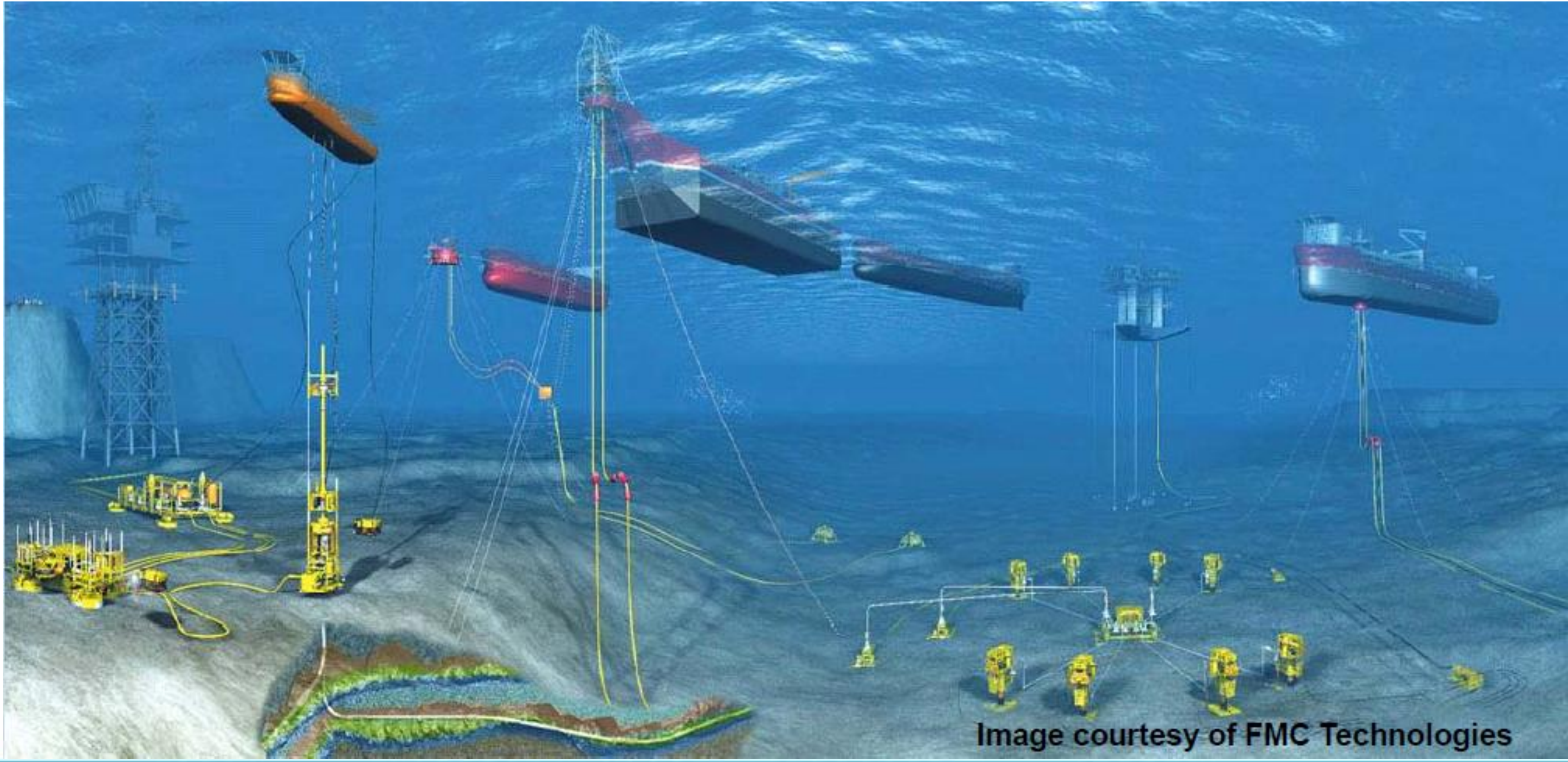


Flow Assurance

서유택

Superficial velocity

- For single phase flow, the mean velocity is defined by the volumetric flowrate of the single phase divided by the cross sectional area.
- If there is more than one phase, there needs to be a way of describing the velocity of each phase. The most convenient way is the phase superficial velocity, which is the velocity the phase would have without the other phase.
- The superficial velocity is defined as the ratio of the liquid or gas volumetric flowrate to the total pipeline cross-sectional area.

$$U_{sl} = \frac{Q_l}{A_f}$$

$$U_{sg} = \frac{Q_g}{A_f}$$

where

U_{sl} = liquid superficial velocity
U_{sg} = gas superficial velocity
Q_l, Q_g = liquid and gas volumetric flowrate, respectively
A_f = pipeline flow cross-sectional area

- Note that superficial velocity is not the same as the velocity at which the phase itself moves, because superficial velocity reflects the relative flow rates.

  : This can be understood by thinking of the case in which most of the cross section is liquid, but a few gas bubbles move with the liquid at the same velocity. The superficial velocity of the gas is then much smaller than the superficial velocity of the liquid even though the velocities of the two phases are the same.

- The ratio between the superficial velocities is not the same as the ratio between the fractions of the cross section occupied by each phase.

  : That can be seen by thinking of a pipeline in which half the cross section is liquid and the other half is gas, but the gas is moving 10 times faster than the liquid.

Basic flow variables

- The fluid <u>Mixture Velocity</u> is defined as the sum of the superficial gas and liquid velocities

$$U_m = U_{sl} + U_{sg} = \frac{Q_l + Q_g}{A}$$

where
U_m = fluid mixture velocity.

- <u>Liquid holdup</u> is defined as the ratio of the liquid volume in a pipeline segment to the whole volume of the pipeline segment. Liquid holdup is a function of both space and time

$$H_l = \frac{V_l}{V}$$

where

H_l = liquid holdup
V_l = pipeline segment volume occupied by liquid
V = whole pipeline segment volume

- Gas void fraction is defined as the ratio of the gas volume in a pipeline segment to the whole volume of the pipeline segment.

$$\alpha_g = \frac{V_g}{V}$$

where

α_g = gas void fraction
V_g = pipeline segment volume occupied by gas

- From the above two equations, the sum of the liquid holdup and gas void fraction equals one.

$$H_l + \alpha_g = 1$$

- Average gas and liquid velocities

  : If the superficial velocity and liquid holdup are known and the liquid holdup would not change longitudinally, the average gas and liquid velocities can be calculated as

$$u_g = \frac{Q_g}{A_g} = \frac{Q_g}{A\alpha_g} = \frac{Q_g}{A(1 - H_l)} = \frac{U_{sg}}{1 - H_l}$$

$$u_l = \frac{Q_l}{A_l} = \frac{Q_l}{AH_l} = \frac{U_{sl}}{H_l} = \frac{U_{sl}}{1 - \alpha_g}$$

where

u_l, u_g = average liquid and gas velocity, respectively
A_l, A_g = pipeline cross-sectional area occupied by liquid and gas, respectively

- Slip velocity

  : Due to the density difference, when gas and liquid flow simultaneously inside a pipeline, the gas phase tends to flow faster than the liquid phase. The gas is “slipping” away from the liquid. The Slip Velocity is defined as the difference of the average gas and liquid velocities

$$u_s = u_g - u_l = \frac{U_{sg}}{1 - H_l} - \frac{U_{sl}}{H_l}$$

- In homogeneous gas and liquid two phase flow, there is not slippage between gas and liquid, and the slip velocity equals zero.
  : Then the liquid holdup can be calculated as

$$H_l = \frac{U_{sl}}{U_{sl} + U_{sg}} = \frac{Q_l}{Q_l + Q_g}$$

- Water cut is defined as the ratio of the water volumetric flowrate to the total water and oil volumetric flowrates,

$$f_w = \frac{Q_w}{Q_w + Q_o} = \frac{Q_w}{Q_l}$$

where

f_w = water cut
Q_o, Q_w = oil and water volumetric flowrate, respectively

- The density of gas and liquid homogeneous mixture is expressed as

$$\rho_m = \rho_l H_l + \rho_g (1 - H_l)$$

where

ρ_m = gas-liquid mixture density
ρ_l, ρ_g = liquid and gas density, respectively

Exercise 2.

- Determine actual and superficial velocities

  : Determine the superficial liquid and vapor velocities (V_{sl} and V_{sg}) and the actual liquid and vapor velocities (V_l and V_g) for the following liquid holdups, flow rates, temperatures and pressures.

  : Use the gas composition and spreadsheet from exercise 1 to calculate the vapor volume per MMscf.

