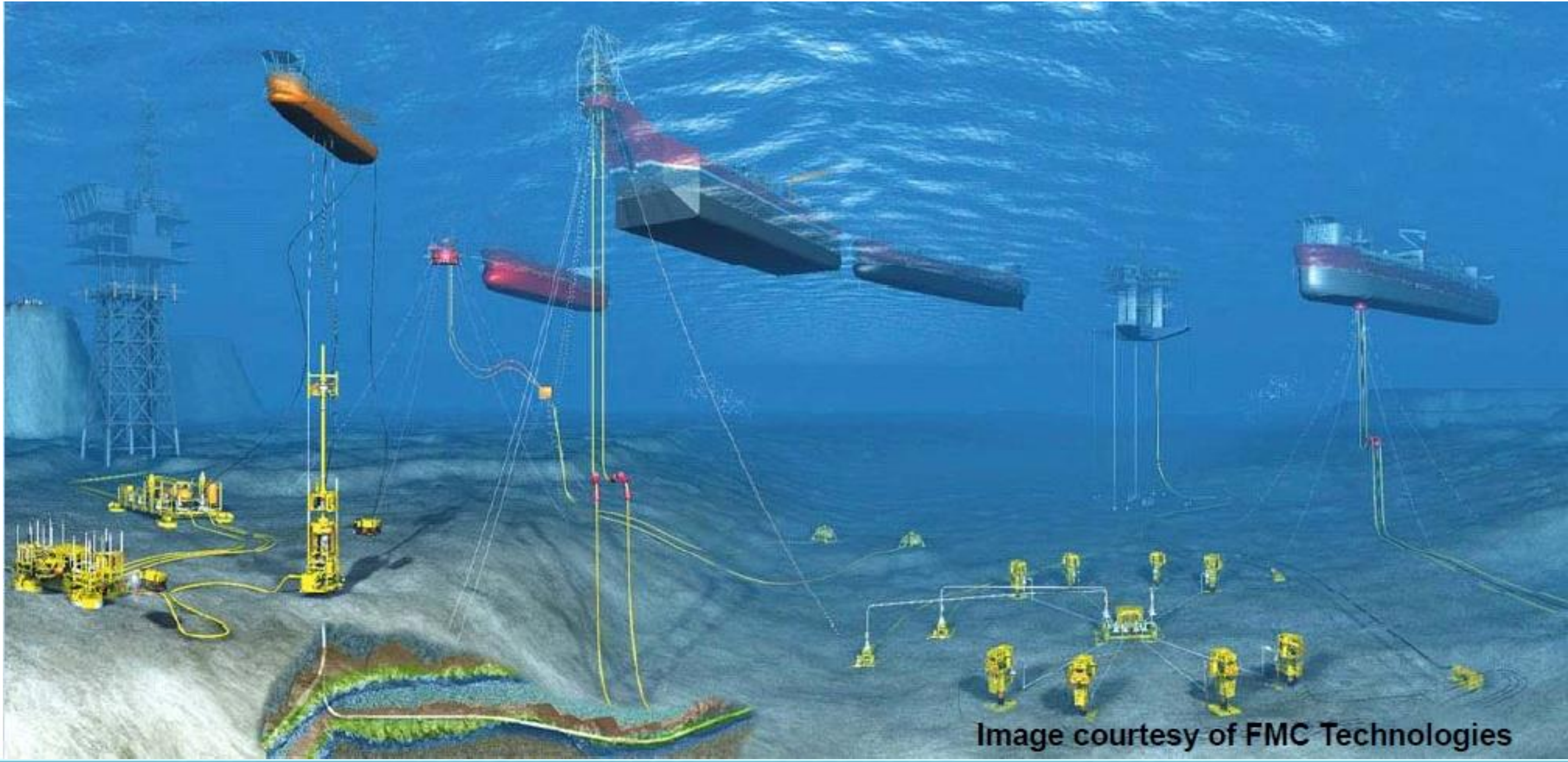


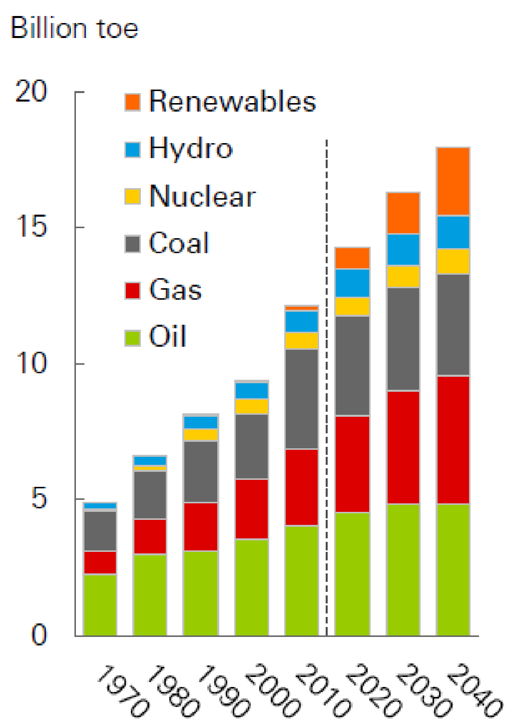
Offshore Equipment

Yutaek Seo

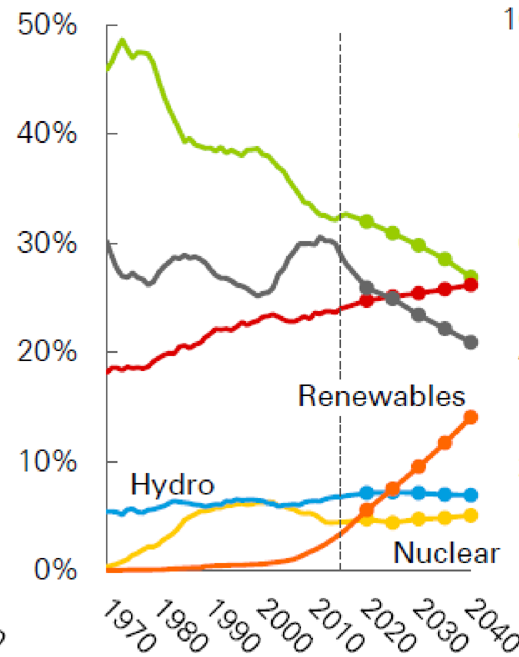
BP Primary Energy Consumption forecast (2018)

