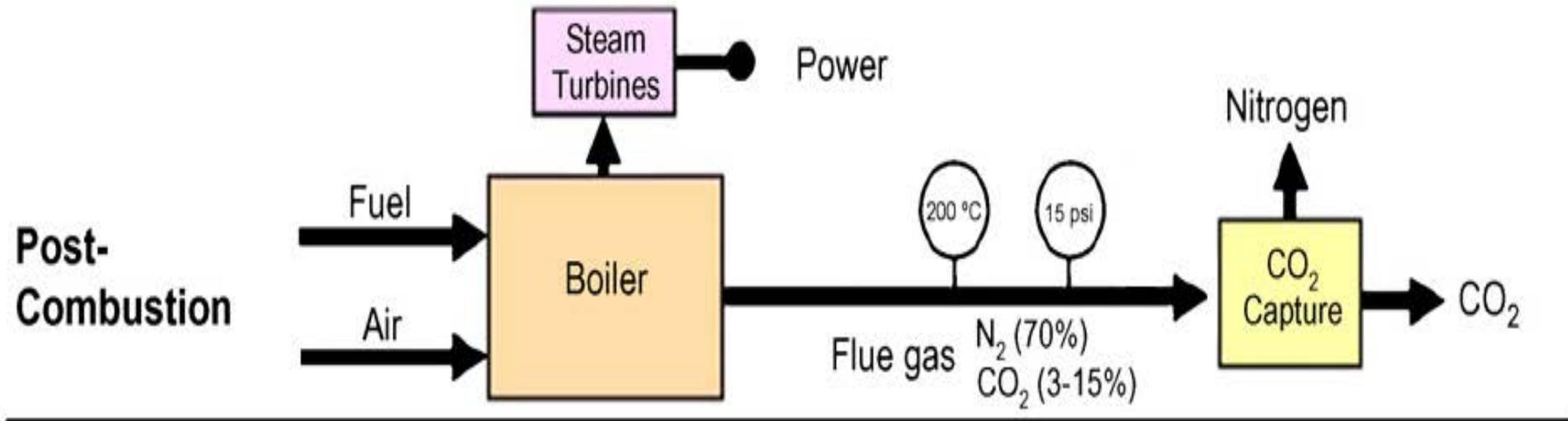


Separate O₂ from O₂+N₂



Post combustion Separation or Capture

Separate CO_2 from $\text{CO}_2 + \text{N}_2$



Courtesy Elsevier, Inc., <http://www.sciencedirect.com>. Used with permission.

The Petra Nova Project

Petra Nova plant has been operating since Jan 2017, it is the world's largest coal-fired power plant post-combustion carbon dioxide (CO₂) capture system. Since coming online it has **captured over 3.9 million short (US) tons of CO₂**, which was used to **produce over 4.2 million barrels of oil (from West ranch oil field)** through enhanced oil recovery (EOR).



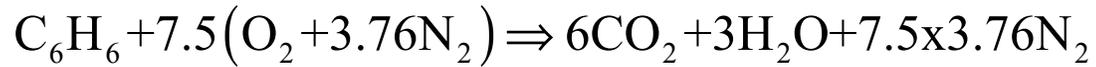
Image courtesy of DOE.

Partially funded by the U.S. DOE and managed NETL. Owned and operated by NRG Energy, Inc. and JX Nippon Oil and Gas Exploration Corporation, the Petra Nova project successfully retrofit carbon-capture technology onto a unit at the coal-fired W.A. Parish Generating Station located southwest of Houston, TX.

Petra Nova is designed to capture approximately 90 percent of the CO₂ from a 240-megawatt equivalent flue gas slipstream—which is approximately 1.6 million tons of CO₂ per year. The captured CO₂ is compressed, dried, and transported to the West Ranch Oil Field in Jackson County, TX. Then, the CO₂ is used in EOR to boost oil production.

<https://www.energy.gov/fe/articles/happy-third-operating-anniversary-petra-nova>

Estimate of the ideal and actual Work required for CO₂ Separation from combustion products in a post combustion capture process



The concentrations are: $X_{\text{CO}_2} = 0.16$ and $X_{\text{H}_2\text{O}+\text{N}_2} = 0.84$

Minimum CO₂ separation work equation: $-0.137 T_0$ MJ/kmol of benzene

At $T_0 = 300\text{K}$, the work is **41.1 MJ/kmol of benzene**.

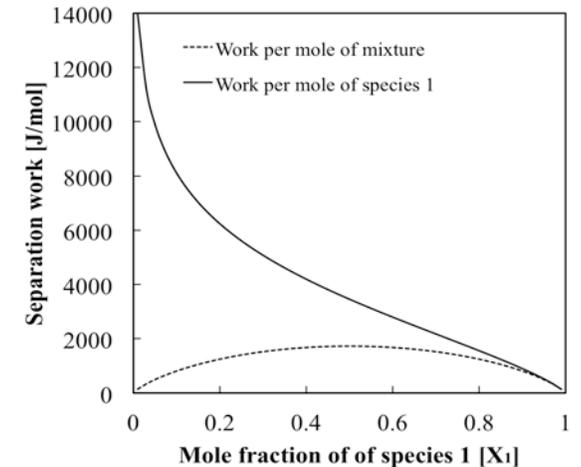
The enthalpy of reaction of benzene is 3171 MJ/kmol.

Taking 40% efficient cycle, the “useful” work produced is **1268.4 MJ/kmol of benzene**.

There is a penalty of 3.25% for the separation of CO₂ at $T = 27\text{C}$.

Actual separation processes require more work (5-10 time) due to irreversibility.

Thus, WORK penalty can be as large as **32.5% of the original work**.



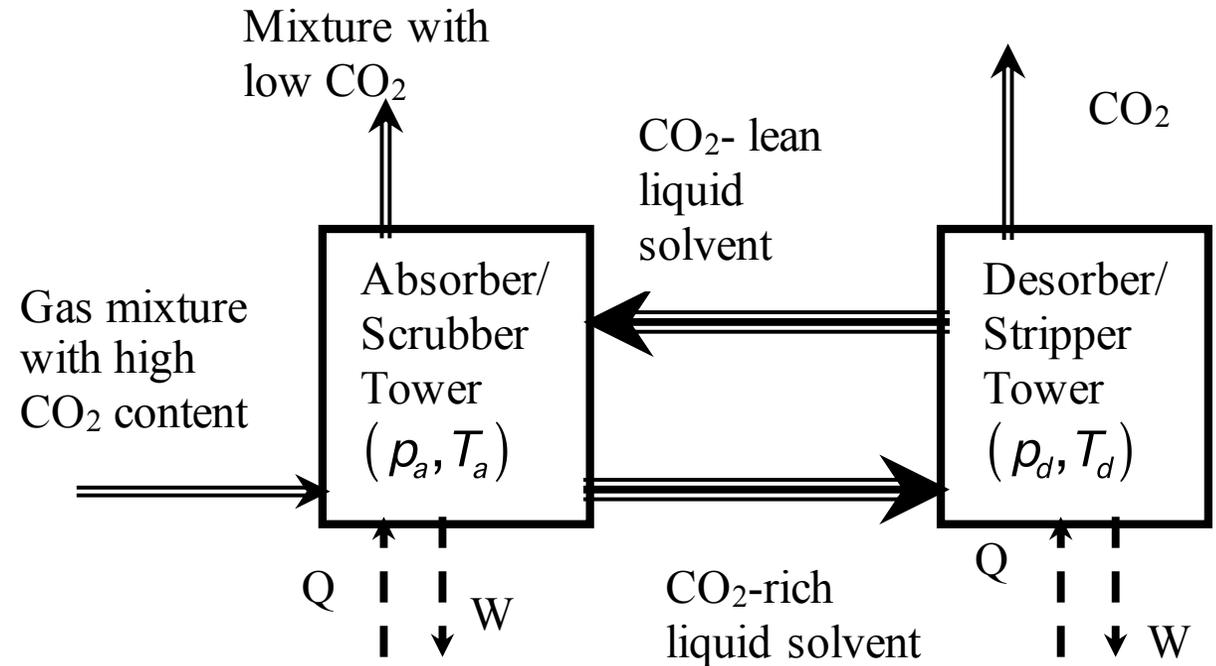
$$\dot{w}_{\text{mole of mixture}} = -\mathcal{R}T_o \left(X_1 \ln \frac{X_1}{1-X_1} + \ln(1-X_1) \right)$$

$$\dot{w}_{\text{mole of } X_1} = -\mathcal{R}T_o \left(\ln \frac{X_1}{1-X_1} + \frac{\ln(1-X_1)}{X_1} \right)$$

Separation or “capture” of CO₂ by absorption or adsorption

Relies on how much gas (CO₂) can be absorbed/adsorbed, chemically or physically, into a liquid/solid at a given temperature and/or pressure. The sorbent must be regenerated/brought back to its original form by releasing CO₂ in a different vessel and under different temperature and/or pressure.

- Separation of CO₂ by continuous absorption & desorption using a looping solvent, between two vessels (at different T/p).
- The liquid solvent can bind with CO₂ chemically or physically.
- The temperature and pressure are different in the two vessels, leading to favorable conditions for absorbing and desorbing CO₂ as the solvent loops between them



(1) SEPARATION By Chemical Absorption in a liquid solution

Works well for separating CO₂ from combustion products (mixture of N₂+CO₂)

use a reversible reaction whose equilibrium is temperature dependent

for an "amine solution" $2\text{R-NH}_2 + \text{CO}_2 \leftrightarrow \text{R-NH}_3^+ + \text{R-NH-COO}^-$

or $\text{CO}_2\text{-lean chemical solvent} + \text{CO}_2 \xrightleftharpoons[\text{desorber @ high T}]{\text{absorber @ low T}} \text{CO}_2\text{-rich chemical solvent}$

o forward absorption reaction is exothermic and lower temperatures favor absorption (products)

$$\frac{[\text{R-NH-COO}^-]}{[\text{R-NH}_2]} \text{ is proportional to } K_p(T) = \exp\left(-\frac{\Delta G_R}{\mathfrak{R}T}\right) = e^{\Delta S_R/\mathfrak{R}} \exp\left(\frac{|\Delta H_R|}{\mathfrak{R}T}\right) \uparrow \text{ as } T \downarrow$$

o lower T ~ 50 C, equilibrium favors products' formation or CO₂ absorption into the amine solution