Component	Mole %
N_2	0.50
CO_2	1.50
H_2S	0.00
He	0.00
C_1	88.00
C_2	5.00
C_3	3.00
i-C_4	0.40
n-C_4	1.00
i-C_5	0.20
n-C_5	0.30
C_6	0.10

Flowline ID (in)	Temp. (F)	Press. (psia)	In-situ vap or (MMscfd)	STO (kbpd)	Oil Swelling (%)	Liquid Holdup (%)	V_{sl} (ft/s)	V_{sg} (ft/s)	V_l (ft/s)	V_g (ft/s)
8	120	2000	3	30	20	95				
8	80	1200	9	30	12	70				
8	40	300	18	30	8	40				
10	120	2000	5	50	20	96				
10	80	1200	15	50	12	72				
10	40	300	30	50	8	44				

Beggs and Brill model for two phase flow

- No slip liquid holdup

  : λ_L is defined as the ratio of the volume of the liquid in a pipe segment divided by the volume of the pipe segment which would exist if the gas and liquid travelled at the same velocity (no-slippage). It can be calculated directly from the known gas and liquid volumetric flowrates.

$$\lambda_L = \frac{q_L}{q_L + q_g}, \quad \lambda_g = 1 - \lambda_L$$

- For no slip condition,

$$For\ No - Slip : u_g = u_L\ or\ \frac{U_{sg}}{1 - \lambda_L} = \frac{U_{sl}}{\lambda_L} \Rightarrow \lambda_L = \frac{U_{sl}}{U_m}$$

- Froude number of the mixture

$$Fr_m = \frac{u_m^2}{gD}$$

where, D is pipe ID and g is gravitational constant

- Transition lines for correlation

$$L_1 = 316\ \lambda_l^{0.302},\ L_2 = 0.0009252\ \lambda_l^{-2.4684},\ L_3 = 0.10\ \lambda_l^{-1.4516},\ L_4 = 0.5\ \lambda_l^{-6.738}$$

- Determining flow regimes

Segregated if $\lambda_l < .01$ and $Fr_m < L_1$ or $\lambda_l \geq .01$ and $Fr_m < L_2$

Transition if $\lambda_l \geq .01$ and $L_2 < Fr_m \leq L_3$

Intermittent if $.01 \leq \lambda_l < 0.4$ and $L_3 < Fr_m \leq L_1$ or $\lambda_l \geq .4$ and $L_3 < Fr_m \leq L_4$

Distributed if $\lambda_l < .4$ and $Fr_m \geq L_1$ or $\lambda_l \geq .4$ and $Fr_m > L_4$

- For segregated, intermittent and distributed flow regimes used the following

$$H_l = H_{l0}\varphi, \qquad H_{l0} = \frac{a\lambda_l^b}{Fr_m^c} \text{ (horizontal liquid holdup)}$$

- Actual liquid holdup is obtained by multiplying H_{l0} by a correction factor φ

$$\varphi = 1 + C\left[sin\left(1.8\theta\right) - 0.333sin^3\left(1.8\theta\right)\right]$$

$$C = (1 - \lambda_l)ln(d\,\lambda_l^e N_{vl}^f Fr_m^g)$$

- Liquid velocity number

$$N_{vl} = 1.938\,u_{sl}\left(\frac{\rho_l}{g\,\sigma}\right)^{0.25}$$

u_{sl}: no slip velocity

Beggs and Brill holdup constants

Flow regime	a	b	c	
Segregated	0.98	0.4846	0.0868	
Intermittent	0.845	0.5351	0.0173	
Distributed	1.065	0.5824	0.0609	
	d	e	f	g
Segregated uphill	0.011	−3.768	3.539	−1.614
Intermittent uphill	2.96	0.305	−0.4473	0.0978
Distributed uphill	No correction, $C = 0$, $\psi = 1$			
All regimes downhill	4.70	−0.3692	0.1244	−0.5056

- For transitional flow, the liquid holdup is calculated using both the segregated & intermittent equations and interpolating using the following

$$H_l = AH_l(Segregated) + B\,H_l(Intermittent)$$

$$A = \frac{L_3 - Fr_m}{L_3 - L_2}, \; B = 1 - A$$

- General pressure gradient equation

  : The pressure gradient equation which is applicable to any fluid flowing in a pipe inclined at an angle φ from horizontal was derived previously. This equation is usually adapted for two-phase flow by assuming that the two-phase flow regime and two-phase properties can be considered homogeneous over a finite volume of the pipe.

In-situ average density

$$\rho_L H_L + \rho_g H_g$$

Depend on the using correlation

$$U_{sL} + U_{sg}$$

$$\left(-\frac{dP}{dZ}\right) = \underbrace{\frac{g}{g_c}\rho_s \sin\varphi}_{\Delta P_{elevation}} + \underbrace{\frac{f_{tp}\rho_m u_m^2}{2g_c d}}_{\Delta P_{friction}} + \frac{\rho_s}{2g}\frac{du_m^2}{dL}$$

$\Delta P_{elevation}$
Pressure change due to hydrostatic head

$\Delta P_{friction}$
Frictional pressure gradient

- The no slip friction factor f_n is based on smooth pipe ($\varepsilon/D=0$) and the Reynolds number, Re.

$$f_{tp} = f_n e^S$$

$$S = \frac{ln(x)}{(-0.0523 + 3.182ln(x) - 0.8725[ln(x)]^2 + 0.01853[ln(x)]^4}$$

$$x = \frac{\lambda_l}{H_l^2}$$

- Liquid Viscosity (μ_L): μ_L may be calculated from the oil and water viscosities with assumption of no slippage between the oil and water phases as follows:

$$\mu_L = \mu_o f_o + \mu_w f_w$$

- Two-Phase Viscosity: Calculation of the two-phase viscosity requires knowledge of the liquid holdup. Two equations for two-phase viscosity are used by various investigators in two-phase flow:

$$\mu_m = \mu_L \lambda_L + \mu_g \lambda_g \;\; , \mu_s = \mu_L^{H_L} \times \mu_g^{H_g}$$

- Liquid Surface Tension (σ_L):

$$\sigma_L = \sigma_o f_o + \sigma_w f_w$$

Two phase flow correlations

- Many correlations have been developed for predicting two-phase flow pressure gradients which differ in the manner used to calculate the three terms of pressure gradients equation (elevation change, friction and acceleration terms):

a. *No slip, no flow regime considerations:* the mixture density is calculated based on the no slip holdup. No distinction is made for different flow regimes.

b. *Slip considered, no flow regime consideration:* The same correlations for liquid holdup and friction factor are used for all flow regimes.

c. *Slip considered, flow regime considered:* Usually a different liquid holdup and friction factor prediction methods are required in each flow regimes.

Exercise 3. Determine pressure drop

- In-situ Vapor MW = 20.6
- Q_l =200 bpd, ρ_l = 49.9 lb/ft^3, μ_l = 2 cp
- Q_g=1mmscfd, ρ_g=2.6 lb/ft^3, μ_g=0.0131 cp

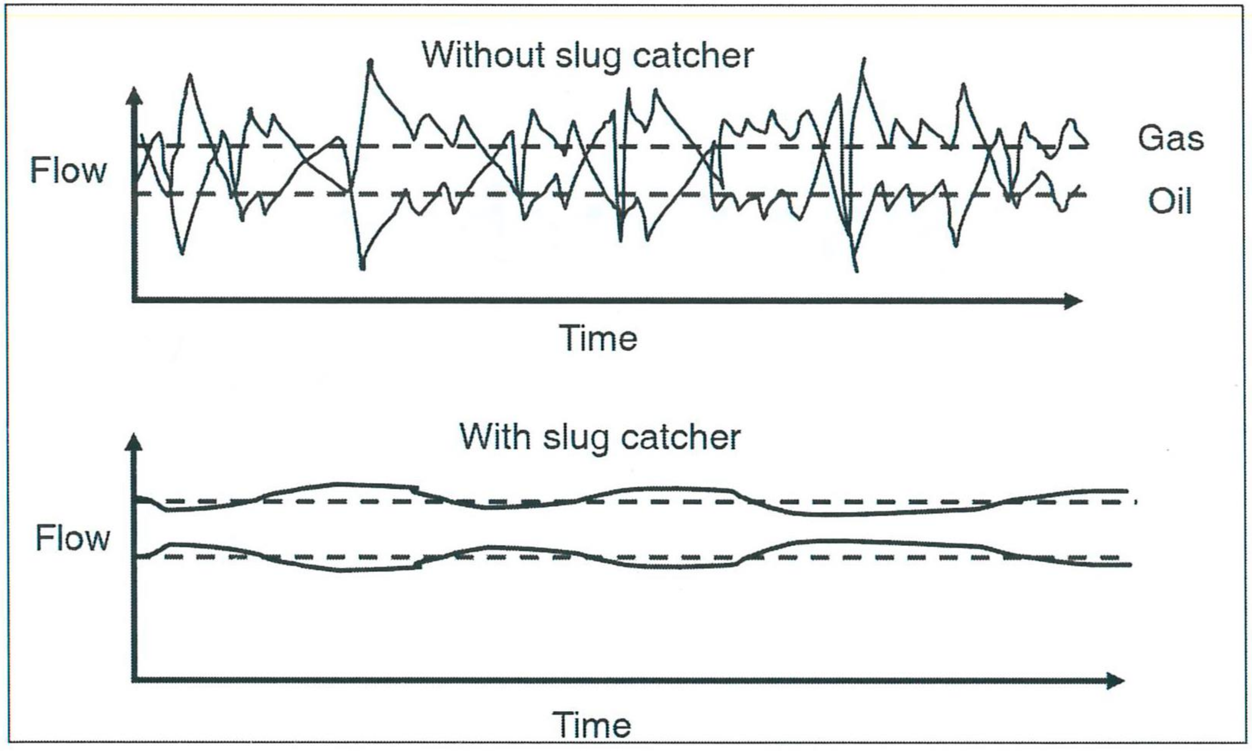
1. Find the flow regime, from Froude number and no-slip liquid holdup