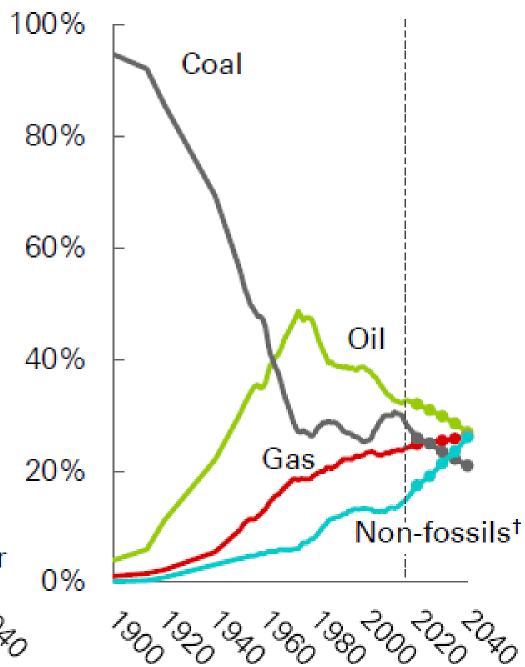
- Natural gas (1.6% p.a.) grows much faster than either oil or coal, with its share in primary energy overtaking coal and converging on oil.
- Oil grows (0.5% p.a.) until 2040, although is projected to plateau in the final part of the outlook. Coal consumption is broadly flat until 2040, with its share in primary energy declining to 21%, its lowest share since the industrial revolution.