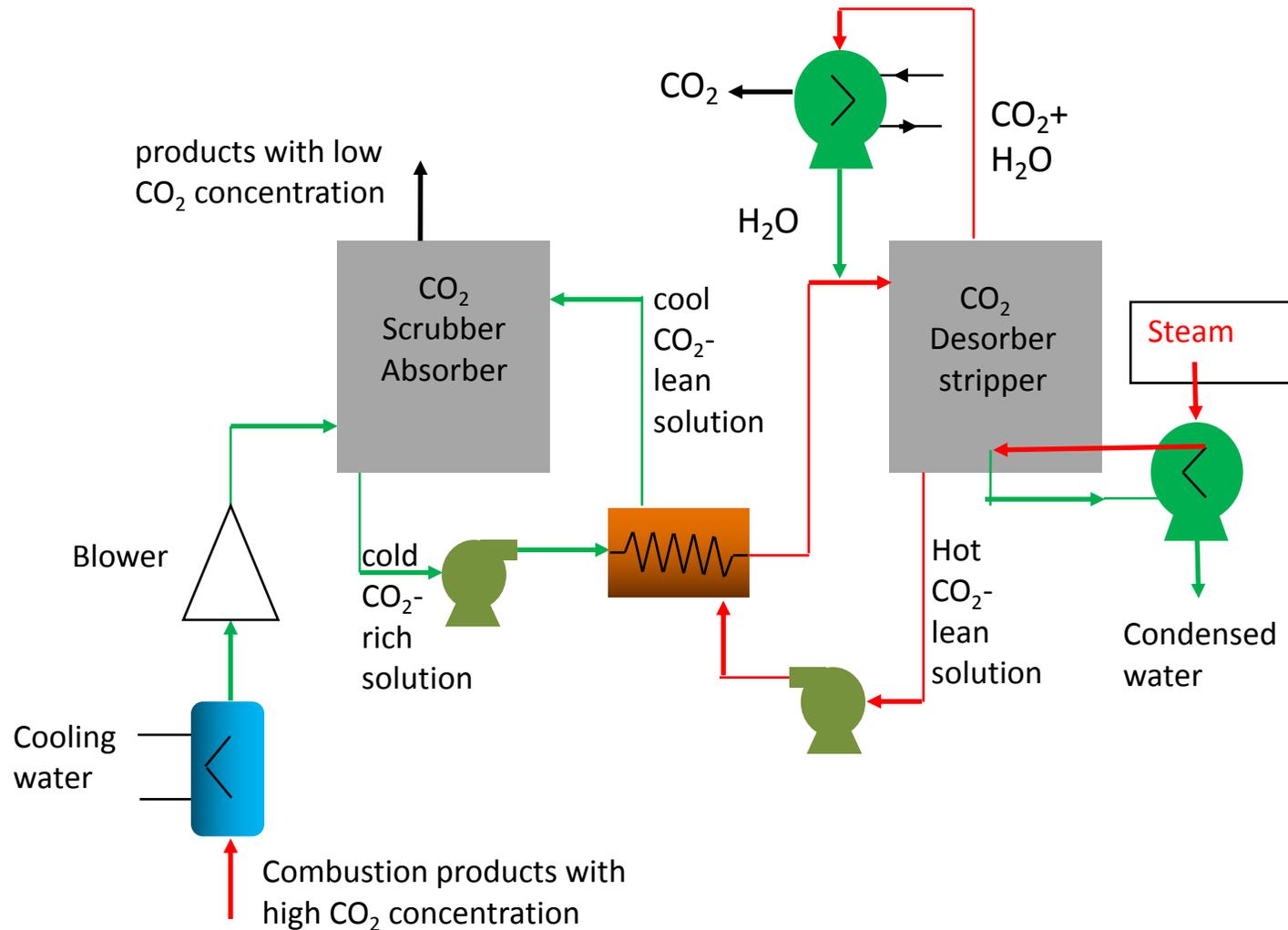
Must keep the absorber "tower" cool (reject "low quality" heat) to increase CO₂ in the amine phase.

o opposite is true at higher T ~ 150 C, the reaction reverses, and CO₂ is desolved from the solution

Need to heat up the solvent to release CO₂ and regenerate the solvent.

for R ≡ (-CH₂CH₂OH), monoethanolamine (MEA), $\Delta H_R = -1.919 \text{ MJ/kgCO}_2$

Using heat recuperation to reduce the heat required in the desorber (heat duty)



Energy required by the separation plant:

energy for separation
 = enthalpy of the desorption reaction
 + thermal enthalpy of solution (water)
 + enthalpy of evaporation of vaporized water
 + enthalpy of evaporation of vaporized absorbent
 + enthalpy of desolution of CO₂ from water
 + *pumping work*

Most important components are:

$$Q_{recovery} = n_{solution} \dot{\epsilon}_{solution} \Delta T + n_{water} \Delta \dot{h}_{evap} + n_{sorbent} \left(\Delta \dot{h}_{desorption} + \Delta \dot{h}_{CO_2, sol} \right)$$

Total heat required for separation (simplified calculations)

Burning 1 kmole of coal generates one kmole of CO₂, and ~ 360 MJ of heat.

For each kmole of CO₂, 2 kmoles of MEA are needed. With 10% sorbent molar concentration in water (or 0.27 by mass, MW_{sorb}=61), one needs 7.3 kmoles solvent.

The reaction energy of desorption is 1.919 MJ/kgCO₂, or (1.919x44)= **84.5 MJ** for 1 kmole CO₂.

But some water evaporates, $\Delta h_{\text{evap}} \sim 2.25$ MJ/kg_{water}. Assume ~ 1 kmole of water evaporates per mole CO₂ removed, we need (2.25x18)= **40 MJ**.

Thus, the total thermal energy in the reboiler is ~ 84.5+40= **124.5 MJ**.

Thus ~ (124/36) or one third of the original thermal energy is needed for CO₂ capture.

More energy is used to raise the solvent T (some comes from regeneration), and for pumping the solvent through the loop. The work penalty is less because this is “lower quality heat”.

APPLICATIONS: Chemical scrubbing of CO₂ from flue gases has been used extensively:

During 1982-1986, an aqueous solution of MEA was used in: Lubbock Power plant, Texas, NG was fired in a 50 MW plant, producing near 1000 t/d of CO₂ (for EOR)

And in a coal-steam generator in Carlsbad NM producing 113 t/d (also EOR)

Since 1991, CO₂ scrubbing using 15-20% MEA solutions in the 300 MW Shady Point CHP Plant in Oklahoma, producing nearly 400 t/d CO₂, used in food industry and EOR.

Norway Sleipner Vest gas field separates CO₂ from the recovered natural gas to reduce CO₂ concentration in the produced gas from 95% to 2.5%. The separated CO₂ is then injected back into a 250 m deep aquifer located 800 m below the ocean surface.

A similar project in Indonesia in which the CO percentage of CO₂ in the recovered gas will be reduced from 71 to almost zero.

In Algeria, the In Salah field

Malaysia



North Dakota



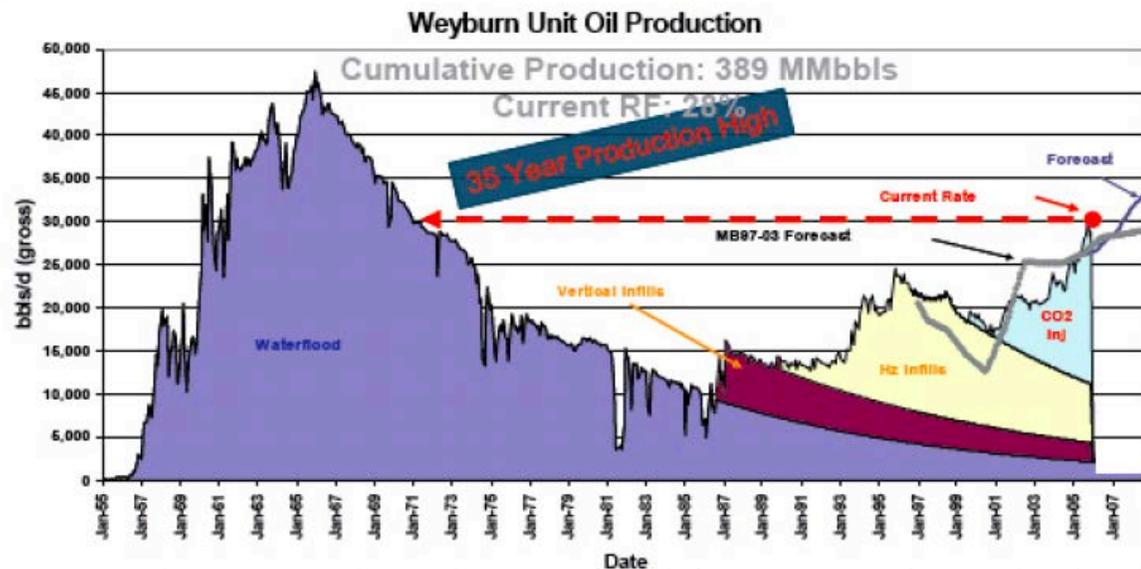
Figure TS.4. (a) CO₂ post-combustion capture at a plant in Malaysia. This plant employs a chemical absorption process to separate 0.2 MtCO₂ per year from the flue gas stream of a gas-fired power plant for urea production (Courtesy of Mitsubishi Heavy Industries). (b) CO₂ pre-combustion capture at a coal gasification plant in North Dakota, USA. This plant employs a physical solvent process to separate 3.3 MtCO₂ per year from a gas stream to produce synthetic natural gas. Part of the captured CO₂ is used for an EOR project in Canada.