2. Find the pressure drop for elevation change per length (assume θ=90)

3. Find the frictional pressure drop per length

Slug catcher

- What are they?
  - Large capacity/volume separators designed to receive unsteady multiphase flow from multiphase pipelines
  - Two main types
    : Finger or pipe type
    : Vessel type

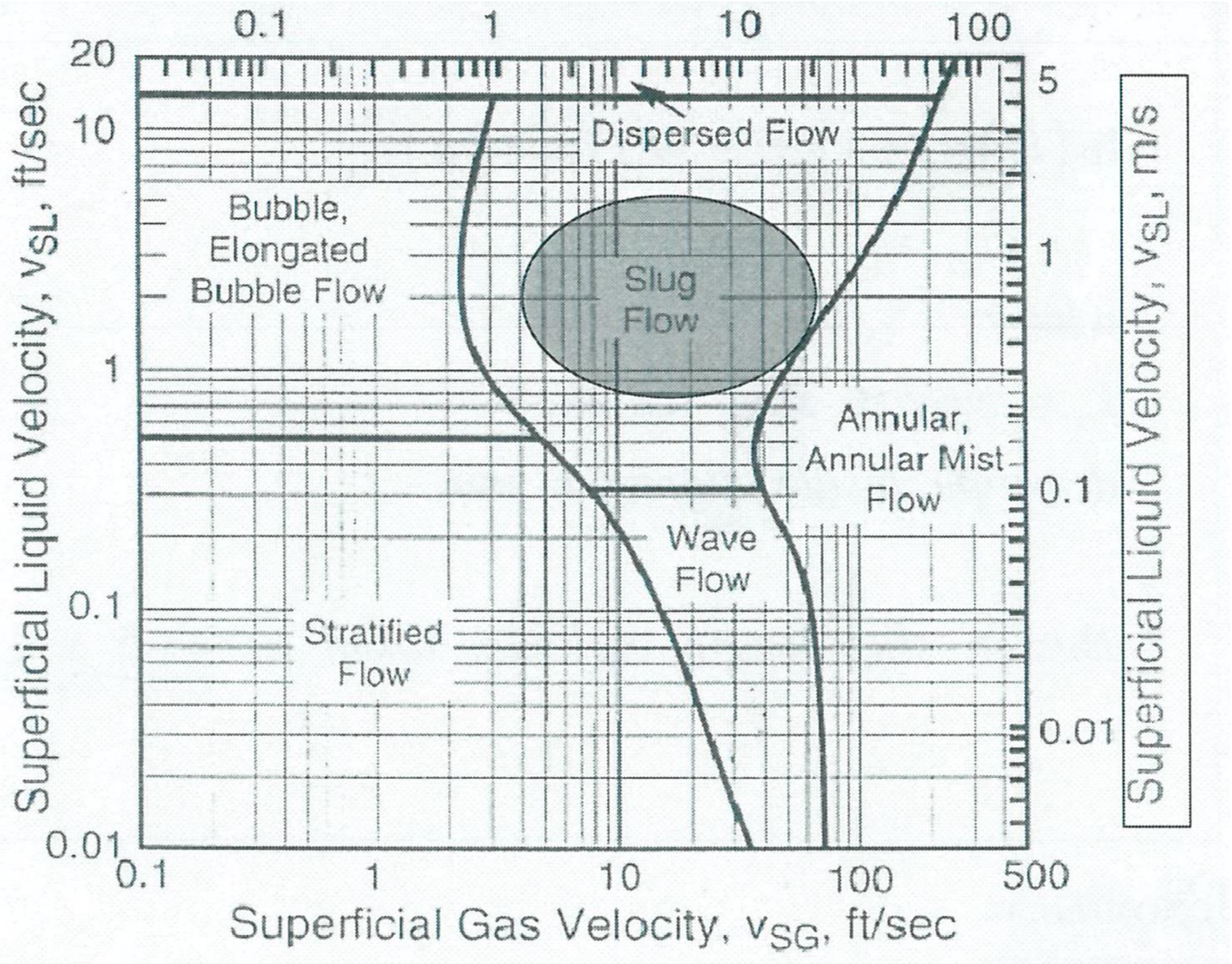


Causes of slug/intermittent flow

- Operation in the slug flow regimes
  : Hydrodynamic slugging
  : Gas velocity > Fluid velocity
- Terrain induced slugging
  : $\pm$ 10% changes
  : Start up operation
- Flow rate changes
  : Ramping up
- Pigging
  : Sweeping liquids
- Flowline/riser geometry: Severe slugging

Flow pattern map of Mandhan

- Horizontal two phase flow in pipes



Hydrodynamic slug size prediction (FPS vs SI)

- $$\ln(L_s) = -25.4144 + 28.4948\left(\ln(d)\right)^{0.1}$$

  where :

  L_s = *average slug length , ft*

  d = *pipe inside diameter , in*

- $$\ln(L_s) = -65.807 + 59.115\left(\ln(d)\right)^{0.1}$$

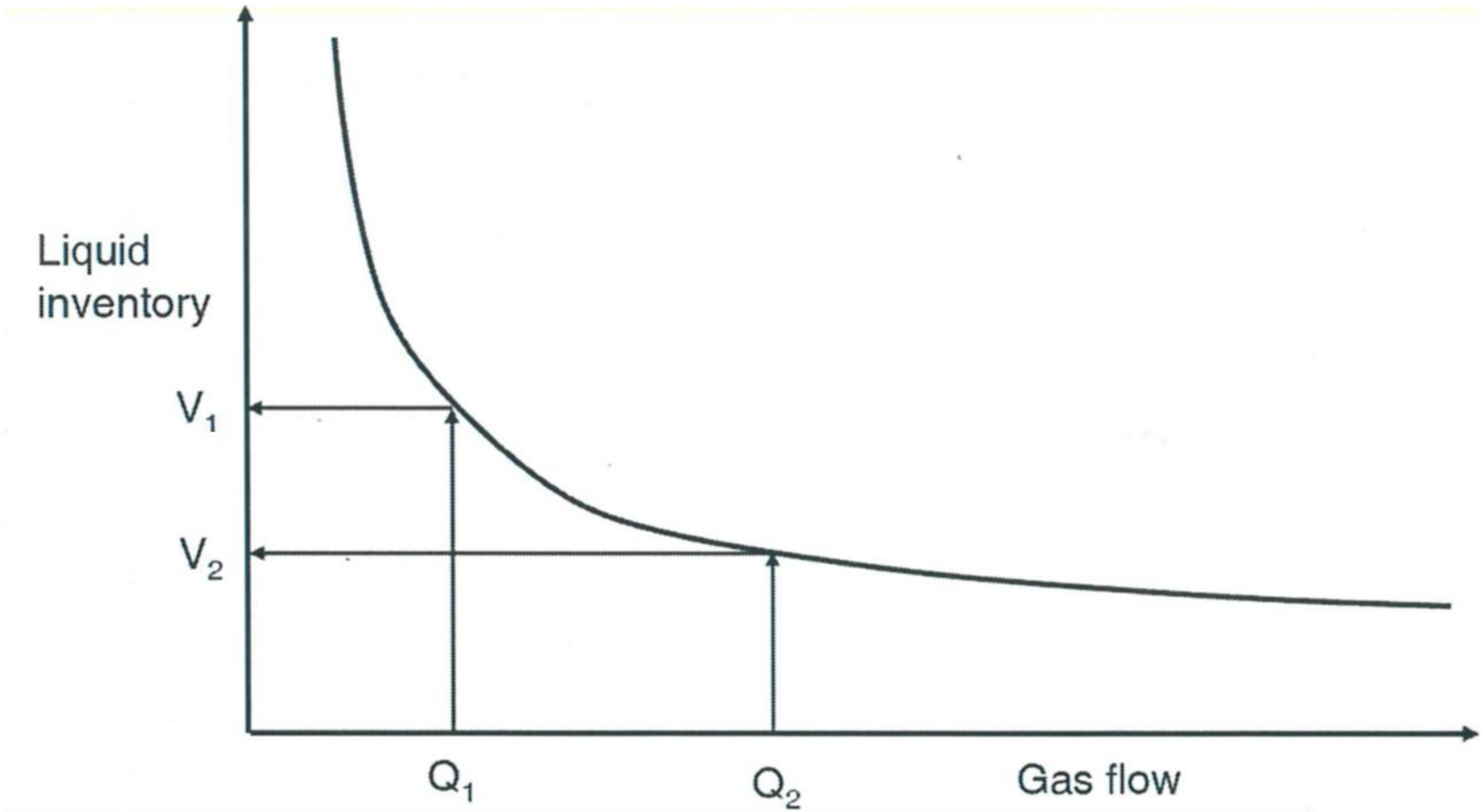
  where :