Primary energy consumption by fuel



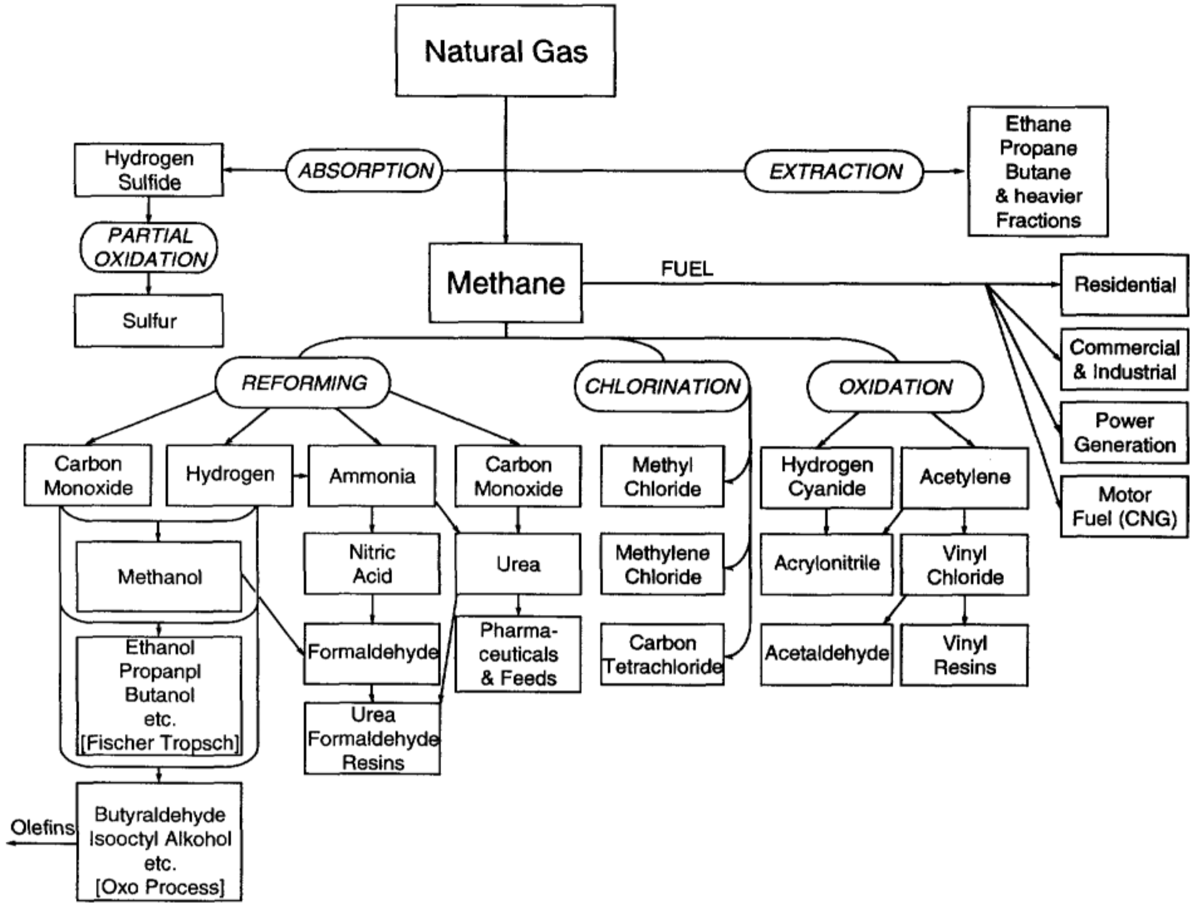
Shares of primary energy





Natural gas

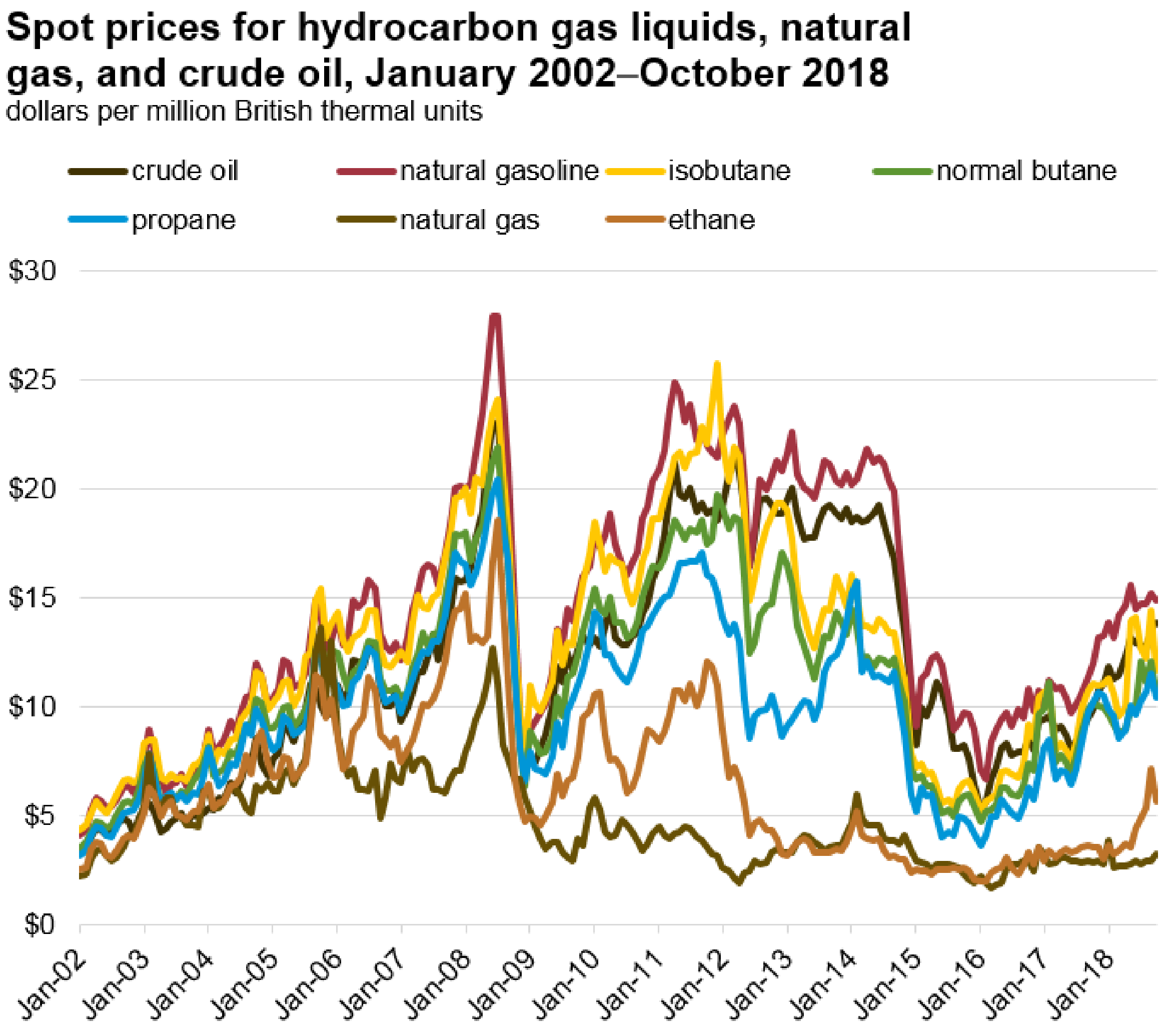
- The two main uses for natural gas are fuel (residential, commercial, industrial, power generation, and transportation) and chemical manufacturing feedstock.
- Worldwide, the primary chemicals manufactured from natural gas are methanol and urea (fertilizer).



Natural gas demands and price history

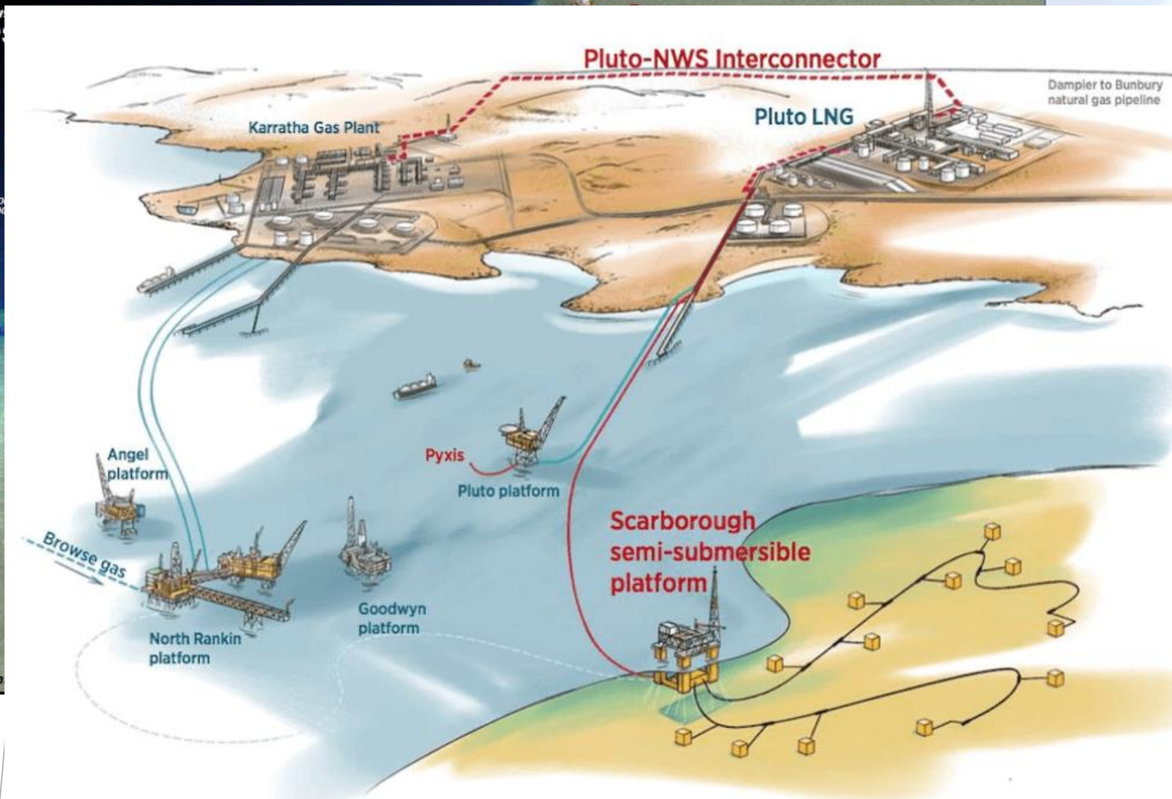
- The mobility of LNG cargoes and their ability to be diverted in response to price signals causes the gas market to become increasingly integrated, with movements in global gas prices becoming more synchronized.