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EOR and the Dakota Gasification Company

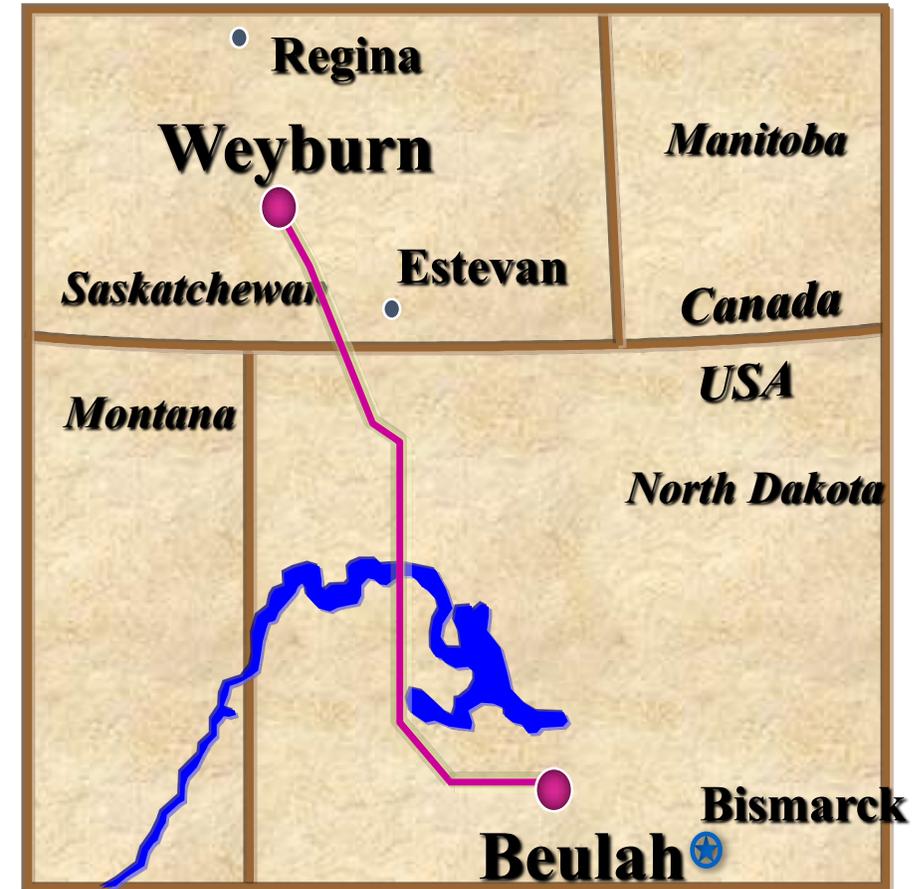
250 million cu.ft. per day CO₂ by-product of coal gasification
95 million cu.ft. per day contracted for Weyburn project
320 km pipeline
CO₂ purity 95%
CO₂ pressure 14MPa

Weyburn Oil Production

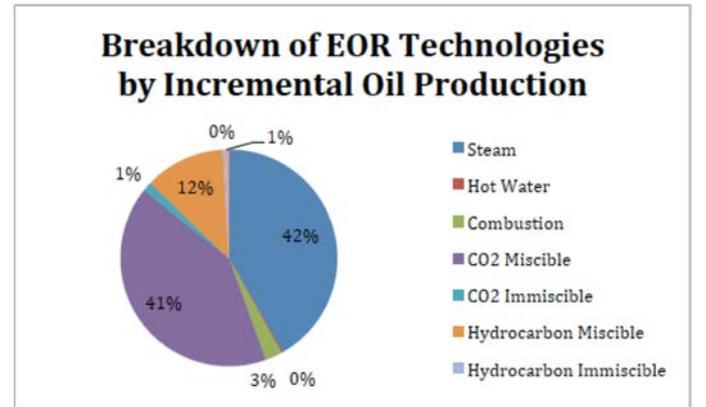
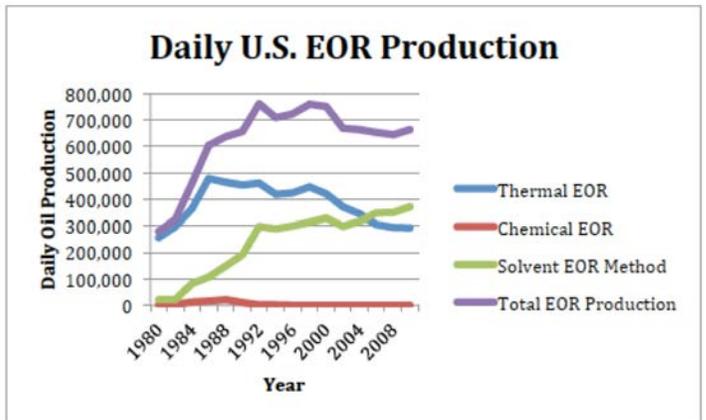
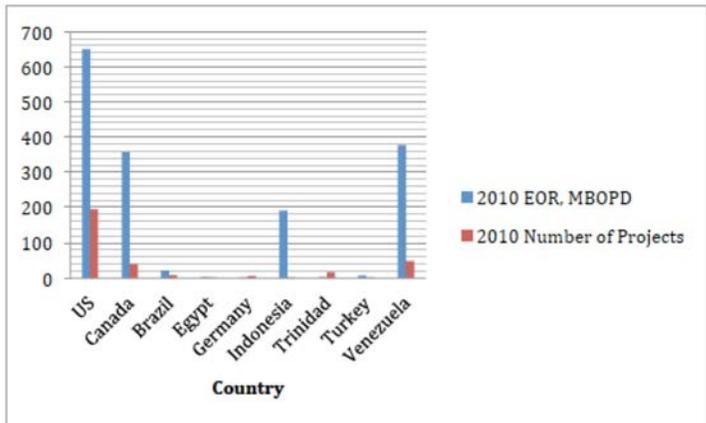


Source: EnCana Corporation (via Don Langley, B&W)

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Koottungal "Special Report: 2010 worldwide EOR survey," the oil and gas J., 14 (2010)

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Carbon Dioxide Pipelines in Operation in North America

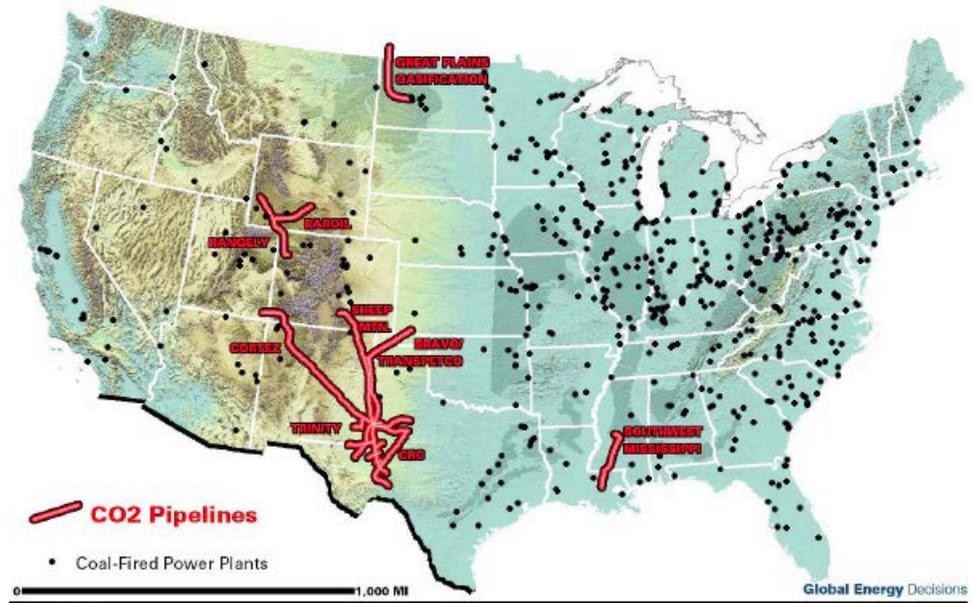
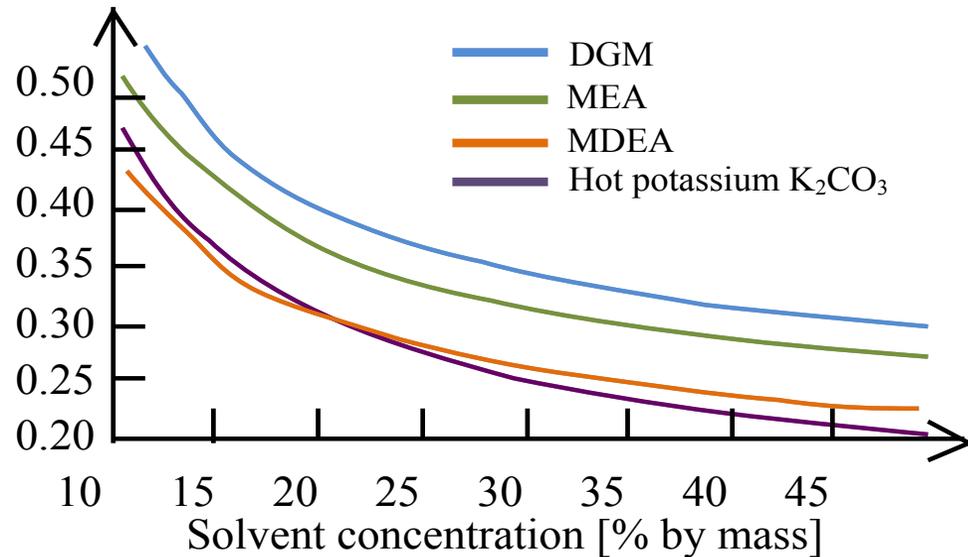


Figure 6.3

Coal: America's Energy Future, The National Coal Council, March 2006.

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Sorbents with lower desorption reaction enthalpy are under development