  L_s = *average slug length, m*

  d = *pipe inside diameter, mm*

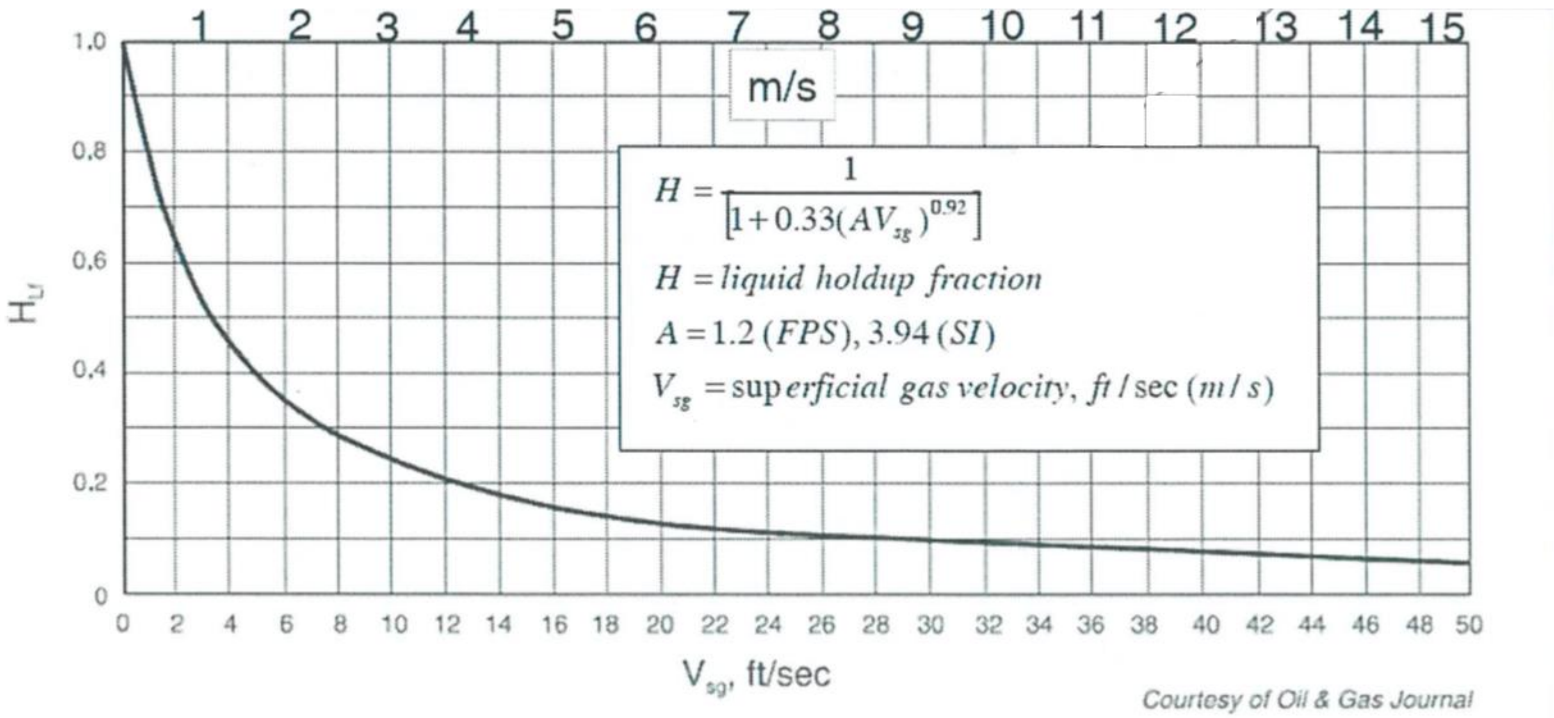
- Design slug length typically taken as 4 ~ 5 times L_s

Pipeline liquid holdup



Simple holdup correlation - Flanigan

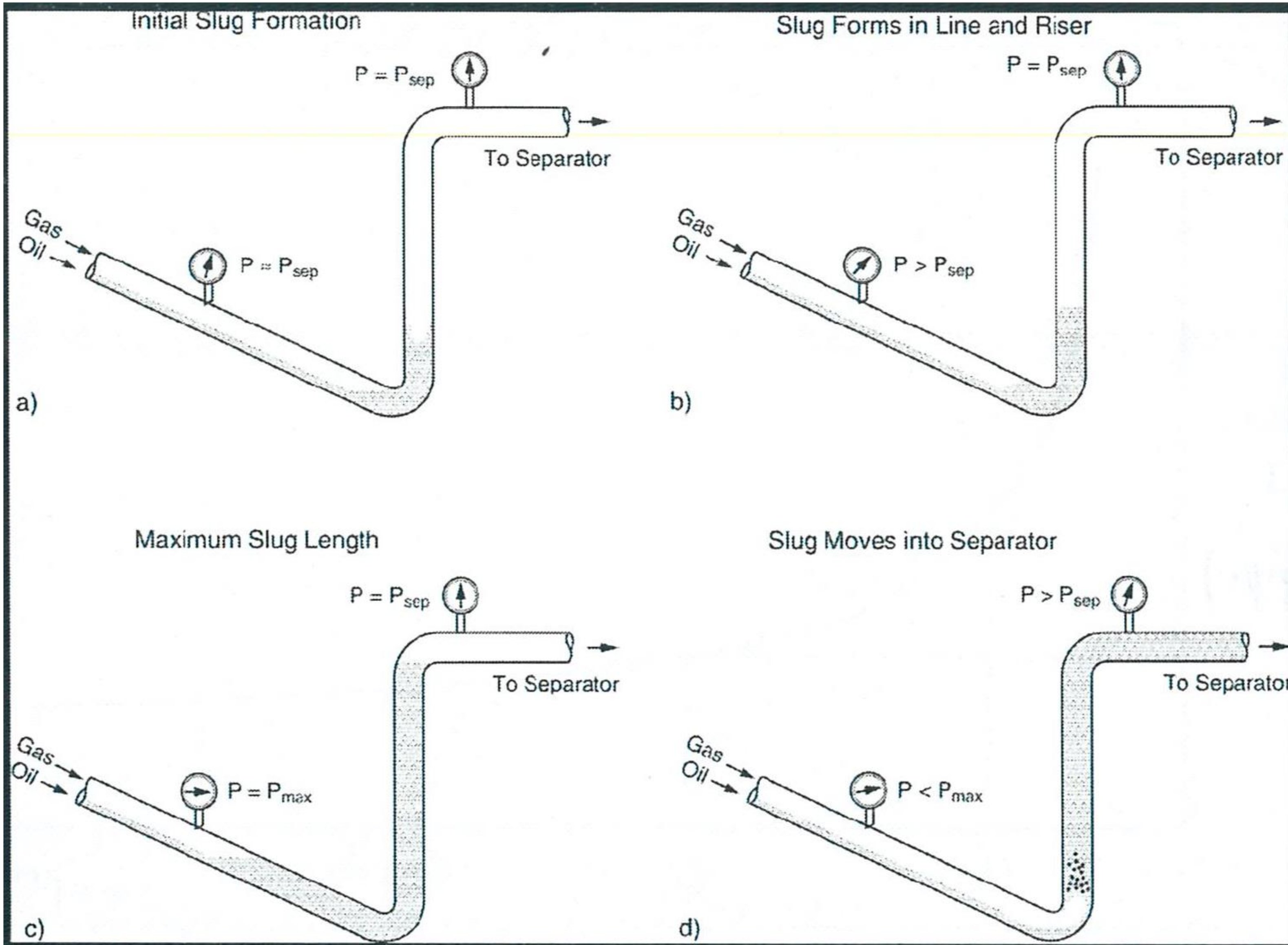
- Slug size is based on “Hold Up” difference between flow rate 1 and 2