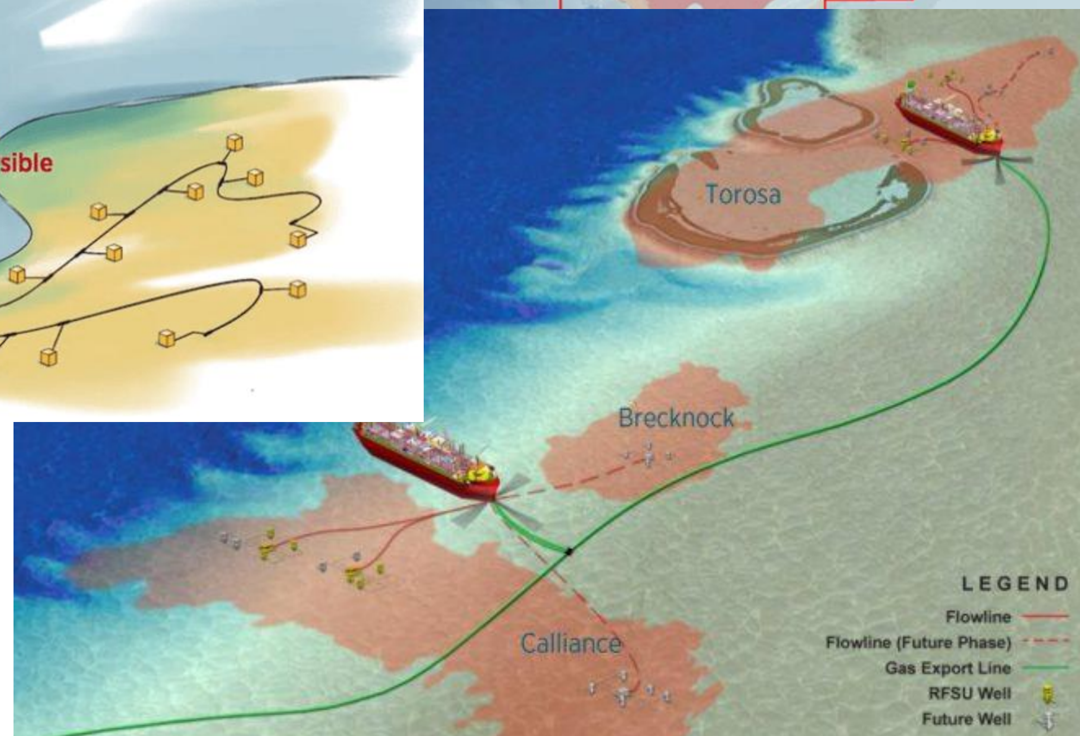
NGL prices



Note: Prices are monthly average of close-of-day spot prices; crude oil is Brent; natural gas is Henry Hub; HGL products are at Mt. Belvieu non-LST (Lone Star Terminal).
Source: U.S. Energy Information Administration from Bloomberg.

Long distance gas pipeline & gFPSO



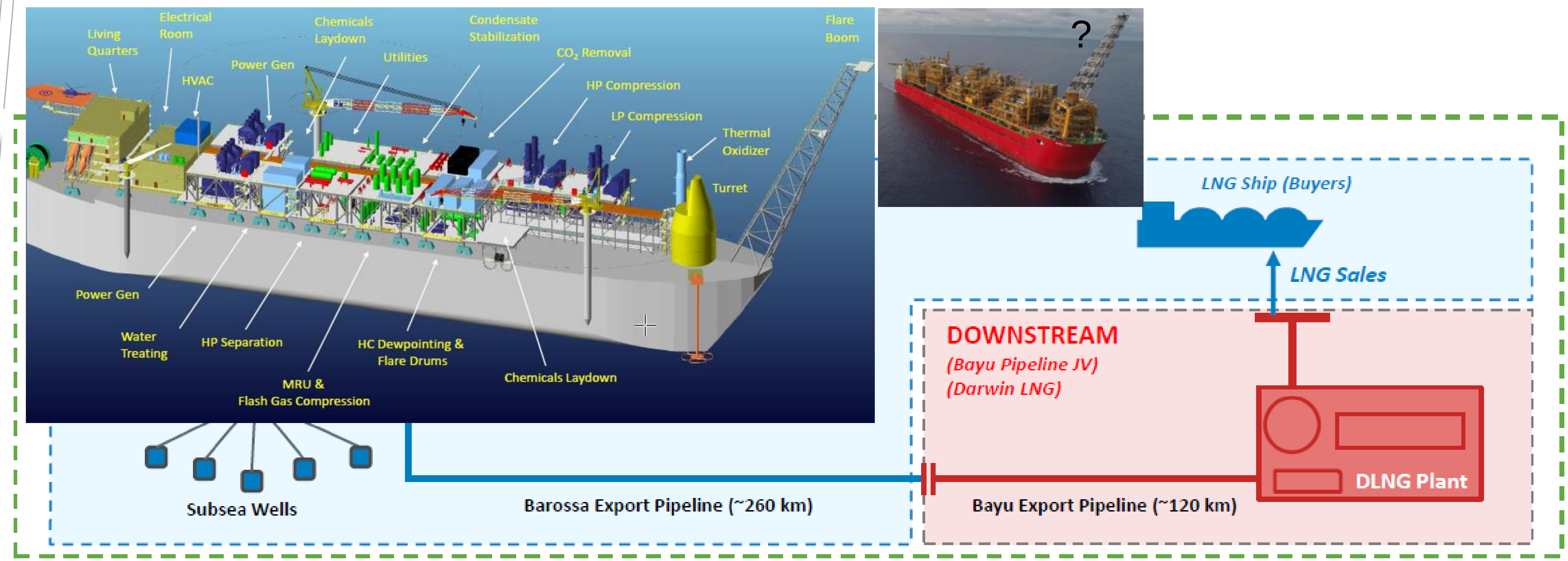


- Woodside is leaning towards the development using two gas Floating Production Storage and Offloading Units (gFPSO) delivering around 10 mtpa of gas to NWS infrastructure by an approximately 900 km pipeline. (2018. 10. 11)