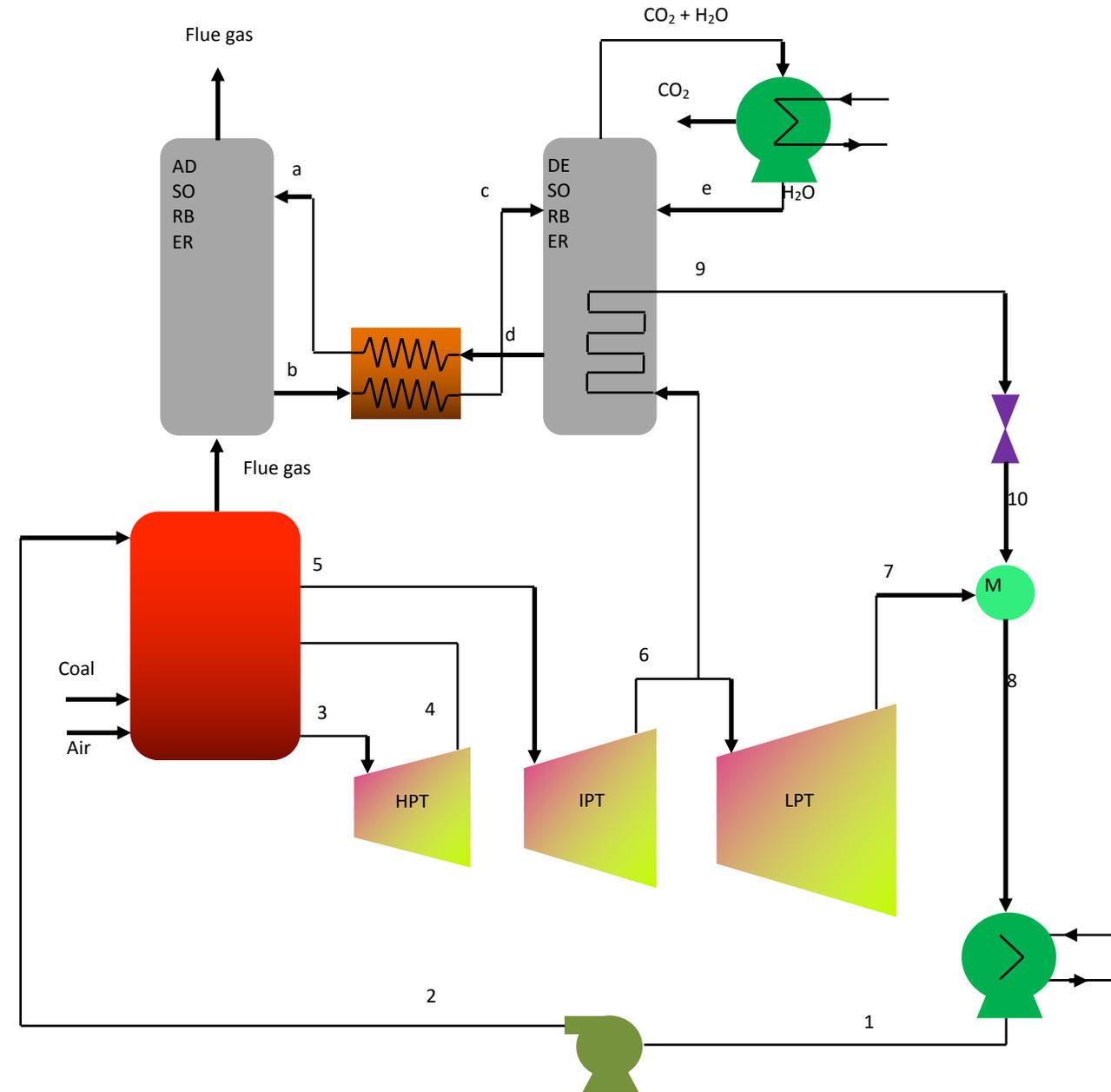


		MJ/kg_CO2
Monoethanolamine (MEA)	$R \equiv (-CH_2CH_2OH)$	1.919
Diethanolamine (DEA)	$R = (-CH_2CH_2OH)_2$	1.519
Triethanolamine (TEA), ₃		0.989
Methyldiethanolamine (MDEA), ₂		1.105

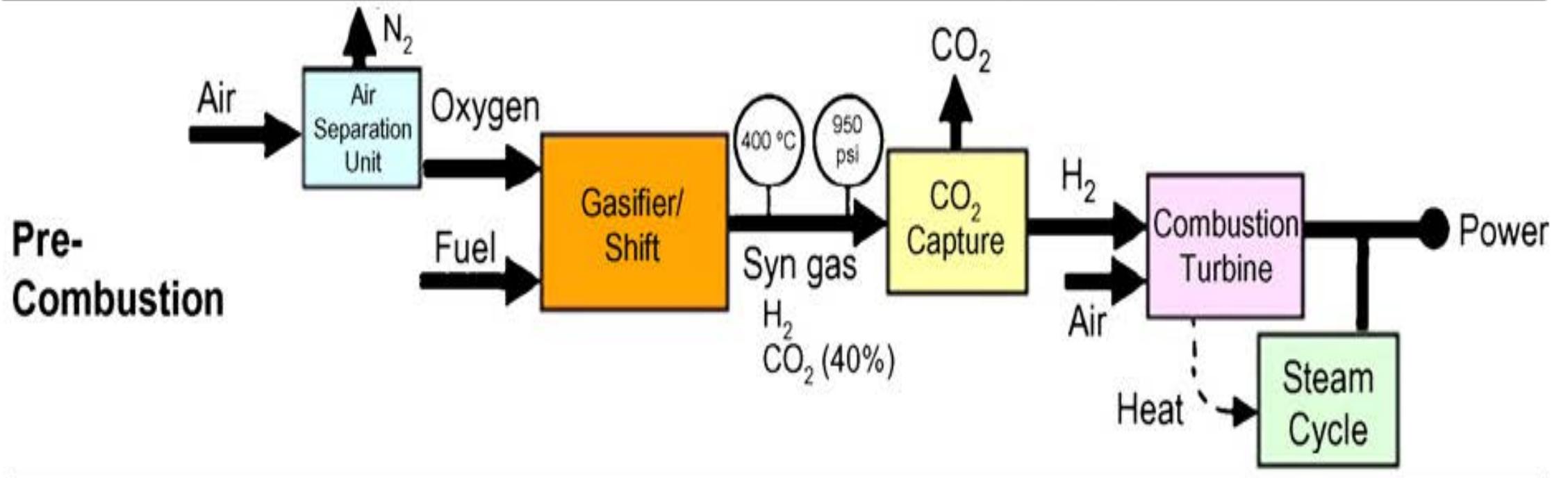
Energy penalty for CO₂ separation using chemical absorption expressed as work (exergy) or reduction of electricity production in kWh/kgCO₂. Increasing the concentration of the sorbent in the solution reduces the energy required for CO₂ removal. With higher sorbent concentration, more energy is expended in the desorption reaction enthalpy than in heating and evaporating the solvent. However, higher sorbent concentration can be corrosive, and mass concentrations below 30% have been recommended (kWh = 3.6 MJ). DGA is known as Econamine (n diglycolamine in aqueous solution), MDEA is 2n mythyldiethanolamine.

Integration of Separation with the Power Cycle

- Works by extracting steam at lower T (~ 150 C), i.e., from lower pressure sections of steam turbine, to reduce exergy losses in the cycle.
- MAE works with (in the presence of) SO_x .
- For monoethanolamine, or MEA, a SO_x level of less than 10 ppmv is desired for the Fluor Daniel EconamineTM process.
- A level of 50-100 ppmv can be tolerated by the KEPCO/MHI process



Approaches for CO₂ capture: pre-combustion capture



Separate CO₂ from CO₂+N₂

(2) SEPARATION by Physical Absorption (in a liquid)

Works well for separating H₂+CO₂ mixture (pre-combustion capture)

- In a given solvent, different gases have different solubility (different values of Henry's constant), and hence get absorbed at different fractions by the liquid solvent.
- "Selectivity" is measured by the relative value of the solubility, or He.
- Higher pressure and lower temperature favors absorption.
- Lowering the pressure or raising the temperature of the solvent causes the gas to escape.

In gas-liquid equilibrium (at low concentration)

$$X_{L,j} = \frac{p_j}{He_j} = \frac{p}{He_j(T)} X_{V,j}$$

$X_{L,j}$ Mole fraction of gas "j" dissolved in liquid

p_j Gas "j" partial pressure in the gas phase

p Total pressure of gas mixture in contact with liquid

$X_{V,j}$ "j" mole fraction in gas mixture in contact with liquid

He_j Henry's constant of gas "j" in the liquid

Henry's constant for a number of gases in methanol, measured in MPa/(mol/kg methanol), at two different temperatures. It decreases rapidly with T

gas	- 40 C	- 70 C
Hydrogen sulfide H ₂ S	0.28	0.026
Carbon dioxide	1.4	0.31
methane	70	30
hydrogen	670	930

A continuous separation process using physical absorption

the fraction of CO₂ in liquid is: $X_{L,CO_2} = \frac{p_{CO_2}}{He_{CO_2}} = \frac{X_{V,CO_2}}{He_{CO_2}} p$

this fraction increases significantly at high p and low T (low He)

If all the CO₂ (fraction $X_{V,CO_2,in}$) in the gas stream $\dot{n}_{prod,in}$ is absorbed by solvent \dot{n}_{sol} (leaving with $X_{L,CO_2,out}$):

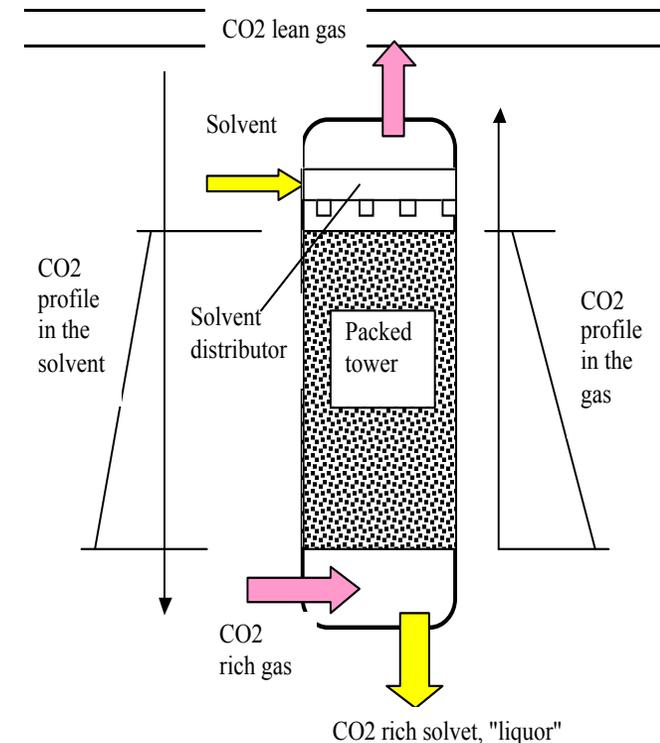
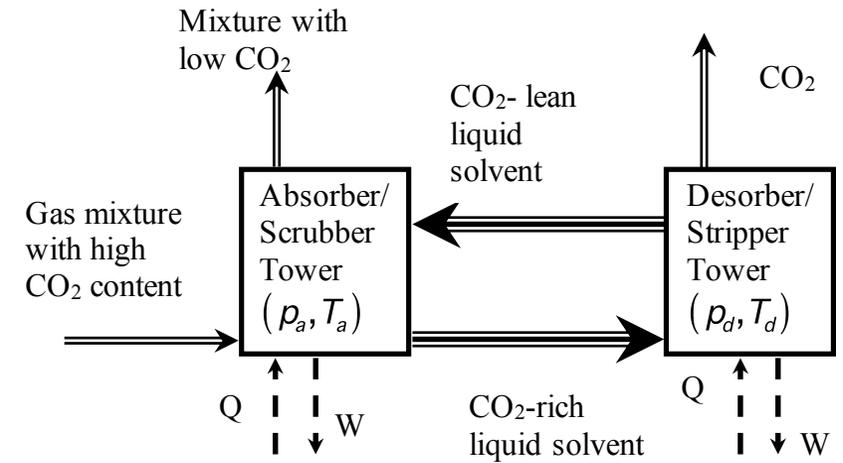
$$\dot{n}_{prod,in} X_{V,CO_2,in} = \dot{n}_{sol} X_{L,CO_2,out}$$

The required solvent flow rate (from the two equations):

$$\dot{n}_{sol} = \frac{He}{p} \dot{n}_{prod,in}$$

Power required to move fluid: $\mathcal{P}_{pump} = \dot{V}_{sol} \frac{\Delta p}{\eta_{pump}}$