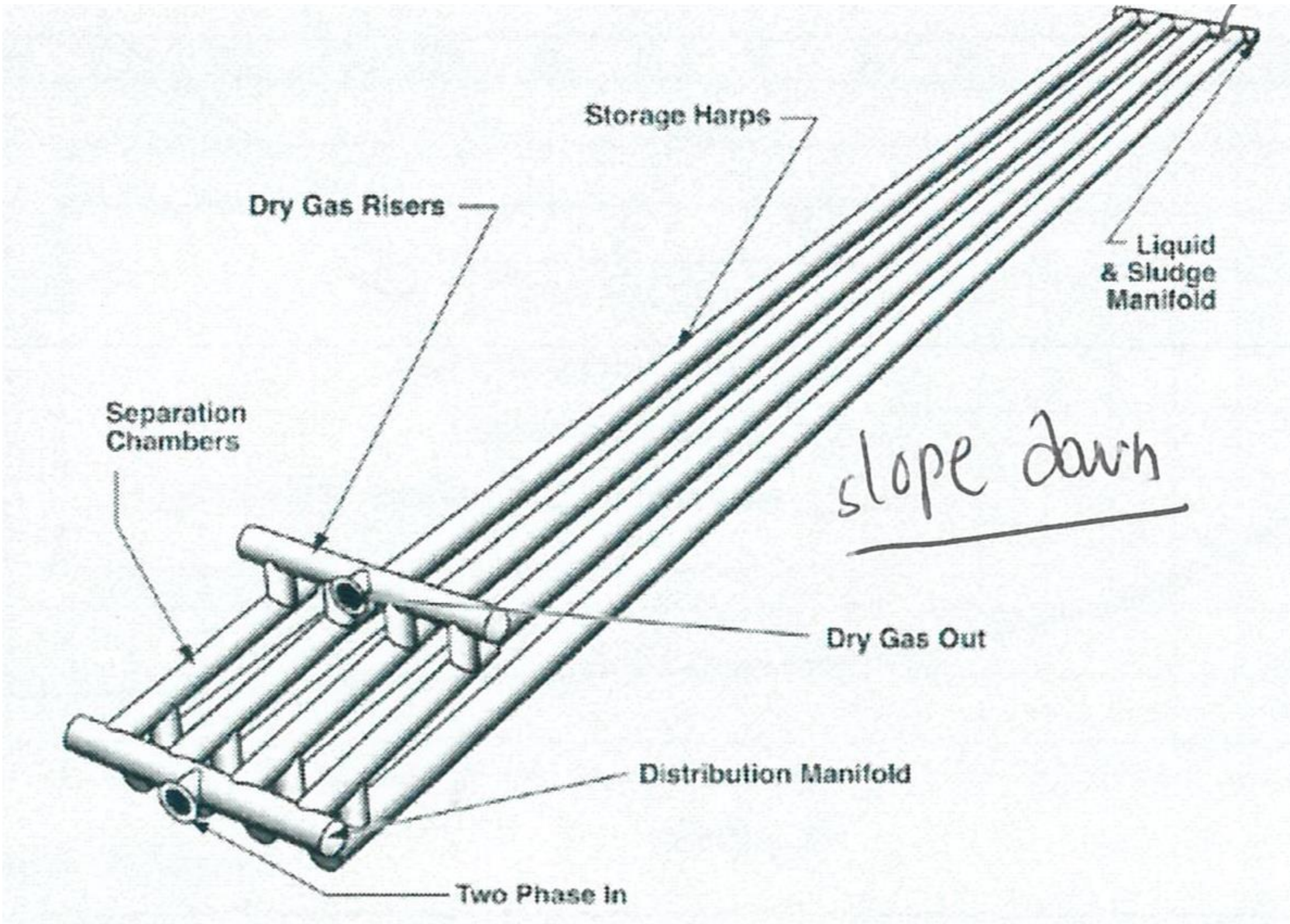
where, $V_{sg} = A\frac{q_g\, z\, T}{d^2\, P}$

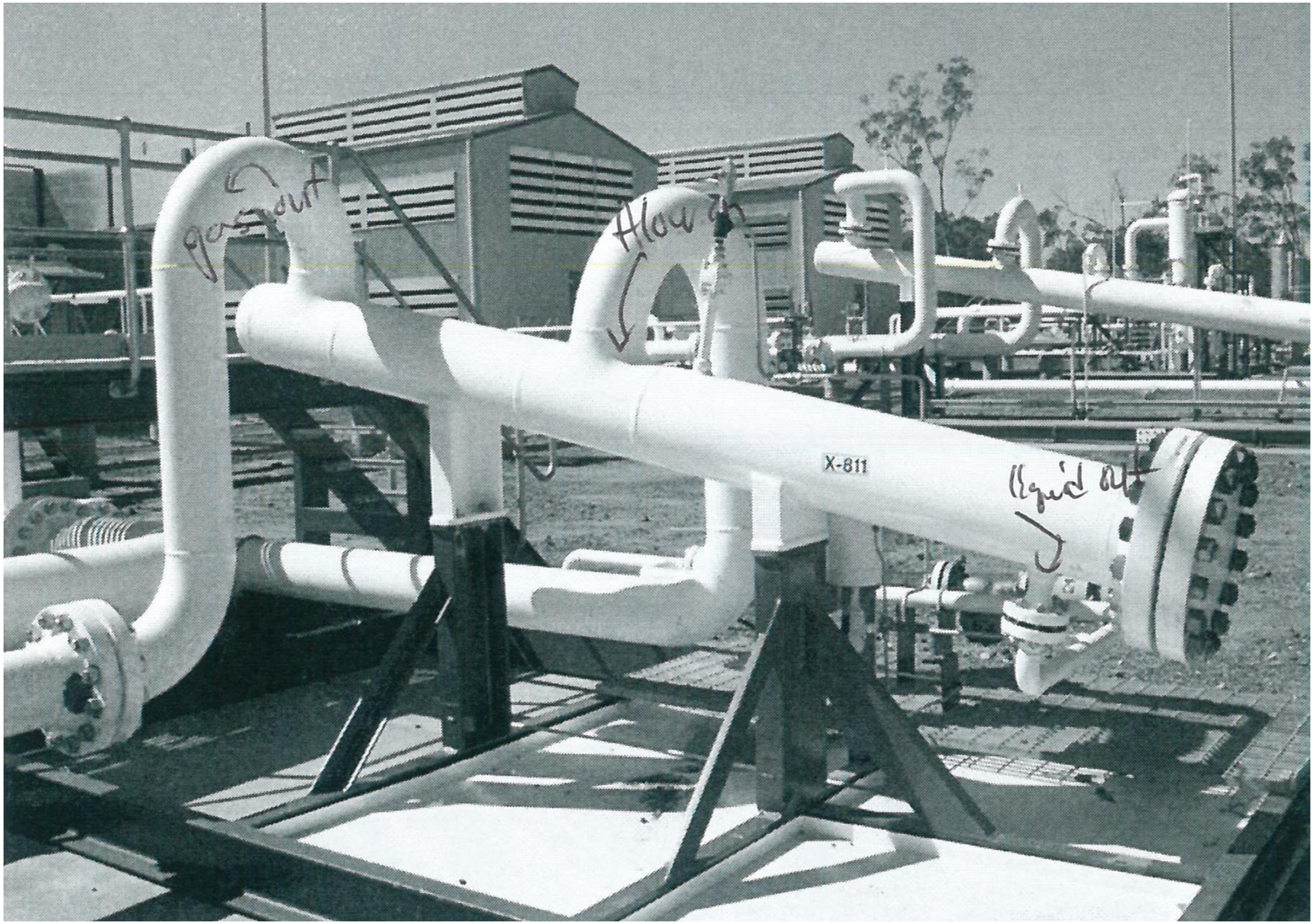
q_g in MMm^3/d, T in K, d in m, P in kPa, A in 5.19 (SI)