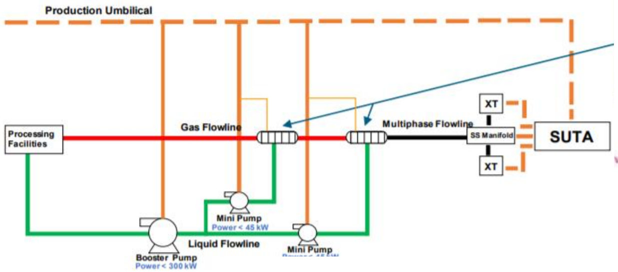
gFPSO (gas-only FPSO)

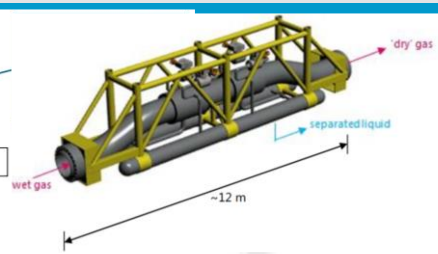
- A GFPSO would essentially be a floating gas production and conditioning facility. Principal export products from a GFPSO would be a LPG liquid, a C5+ condensate liquid and pipeline quality residue gas.
- Technip FMC has developed gas FPSO (Barossa, Tortue and Abadi projects) which is an alternative to FLNG to develop stranded offshore gas fields to export gas for feeding either existing new onshore LNG plants or domestic market.

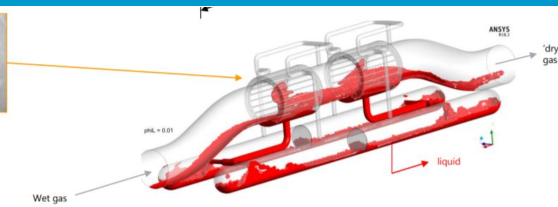
  : From lean to rich gas (LPG can also be produced)
  : Capability and Knowledge to process any feed gas flow characteristics
  : HSE design capacity to propose mixed solutions (safety gaps, fire walls)