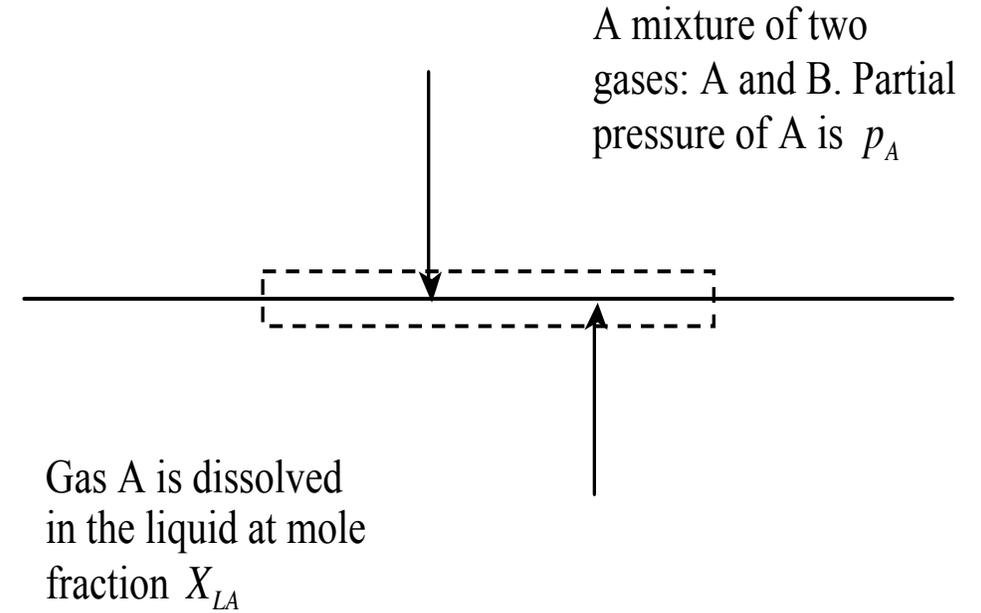
$$\Delta p_{total} = \Delta p_{between\ tanks} + \Delta p_{across\ Absorber} + \Delta p_{across\ Desorber}$$



$$X_{L,A} = \frac{p_A}{He_A} = \frac{p}{He_A(T)} X_{V,A}$$

Henry's constant (in bars) in water at low to moderate pressures

Solute	290 K	300 K	310 K	320 K	330 K	340 K
H ₂ S	440	560	700	830	980	1140
CO ₂	1280	1710	2170	2720	3220	-
O ₂	38000	45000	52000	57000	61000	65000
H ₂	67000	72000	75000	76000	77000	76000
CO	51000	60000	67000	74000	80000	84000
Air	62000	74000	84000	92000	99000	104000
N ₂	76000	89000	101000	110000	118000	124000



- Higher He means that less gas is absorbed in the liquid at the given temperature
- As T decreases, more gas is absorbed, and vice versa (temperature swing processes)
- Water absorbs a lot more CO₂ than nitrogen, i.e., it is more selective to CO₂.
- Gas can also be desorbed from the liquid by reducing the pressure (pressure swing processes)

Changing the pressure to release the gas, Pressure Swing Absorption (PSA):

@ 10 atm and -70 C , a gas mixture in which the molar concentration of CO_2 is 10% is in equilibrium with a methanol solvent in which the molar concentration of CO_2 is (He is in $\text{MPa}/(\text{mol}/\text{kg_methanol})$) is:

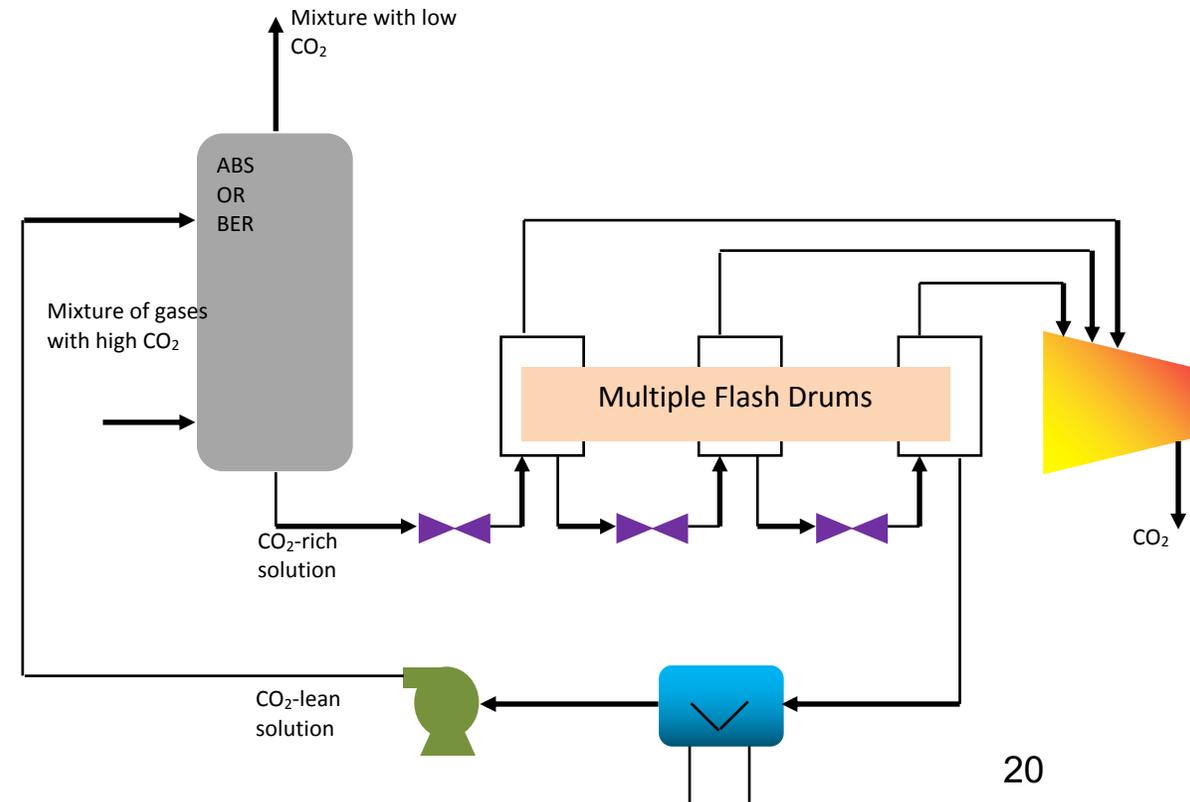
$$X_{L,\text{CO}_2}(1\text{ MPa}, -70\text{C}) = \frac{X_{V,\text{CO}_2} = 0.1 \cdot p = 1}{He = 0.31} = 0.32\text{ mol}/\text{kg_methanol}.$$

If this solvent is transported to another vessel in which the pressure is 0.1 atm (pure CO_2 , $X_{V,\text{CO}_2}=1$), the equilibrium concentration of CO_2 in methanol drops to

$$X_{L,\text{CO}_2}(0.01\text{ MPa}, -70\text{C}) = \frac{X_{V,\text{CO}_2} = 1 \cdot p = 0.01}{He = 0.31} = 0.032\text{ mol}/\text{kg_methanol}$$

and 0.288 mol of CO_2 per kg of solvent is released.

- (1) Methanol, used in **the Rectisol process**;
- (2) N-methyl-2-pyrrolidone (NMP), used in **the Purisol process**;
- (3) Dimethylether polyethylene glycol (DMPEG), used in **the Selexol process**.



Changing the pressure and temperature to release the gas

@ 10 atm and -70 C , a gas mixture in which the molar concentration of CO_2 is 10% is in equilibrium with a methanol solvent in which the molar concentration of CO_2 is

$$X_{L,\text{CO}_2}(1\text{ MPa}, -70\text{ C}) = \frac{X_{V,\text{CO}_2} = 0.1 \cdot p = 1}{He = 0.31} = 0.32 .$$

If pressure is reduced to 1 atm and the temperature is raised to -40 C , the new equilibrium concentration of CO_2 in the solvent is,

$$X_{L,\text{CO}_2}(0.1\text{ MPa}, -40\text{ C}) = \frac{p = 0.1}{He = 1.4} = 0.071 ,$$

and 0.249 moles CO_2 is released per kg_methanol.

This is the pressure-temperature swing separation (PTSA).

(3) Physical Separation by Adsorption (in a solid) with Pressure or Temperature Swing

Langmuir Isotherm for gas-solid surface equilibrium

$$\theta_i = \frac{p_i}{1 / K_i(T) + p_i}$$

θ_i : fraction of occupied sites by gas "i" on solid surface.

p_i : partial pressure of gas "i" in contact with solid surface.

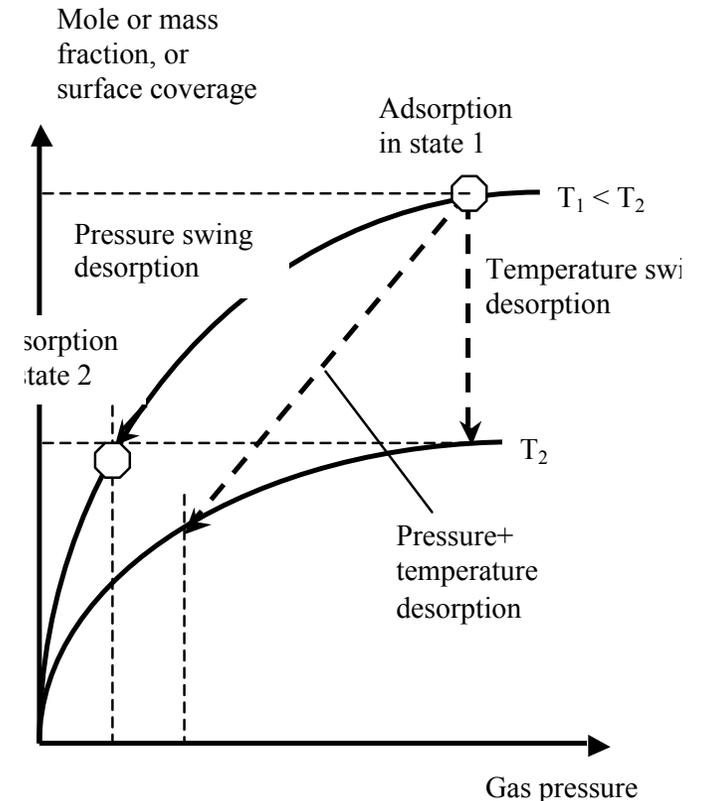
At low p_i , $\theta_i \approx K_i(T) p_i$ and at high p_i , $\theta_i \approx K_i(T)$

In terms of C_i is the concentration of gas in solid,

$$C_i = \frac{\alpha_s p_i}{1 / K_i(T) + p_i} = \frac{\alpha_s X_i}{1 / p K_i(T) + X_i}, \text{ where } \alpha_s = (A_s / \nabla) \tilde{\Gamma} / N_a$$