Severe slugging



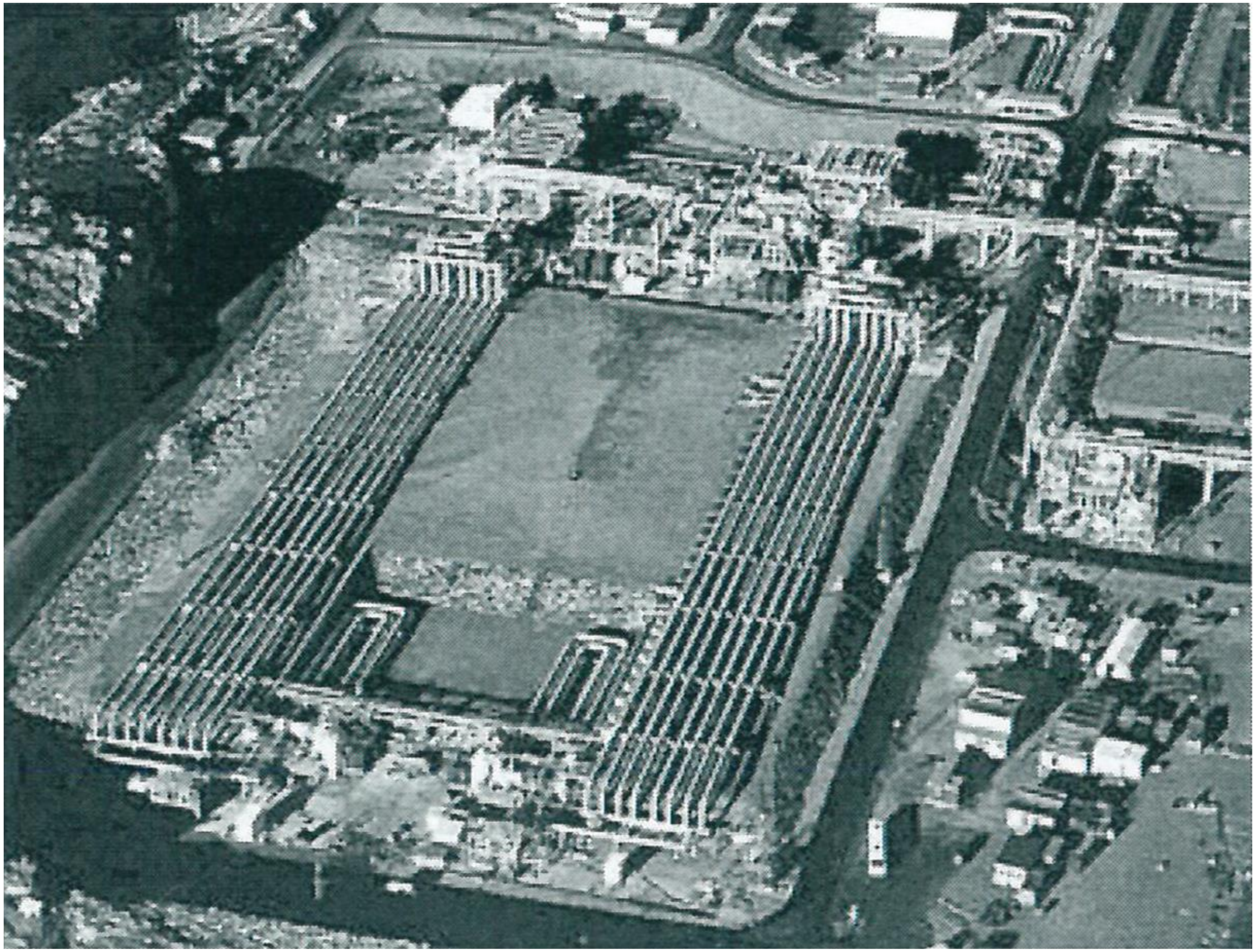
Finger type slug catcher

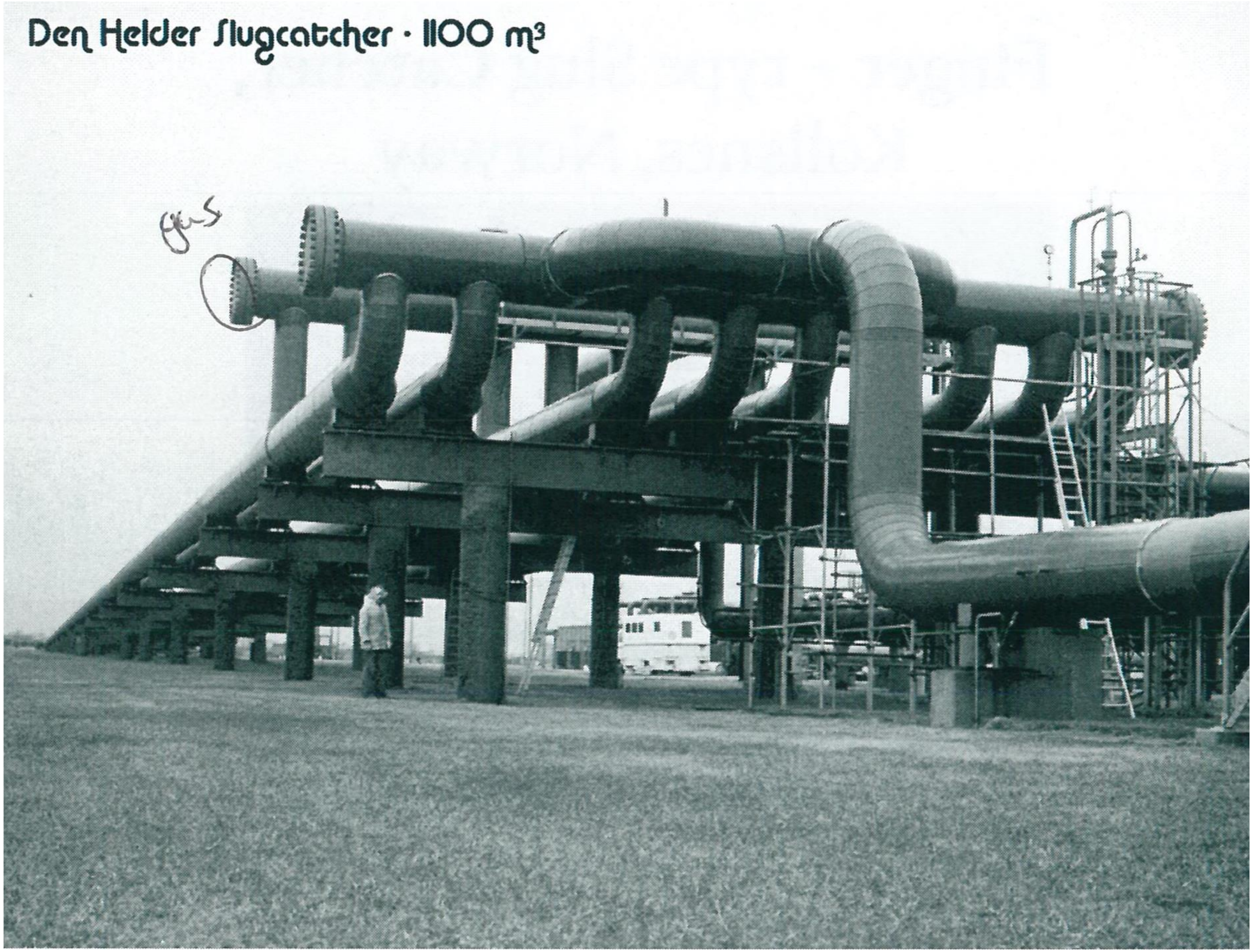



gas out
Flow in
X-811
liquid out

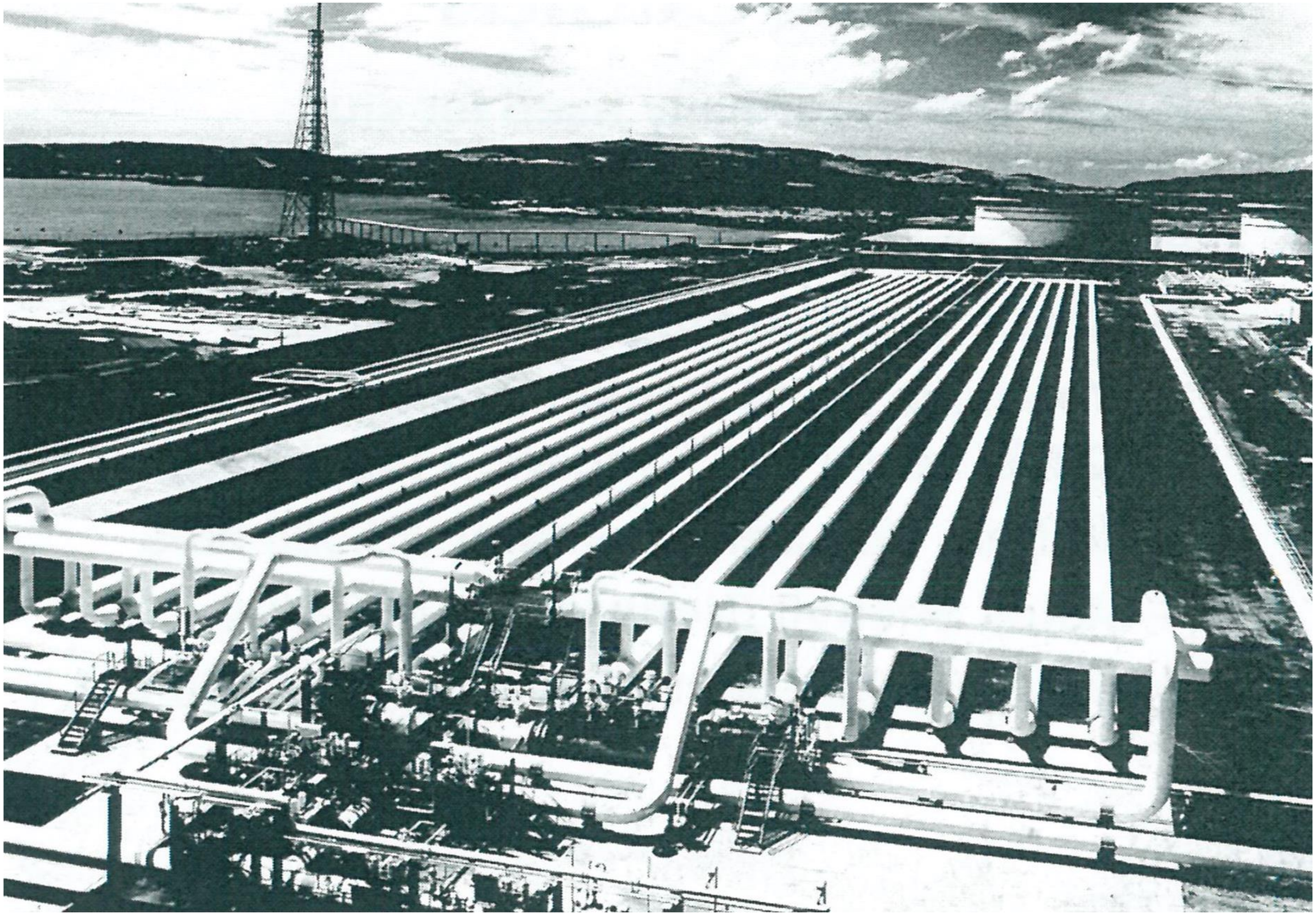
Finger type slug catcher

- Kollsnes, Norway



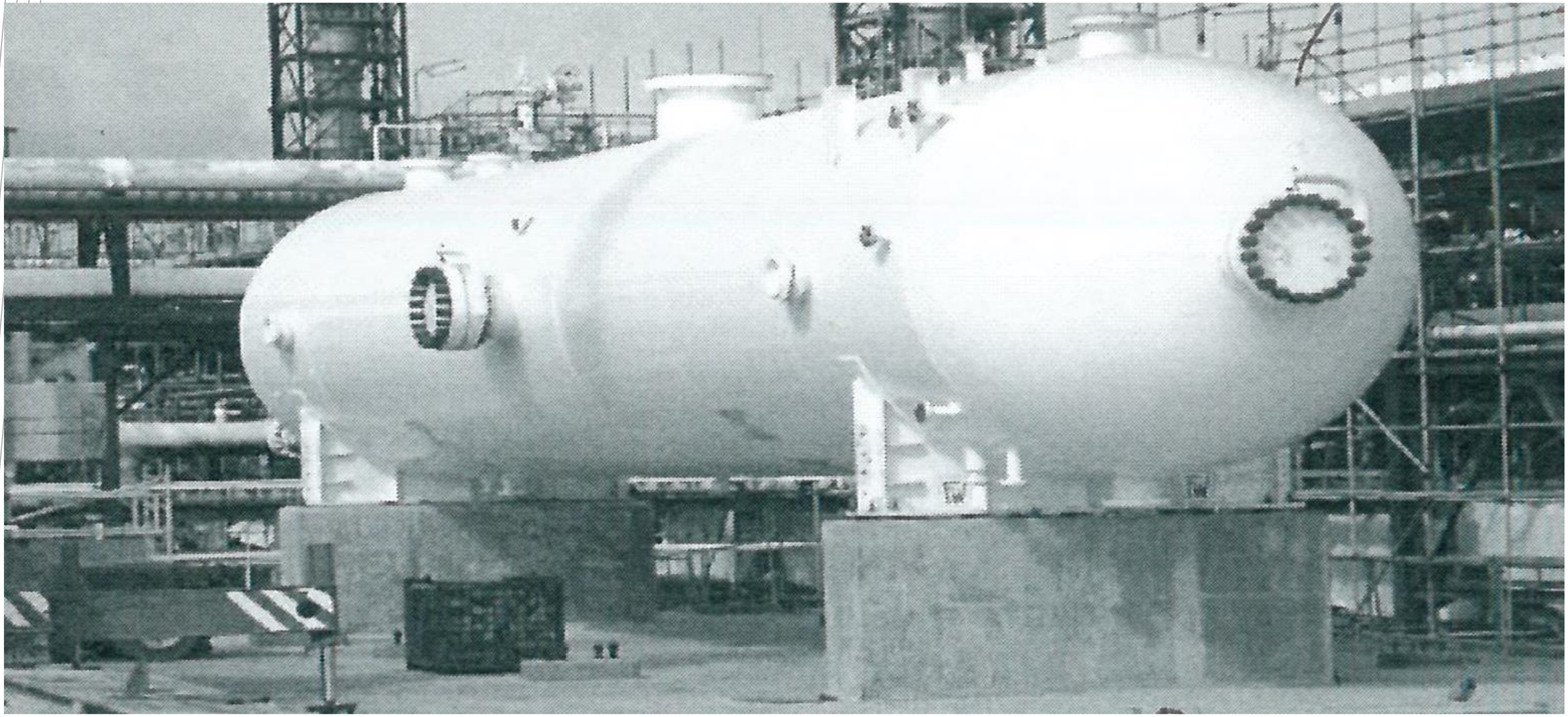

Den Helder Slugcatcher · 1100 m³
gas

- Woodside slug catcher – 32,500 bbls (5000 m3)



Shell Goldeneye Vessel type slug catcher

- 66’ L * 13.1’ D * 3.6’ wt, 240 ton, 1200 bbls slug capacity

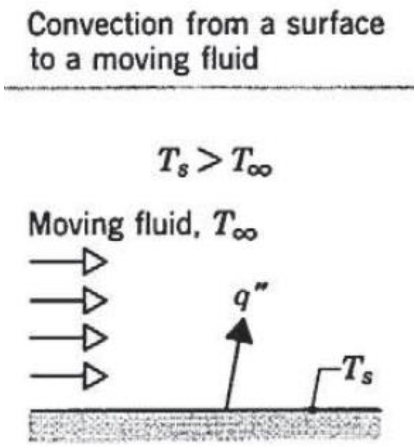


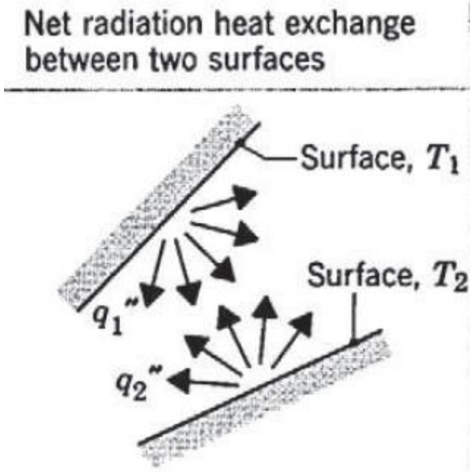
Slug catchers

- Finger type normally used for slug volumes larger than 1500 bbls
- Fingers are normally 36 ~ 48 inch, 300 ~ 800 ft long
- Typical capacities are 3000 ~ 8000 bbls
  - Woodside North Rankin (NWS, Australia)
    : 32,500 bbls, 14~48” fingers, 1150 ft long
  - BP Nam Con Son (Vietnam)
    : 25,000 bbls
  - Statoil Snohvit (Nroway)
    : 17,000 bbls
- Typical installed cost (Finger type): $1500/bbl

Heat transfer on a surface

- If a surface and a moving fluid have a temperature difference, the convection will occur between the fluid and surface.
- All solid surfaces with a temperature will emit energy in the form of electromagnetic waves, which is called radiation