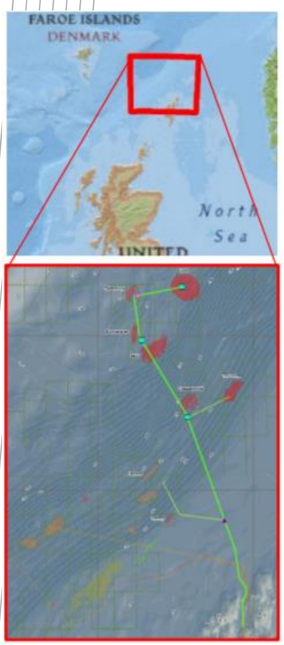


Pseudo Dry Gas







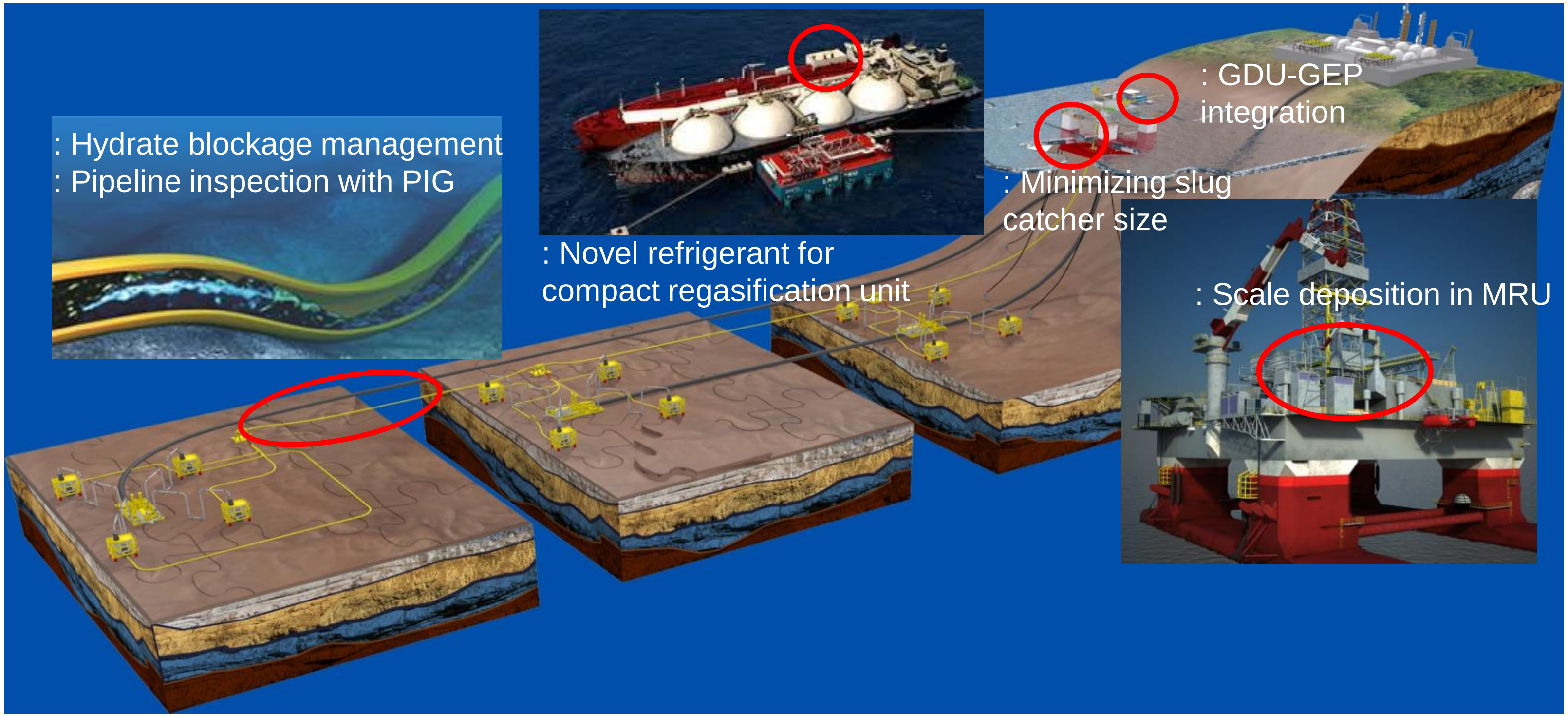


	FPSO, dry gas export	Subsea Tie-back – single flowline	Subsea Tie-back – dual flowlines	WGC - proven (dP=32 bar)	WGC- future (Pr=2)	Pseudo Dry Gas
Optimum line size	2 x 10" (risers)	1 x 22"	2 x 18"	2 x 18"	2 x 18"	1 x 30"
WHP to deliver 500 MMscfd @ HP	114 bar	168 bar	168 bar	132 bar	84 bar	102 bar
Trunkline pressure drop	No wet gas trunkline	98 bar	98 bar	98 bar (incl. 32 bar WGC)	98 bar (incl. 84 bar WGC)	32 bar
Slug (surge) volume generated from short term turn down and ramp-up 100%->50% to 100% *(100%->25%->100%)*		2421 m³ *(9208 m³)*	1034 m³ *(4416 m³)*	1034 m³ *(4416 m³)*	1034 m³ *(4416 m³)*	negligible
Subsea Power	No subsea power demand			7 MW	20 MW	0.5 MW

- FPSO cannot be tied back to distant fields (low WHP for WD=1600m, JT cooling -23°C)
- Single or dual flowlines induces liquid management challenges (large slug catcher)
- Compressor requires high electricity requirement.

Research topics for long distance gas pipeline

- Multiphase flow inside flowlines
  : Slug management at an arrival facilities
- Vapor-Liquid-Solid phase changes
  : Hydrate blockage risk management with MEG or LDHI (KHI/AA)
  : Avoiding scale deposition in subsea flowlines and topside process
  : Novel refrigerant for regasification of LNG



Long pipeline

Flowlines, Manifolds and Piping

- Equipment used to transmit produced fluids from wellhead through treating equipment
  - Piping
  - Connections
  - Valves
  - Fittings
- Flowlines: Usually 2” to 16”
  - API steel line pipe
    - : Standard 5 L < 1000 psi
    - : Strandard 5 LX > 1000 psi
- Pipe:
  - Closed conduit
  - Circular cross-section
  - Constant internal diameter (ID)

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

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

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

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

- 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

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

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

Pressure change due to hydrostatic head

Frictional pressure gradient

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

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

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

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

- 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:

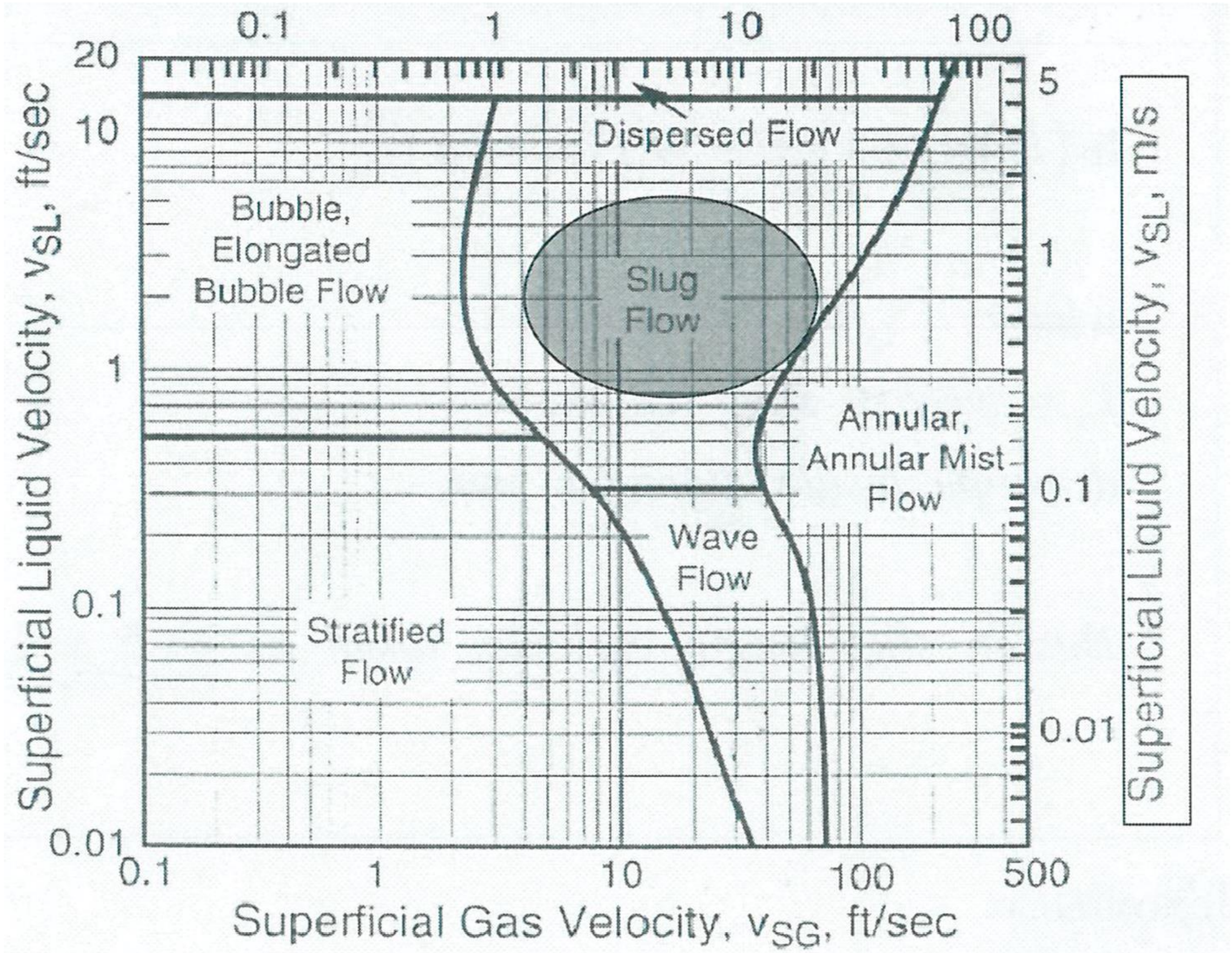
$$\mu_m = \mu_L \lambda_L + \mu_g \lambda_g \;,\; \mu_s = \mu_L^{H_L} \times \mu_g^{H_g}$$