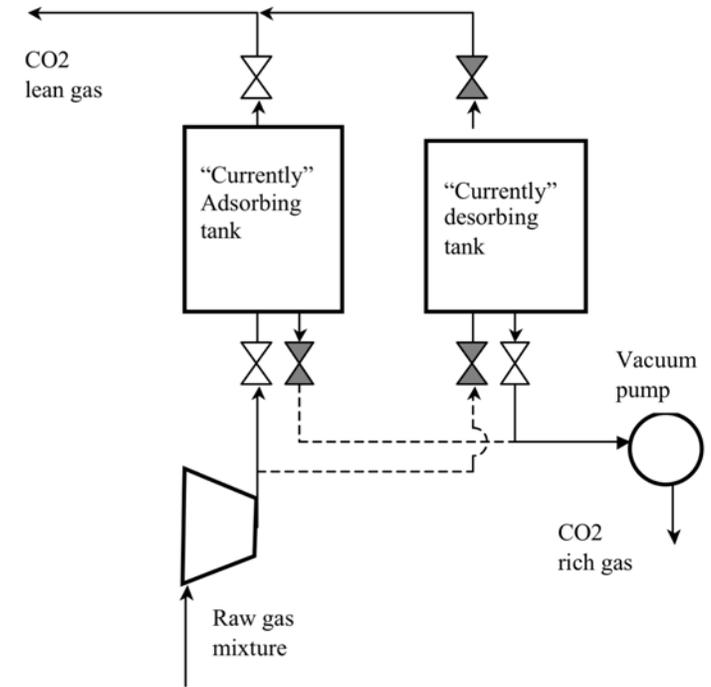
A_s / ∇ : surface area/volume, and $\tilde{\Gamma}$: number of site/area

the gas concentration on the solid depends on $K_i(T)$ and p



Solid Absorbers

- **Activated carbon**, manufactured from carbonaceous matter such as coal, petroleum coke and wood, are used in the powder form.
- Activation involves carbonization, i.e., elimination of volatile matter and partial gasification to develop the porosity and enlarge surface area.
- The pore surface area is high, 300-2500 m²/g, largest among all known adsorbents.
- The pore diameter in activated carbon is 10-35 Å.
- The heat of adsorption of activated carbon is generally lower than that of other adsorbents, and stripping the adsorbate requires less energy.
- Activated carbon is nearly hydrophobic to low humidity mixtures, and hence can be used in separating humid gas mixtures.



Compatible with availability of high p stream as in IGCC, pre-combustion capture application.

- **Molecular sieves carbon (MSC)** is produced by carbonization of polymers such as cellulose and sugar, and other methods.
- The pore structure and pore diameter depends on the original material and the carbonization temperature diameters in the range of 2-8 Å have been demonstrated.
- These are used in ASU's and in other applications in the chemical industry.
- Carbon (3%) deposited into the pores of lignite char by cracking methane at 855 C produces MSC that shows significant molecular sieving between CO₂ (admitted) and N₂ (hindered). CO₂ is a linear molecule that is thought to have a diameter of 3.7 Å.

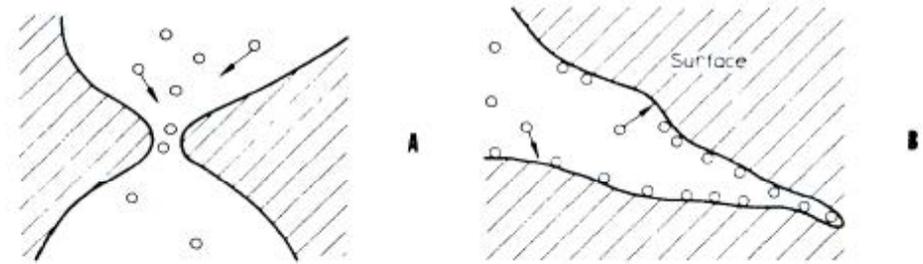


Figure 2.5 Molecular-sieve carbons made by Bergbau-Forschung: (A) Type CMSN2 with bottlenecks near 5 Å formed by coke deposition at the pore mouth; (B) Type CMSH2 formed by steam activation. Source: Jüntgen, Knoblauch, and Harder [34]. Reprinted with permission.

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Yang, R.T., Gas Separation by Adsorption Processes, Imperial College Press, 1997

- **Zeolites** are crystalline aluminosilicates of alkali or alkali earth elements such as sodium, potassium and calcium.
- The basic structure of zeolite is an aluminosilicate skeleton, with a window aperture of 3-10 Å.
- Sorption into zeolites can occur with great selectivity because of the size of the window, and they can also act as molecular sieve.
- Zeolites come in a variety of structures (see figure) depending on how they are manufactured.
- They have been used in hydrogen purification and air separation.

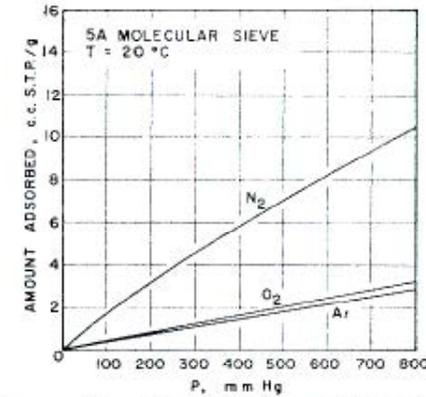


Figure 7.20 Isotherms on 5A zeolite [57]. The selectivity ratio for N_2/O_2 declines gradually to slightly less than 2 as the pressure is increased to 8 atm [58]. Source: Lee and Stahl [20]. Reprinted with permission.

Gottlicher, G., The Energetics of Carbon Dioxide Capture in Power Plants, The US DOE, Office of Fossil Energy, NETL, Feb 2004

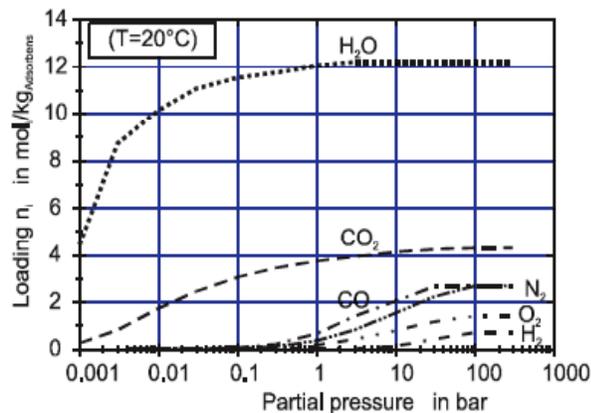
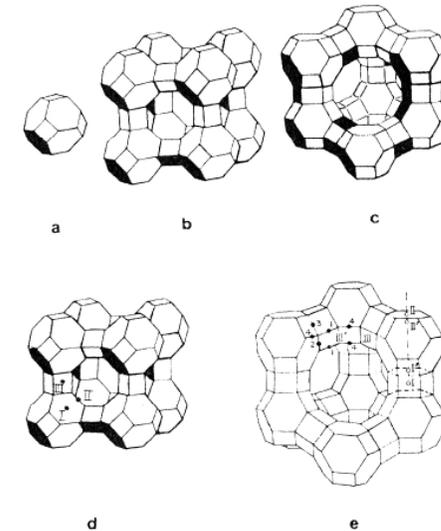


Image courtesy of DOE.

Figure 3.17: Adsorption isotherms of the pure components for zeolite 5A molecular sieve at 20°C ⁴²

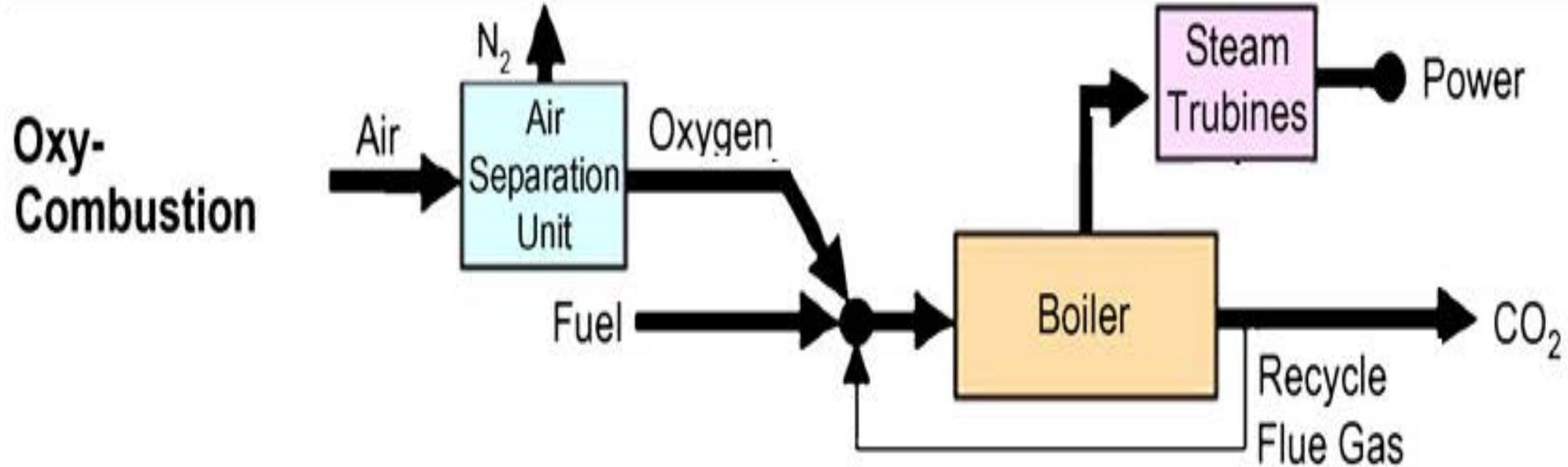
Amount of gas adsorbed in zeolite molecular sieve, in mol/kg_ adsorbent @ 20 C for different partial pressure of the adsorbed gas. Trends follow Langmuir isotherm and show CO₂-H₂ selectivity.