- Although these three heat transfer modes occur at all subsea systems, for typical pipelines, heat transfer from radiation is relatively insignificant compared with heat transfer from conduction and convection.

Conduction

- For a one-dimensional plane with a temperature distribution T(x), the heat conduction is quantified by the following Fourier equation:

$$q'' = -k \cdot \frac{dT(x)}{dx}$$

Where,

q'': heat flux, Btu/(hr ft^2) or W/m^2, heat transfer rate in the x direction per unit area;

k: thermal conductivity of material, Btu/(ft hr ^{o}F) or W/(m K);

dT/dx: temperature gradient in the x direction, ^{o}F/ft or ^{o}C/m.

- When the thermal conductivity of a material is constant along the wall thickness, the temperature distribution is linear and the heat flux becomes:

$$q'' = -k \cdot \frac{T_2 - T_1}{x_2 - x_1}$$

- For a steady heat transfer, the right side of equation is equal to zero.
- The total heat flow per unit length of cylinder is calculated by following equation:

$$q_r = -2\pi k \frac{T_2 - T_1}{\ln(r_2/r_1)}$$

Where,

r_1, r_2: inner and outer radii of the cylinder medium, ft or m;

T_1, T_2: temperatures at corresponding points of r_1, r_2, ^oF or ^oC;

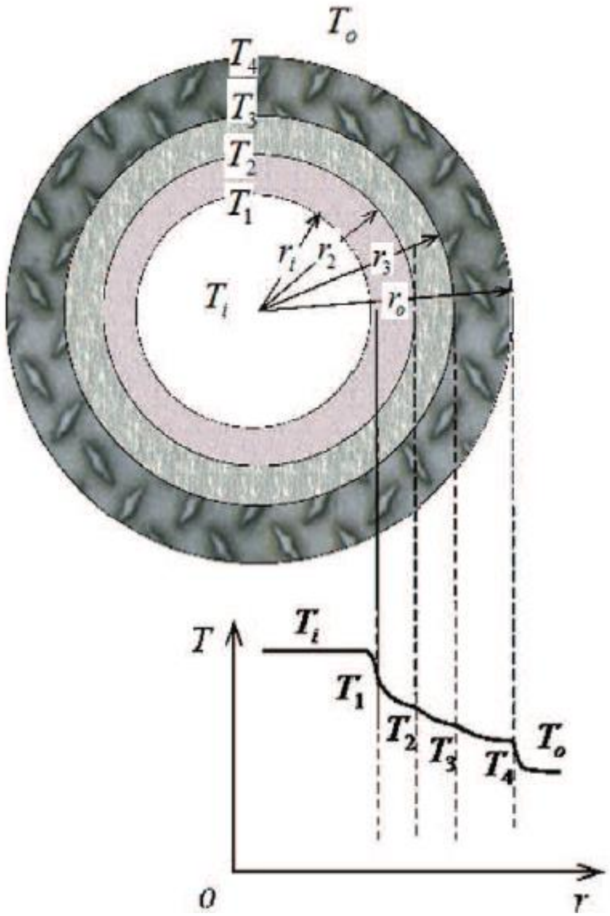
q_r: heat flow rate per unit length of cylinder, Btu/(hr ft) or W/m.