- Liquid Surface Tension (σ_L):

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

Flow pattern map of Mandhan

- Horizontal two phase flow in pipes



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

Slugging during ramp up and pigging

- Ramp Up:

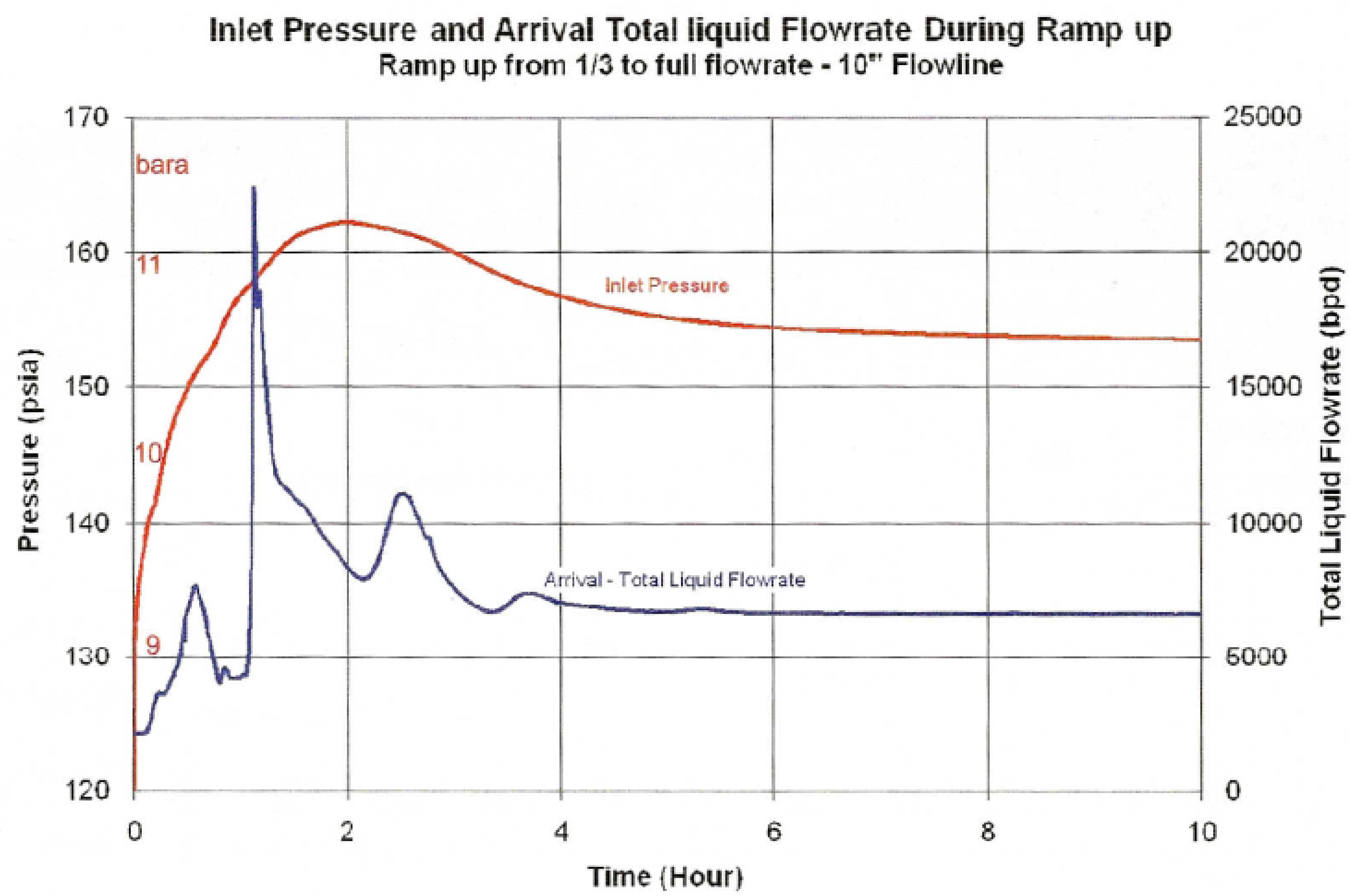
  : Total Liquids Produced

  = holdup at the lower flowrate (minus) holdup at the higher rate.

  : The actual liquid production rate during this period will depend on the fluids, the flowline design and the flow conditions.

- Pigging: The greatest effects on liquid production during pigging occur with gas condensate flowlines. The entire flowline liquid holdup (except for the pig by-pass volume) will be produced in front of the pig.

Ramp up flowrates and pressure



Hydrodynamic slug size prediction (FPS vs SI)

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

  where :