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Yang, R.T., Gas Separation by Adsorption Processes, Imperial College Press, 1997

Approaches for CO₂ capture: Oxy-combustion capture



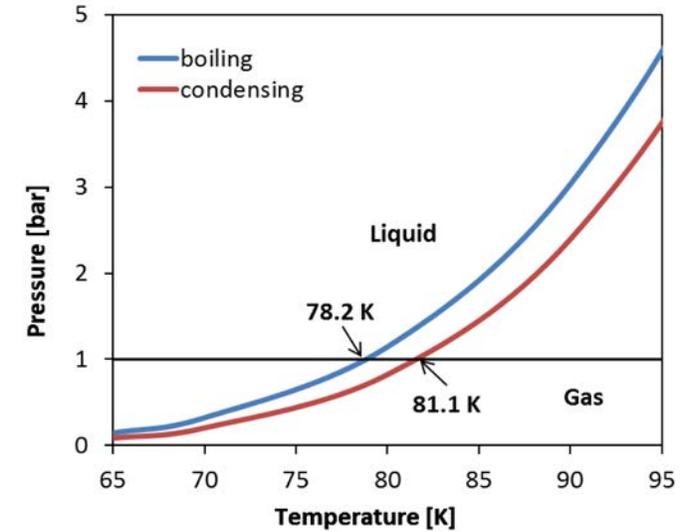
Separate O₂ from O₂+N₂

Liquefaction & Cryogenic Air Separation

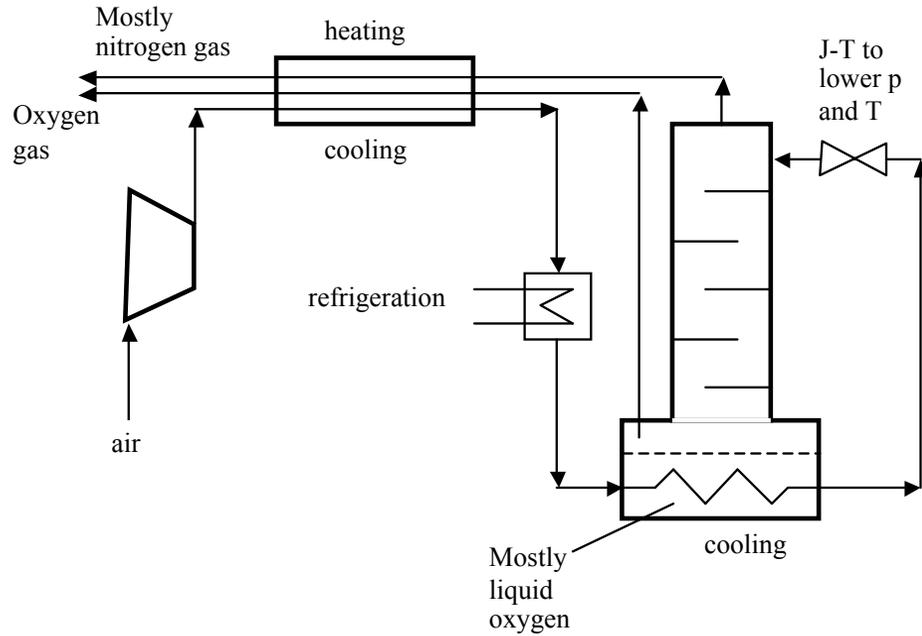
Ref: <http://www.thermopedia.com/content/553>

At atmospheric pressure, the boiling/condensation temperature of oxygen is 90 K (-183 C) and of nitrogen is 77 K (-196 C) (nitrogen is more volatile than oxygen as it evaporates at lower temperature).

When air is cooled at atm. pressure, it remains gas till 81.6 K. and completely liquefy at 79 K (inverse when heated)



Two figures have been removed due to copyright restrictions. Please see the figures on page: <http://hyperphysics.phy-astr.gsu.edu/hbase/thermo/liqair.html>.



A column distillation cycle for air separation. Air is compressed, then cooled in a regeneration unit using the separation products (N_2 and O_2), and a refrigeration unit. A Joule-Thomson valve lowers p and T , i.e., flashes air into a distillation column to separate oxygen and nitrogen. Energy expended is in air compression and refrigeration.

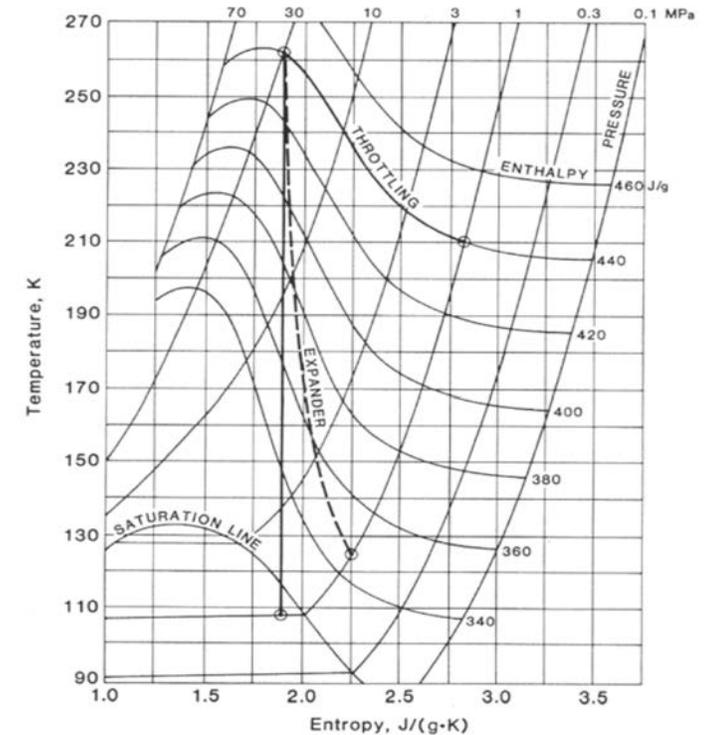


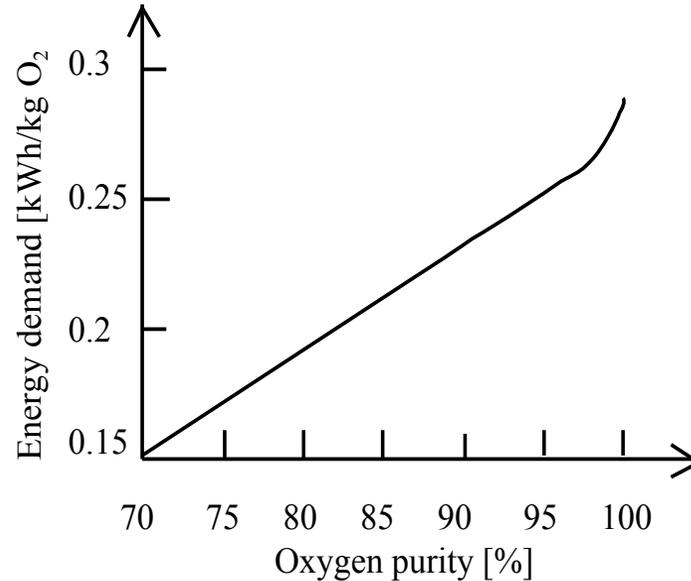
Figure 5.4 Comparison for air of Joule-Thomson cooling with cooling by a mechanical expander.

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Air Separation energy ~ 1 MJ/kg_O2

The energy requirement for air separation depends strongly on the purity of the products streams. To achieve 90% oxygen purity (volume fraction) air must be pressurized to 5 bars. Higher purity requires pressure up to 10 bar. The products are delivered at pressures and temperatures close to the atmospheric values.

Ideal work for air separation is 0.104 kWh/kg_O2, or 0.374 MJ/kg_O2. Air separation efficiencies in cryogenic plants are low, 15-30%.



Energy required to produce oxygen in a large scale double column facility.

Gotticher, G., The Energetics of Carbon Dioxide Capture in Power Plants, The US DOE, Office of Fossil Energy, NETL, Feb 2004

Image courtesy of DOE.

©Ahmed F. Ghoniem

R.F. Probstein and R. E. Hicks, *Synthetic Fuels*,

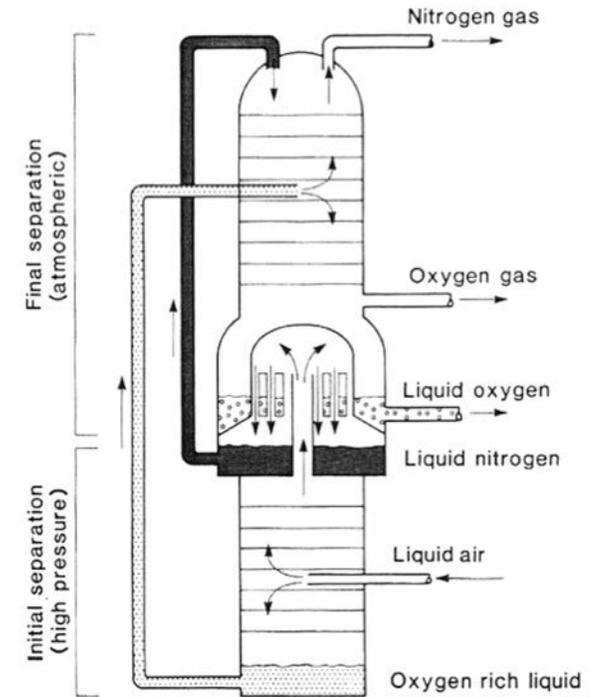
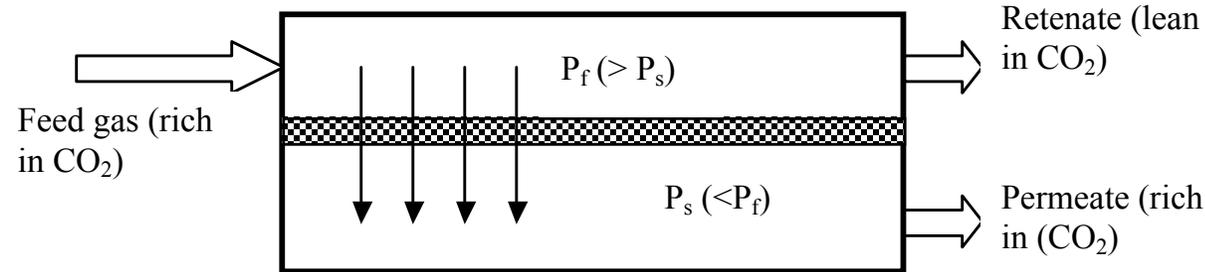


Figure 5.6 Distillation column for fractional separation of liquid air (after Ref. 11).