Convection

- Both internal and external surfaces of a subsea pipeline come in contact with fluids, so convection heat transfer will occur when there is a temperature difference between the pipe surface and the fluid.
- The convection coefficient is also called a film heat transfer coefficient in the flow assurance field because convection occurs at a film layer of fluid adjacent to the pipe surface.

U-value

- Figure shows the temperature distribution of a cross section for a composite subsea pipeline with two insulation layers.



- Radiation between the internal fluid and the pipe wall and the pipeline outer surface and the environment is ignored because of the relatively low temperature of subsea systems.
- Convection and conduction occur in an insulated pipeline as follows,
  : Convection from the internal fluid to the pipeline wall;
  : Conduction through the pipe wall and exterior coatings, and/or to the surrounding soil for buried pipelines;
  : Convection from flowline outer surface to the external fluid

- For internal convection at the pipeline inner surface, the heat transfer rate across the surface boundary is given by the Newton equation:

$$Q_i = A_i \cdot h_i \cdot \Delta T = 2\pi\, r_i\, L\, h_i\, (T_i - T_1)$$

where

Q_i : convection heat transfer rate at internal surface, Btu/hr or W;

h_i : internal convection coefficient, Btu/(ft^2 hr ^{o}F) or W/(m^2 K);

r_i : internal radius of flowline, ft or m;

L: flowline length, ft or m;

A_i : internal area normal to the heat transfer direction, ft^2 or m^2;

T_i : internal fluid temperature, ^{o}F or ^{o}C;

T_1: temperature of flowline internal surface, ^{o}F or ^{o}C.

- Conduction in the radial direction of a cylinder can be described by Fourier's equation in radial coordinates:

$$Q_r = -2\pi\, r\, L\, k \frac{\partial T}{\partial r}$$

where

Q_r : conduction heat transfer rate in radial direction, Btu/hr or W;

r: radius of cylinder, ft or m;

k: thermal conductivity of cylinder, Btu/(ft hr ^{o}F) or W/(m K);

$\partial T / \partial r$: temperature gradient, ^{o}F/ft or ^{o}C/m.

- Integration gives:

$$Q_r = \frac{2\pi\, L\, k (T_1 - T_2)}{\ln(r_2/r_1)}$$

- The temperature distribution in the radial direction can be calculated for steady-state heat transfer between the internal fluid and pipe surroundings, where heat transfer rates of internal convection, external convection, and conduction are the same.
- The following heat transfer rate equation is obtained:

$$Q_r = \frac{T_i - T_o}{\frac{1}{2\pi r_i L h_i} + \frac{\ln(r_1/r_i)}{2\pi k_1 L} + \frac{\ln(r_2/r_1)}{2\pi k_2 L} + \frac{\ln(r_o/r_2)}{2\pi k_3 L} + \frac{1}{2\pi r_o L h_o}}$$

- The heat transfer rate through a pipe section with length of L, due to a steady-state heat transfer between the internal fluid and the pipe surroundings, is also expressed as follows:

$$Q_r = UA(T_i - T_o)$$

where

U: overall heat transfer coefficient based on the surface area A, Btu/ (ft^2 hr ^oF) or W/ (m^2 K);

A: area of heat transfer surface, A_i or A_o, ft^2 or m^2;

T_o : ambient temperature of the pipe surroundings, ^oF or ^oC;

T_i : average temperature of the flowing fluid in the pipe section, ^oF or ^oC.

Calculating U-value for single layer

- For single layer,

$$U = \frac{2k}{D_i \times \ln\left(\frac{D_o}{D_i}\right)}$$

where, D_i =inside diameter and D_o = outside diameter

Multilayer insulation

- The U-value for a multilayer insulation coating system is easily obtained from an electrical-resistance analogy between heat transfer and direct current.
- The steady-state heat transfer rate is determined by:

$$Q_r = UA(T_i - T_o) = (T_i - T_o)/\sum R_i$$

where UA is correspondent with the reverse of the cross section's thermal resistivity that comprises three primary resistances: internal film, external film, and radial material conductance.

- The relationship is written as follows:

$$\frac{1}{UA} = \sum R_i = R_{film,in} + R_{pipe} + \sum R_{coating} + R_{film,ext}$$

- The terms on the right hand side of the above equation represent the heat transfer resistance due to internal convection, conduction through steel well of pipe, conduction through insulation layers and convection at the external surface.
- They can be expressed as follows.

$$R_{film,in} = \frac{1}{h_i A_i}$$

$$R_{pipe} = \frac{\ln(r_l/r_i)}{2\pi L k_{pipe}}$$

$$\sum R_{coating} = \frac{\ln(r_{no}/r_{ni})}{2\pi L k_n}$$

$$R_{film,ext} = \frac{1}{h_o A_o}$$

- Therefore, the U-value based on the flowline internal surface area A_i is:

$$U_i = \frac{1}{\frac{1}{h_i} + \frac{r_i ln(r_1/r_i)}{k_1} + \frac{r_i ln(r_2/r_1)}{k_2} + \frac{r_i ln(r_o/r_2)}{k_3} + \frac{r_i}{r_o h_o}}$$

and the U-value based on the flowline outer surface area A_o is:

$$U_o = \frac{1}{\frac{r_o}{r_i h_i} + \frac{r_o ln(r_1/r_i)}{k_1} + \frac{r_o ln(r_2/r_1)}{k_2} + \frac{r_o ln(r_o/r_2)}{k_3} + \frac{1}{h_o}}$$

- U-value is a function of many factors, including the fluid properties and fluid flow rates, the convection nature of the surroundings, and the thickness and properties of the pipe coatings and insulation.
- Insulation manufacturers typically use a U-value based on the outside diameter of a pipeline, whereas pipeline designers use a U-value based on the inside diameter.
- The relationship between these two U-values is:

$$U_o \times \mathrm{OD} = U_i \times \mathrm{ID}$$

Exercise 4. U-value calculation

ID (in)	WT (in)	K-value (W/m/K)	U-value (W/m^2/K)
6	1	0.20	
6	2	0.20	
6	4	0.20	
6	1	0.01	
6	2	0.01	
10	1	0.01	
10	1	0.20	
10	2	0.20	
10	4	0.20	

Note: U-value is based on the tube ID.

$$U = \frac{2k}{D_i \times \ln\left(\frac{D_o}{D_i}\right)}$$

Insulation design for flowlines

Flowline	Layers	WT mm	OD mm	ID mm	K W/m°K	Cp j/kg°K	Density kg/m³	U-Valve W/m^2-°K
6″ ID PIP Production Well Jumper WD = 1880m	Fluid (Methane)	–	152.4	–	–	3550	170	1.14 (0.200 Btu/hr-ft^2-deg F)
	Steel	12.7	177.8	152.4	45	461	7865	
	FBE	0.30	178.4	177.8	0.30	1500	1300	
	PUF	29.8	238.0	178.4	0.025	900	64	
	Steel	15.9	269.8	238.0	45.000	461	7865	
8″ ID Flexible Flowline WD = 1880m	Fluid (Methane)	–	203.2	–	–	3550	170	3.82 (0.673 Btu/hr-ft^2-deg F)
	316 SS	10.0	223.2	203.2	14.00	470	4491	
	PA 11	12.00	247.2	223.2	0.27	2000	1050	
	Polyester	2.20	251.6	247.2	0.13	450	800	
	Carb. Steel	18.00	287.6	251.6	56.00	450	7860	
	PP Foam	50.0	387.6	287.6	0.16	2180	700	
	PA 11	10.0	407.6	387.6	0.27	2000	1050	
8″ ID Wet Insulation Production Flowline WD = 1760m	Fluid (Methane)	–	203.2	–	–	3550	170	2.70 (0.476 Btu/hr-ft^2-deg F)
	Steel	15.9	235.0	203.2	45	461	7865	
	FBE	0.30	235.6	235.0	0.300	1500	1300	
	PP Adhesive	0.30	236.2	235.6	0.215	2000	900	
	PP Solid	6.00	248.2	236.2	0.220	2000	900	
	TDF	105.0	458.2	248.2	0.185	2000	740	
	PP Solid	4.0	466.2	458.2	0.220	2000	900	
8″ ID PIP Production Flowline WD = 1840m	Fluid (Methane)	–	177.8	–	–	3100	87	1.14 (0.200 Btu/hr-ft^2 deg F)
	Steel	15.9	209.6	177.8	45	461	7865	
	FBE	0.30	210.2	209.6	0.30	1500	1300	
	PUF	28.5	267.2	210.2	0.025	900	64	
	Steel	16.7	300.6	267.2	45.000	461	7865	

Thank you