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(2) Membrane Separation



For porous membranes, the flux of one gas is

$$J_i = \frac{D_{ei}}{\mathfrak{R}T \ell} \Delta p_i = \tilde{Q}_i (p_{i,f} - p_{i,s}) = \tilde{Q}_i (X_{i,f} p_f - X_{i,s} p_s) = \frac{\tilde{q}}{\ell} (X_{i,f} p_f - X_{i,s} p_s)$$

D_{ei} is the effective diffusivity of gas component i

\tilde{Q} is permeability, ℓ is the membrane thickness, \tilde{q} is permeability coefficient

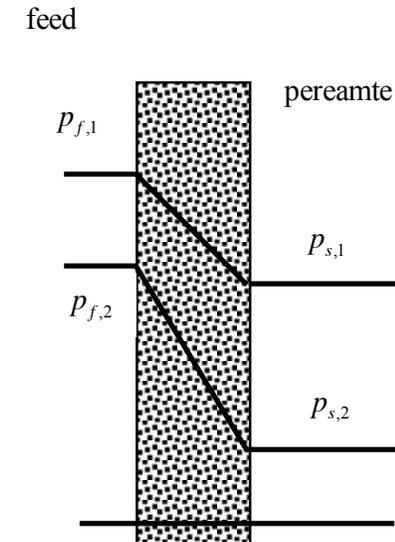
f and s (or p) correspond to feed and sweep (or permeate) sides

Most gases can diffuse across porous and polymer membranes,

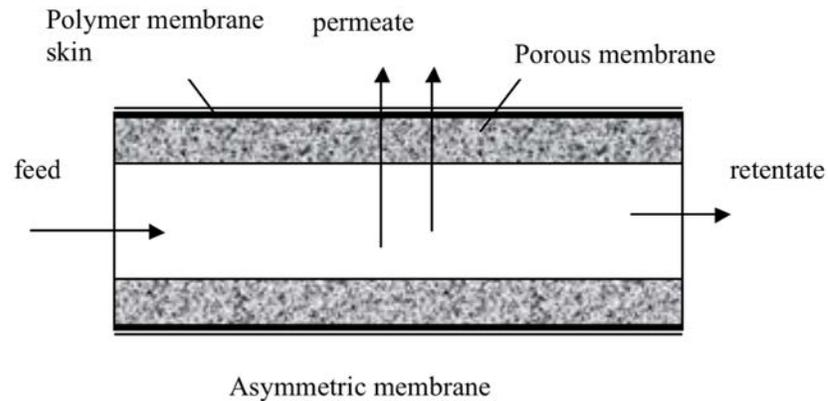
But they diffuse at different rates depending on their value of D_{ei}

The selectivity, $\alpha_{ij} = \frac{\tilde{Q}_i}{\tilde{Q}_j}$, is the ratio of permeabilities of two gases

The selectivity determines the effectiveness of separation



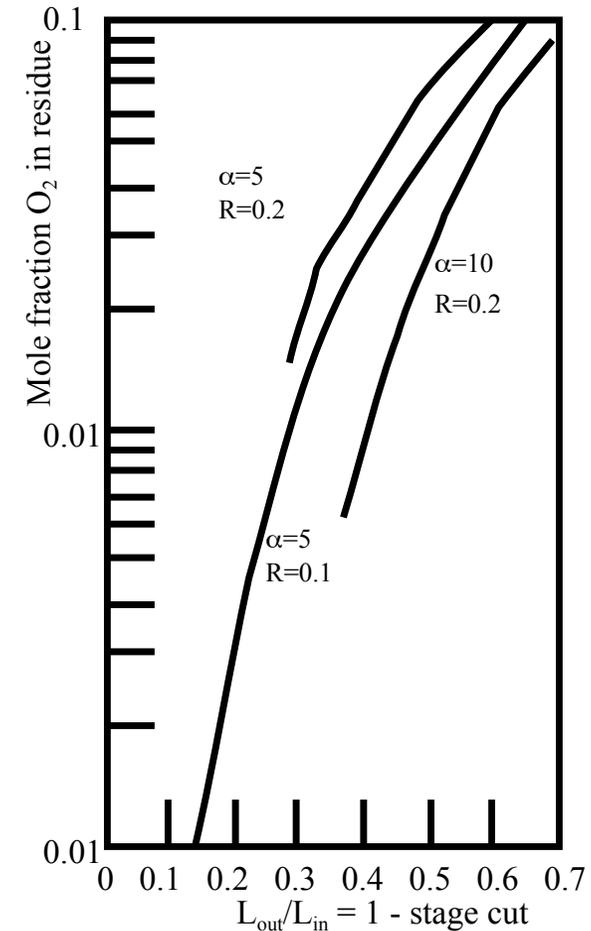
For a porous membrane



Permeability, Q , in $m_N^3 m^{-2} Pa^{-1} s^{-1}$ of the main components of coal gas at 40 °C for the Ube membrane.

Hydrogen	1374
Carbon monoxide	23
Carbon dioxide	461
Nitrogen	13
Argon	31

McCabe, W.L., Smith, J.C., and Harriott, P., Unit Operations of Chemical Engineering, 7th Edition, McGraw Hill, 2005



The production of nitrogen from air in a counter flow membrane separator, showing the residual oxygen in the nitrogen stream as function of the flow rate ratio. α is the membrane selectivity (Q_{O_2}/Q_{N_2}), R is the pressure ratio across the membrane, L is the feed rate.

$$\wp_c = \frac{1}{\eta_c} n \frac{k\mathfrak{R}}{k-1} \left(\left(\frac{p_{out}}{p_{in}} \right)^{\frac{\kappa-1}{n\kappa}} - 1 \right) J_c$$

J_c is the molar flux through the compressor,

n is the number of compression stages,

T_o is the temperature before each compression stage

p_{out} is the pressure after a compression stage,

p_{in} is the pressures before a compression stage,

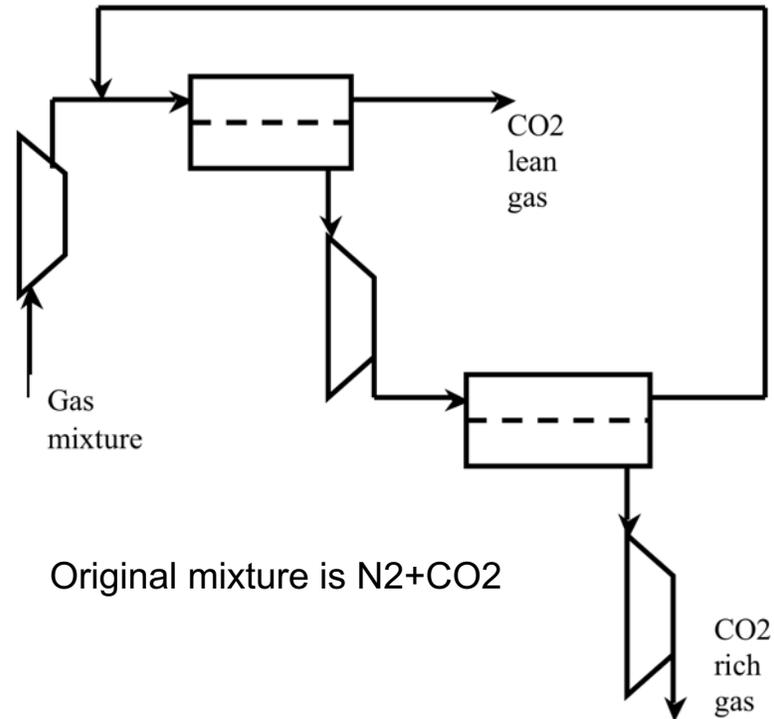
κ is the isentropic index of the gas through the stage

η_c is the isentropic efficiency of the compressor.

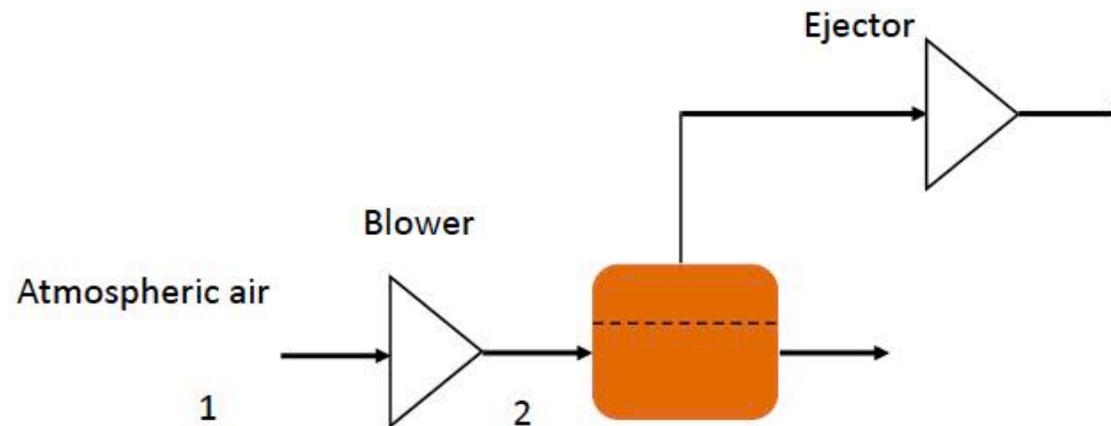
$$\mathfrak{R} = 8.314 \text{ J / mol K}$$

Because of finite permeability of different gases, producing near pure gases requires multi stages, significantly increasing the work.

Energy Requirements:



Example 9.5. An air separation unit (ASU) consisting of a blower, a membrane and an ejector, is shown below. Atmospheric air at 30 °C and 1 atm is sucked through the blower and fed to the membrane at 1.1 atm. The oxygen mole fraction at the permeate side is 35%. The ejector maintains a pressure of 0.3 atm on the permeate side. The membrane is made of a special silicon rubber with 1.5 μm thickness and 10 m^2 area. The permeability coefficient of oxygen is 500 Barrer (at STP). The behavior of gas in all parts of the system can be approximated as an ideal gas. (a) Assuming perfect mixing at the feed and the permeate sides, determine the daily production of the oxygen-enriched stream. (b) Take an isentropic efficiency of 75% for the blower and ejector, and that ; calculate the energy requirement of this unit.



permeability coefficient is $\tilde{p}_{O_2} = 500 \text{ Barrer} = 500 \times 7.500.5 \times 10^{-18} \text{ m}^2 / (\text{s Pa})$

$$J_{O_2} = \frac{\tilde{p}_{O_2}}{t_m} (X_{f,O_2} p_f - X_{p,O_2} p_p) = 3.191 \times 10^{-5} \text{ m}^3 / (\text{m}^2 \cdot \text{s})$$

$$\dot{V}_p = \frac{J_{O_2} A}{X_{p,O_2}} = 0.000912 \text{ m}^3 \text{ s} \quad \text{and}$$

$$\dot{m}_p = \left(X_{p,O_2} M_{O_2} + X_{p,N_2} M_{N_2} \right) \frac{p \dot{V}_p}{\mathfrak{R}T} = 0.001078 \text{ kg / s}$$

total per day is 93.2 kg

$$\dot{W} = \dot{m} c_p \left(\frac{T_{2s} - T_1}{\eta} \right) \quad \text{and} \quad T_{2s} = T_1 \left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}}$$

$$\dot{W}_{blower} = 241.4 \text{ W} \quad \text{and} \quad \dot{W}_{ejector} = 183.7 \text{ W}$$

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