  L_s = *average slug length , ft*

  d = *pipe inside diameter , in*

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

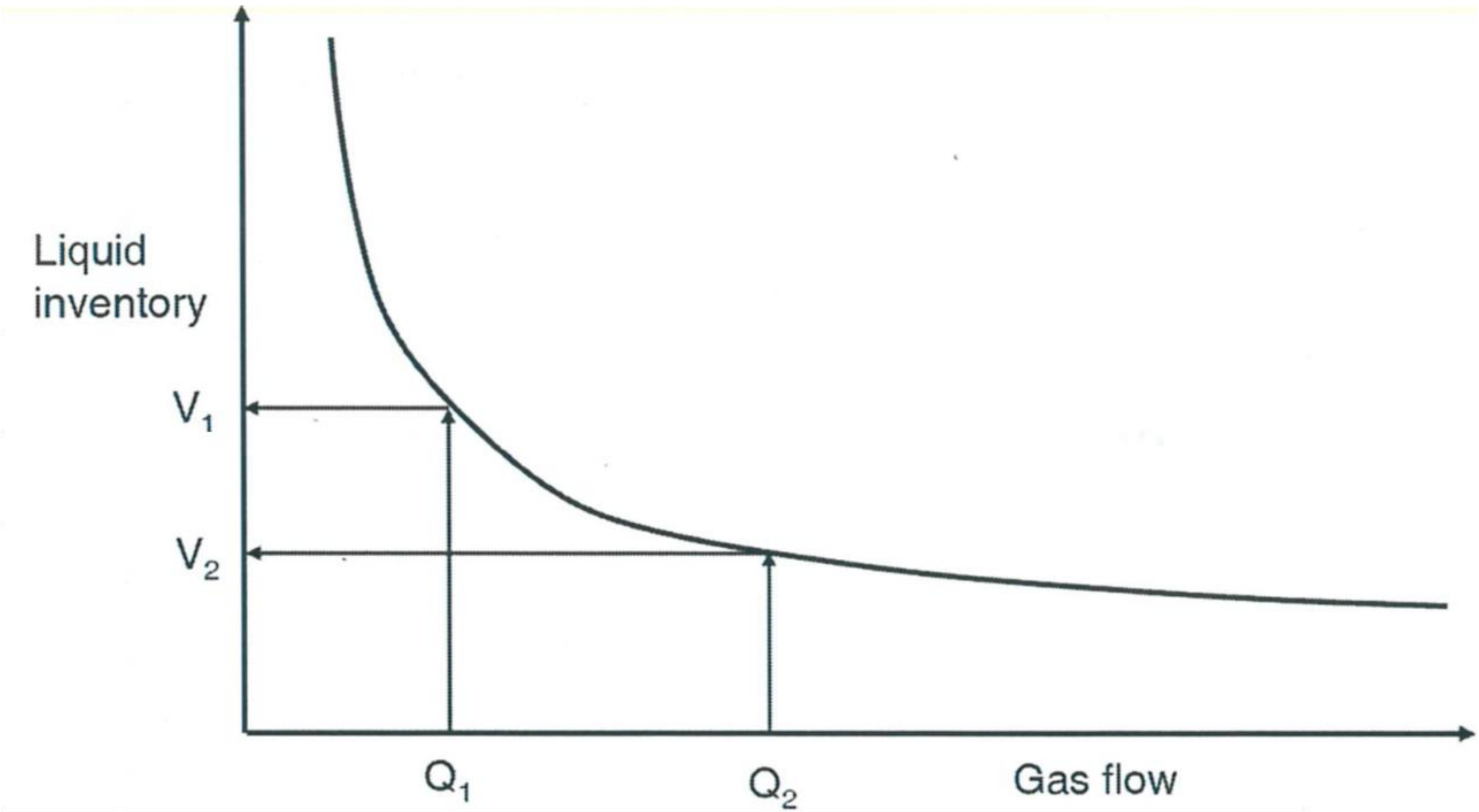
  where :

  L_s = *average slug length, m*

  d = *pipe inside diameter, mm*

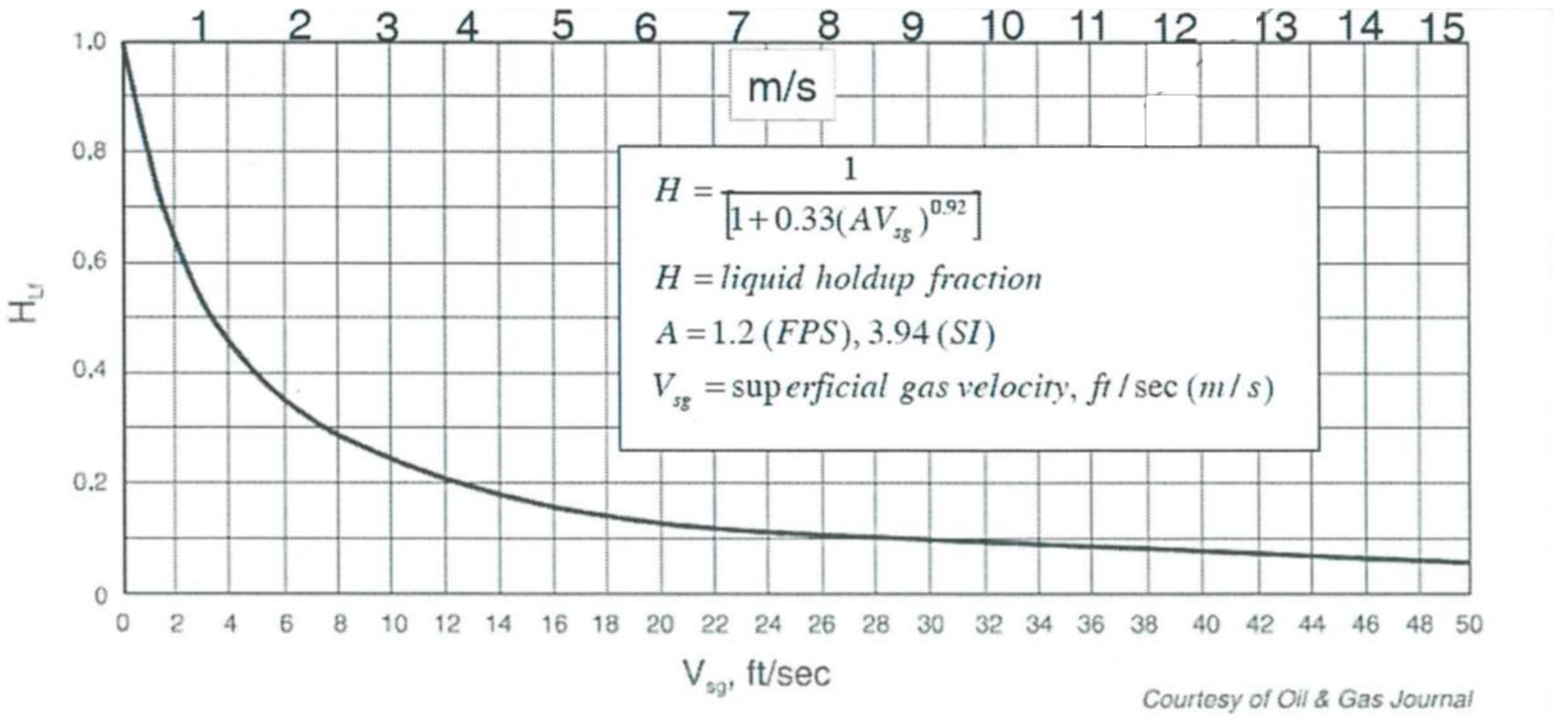
- Design slug length typically taken as 4 ~ 5 times L_s

Pipeline liquid holdup



Simple holdup correlation - Flanigan

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



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

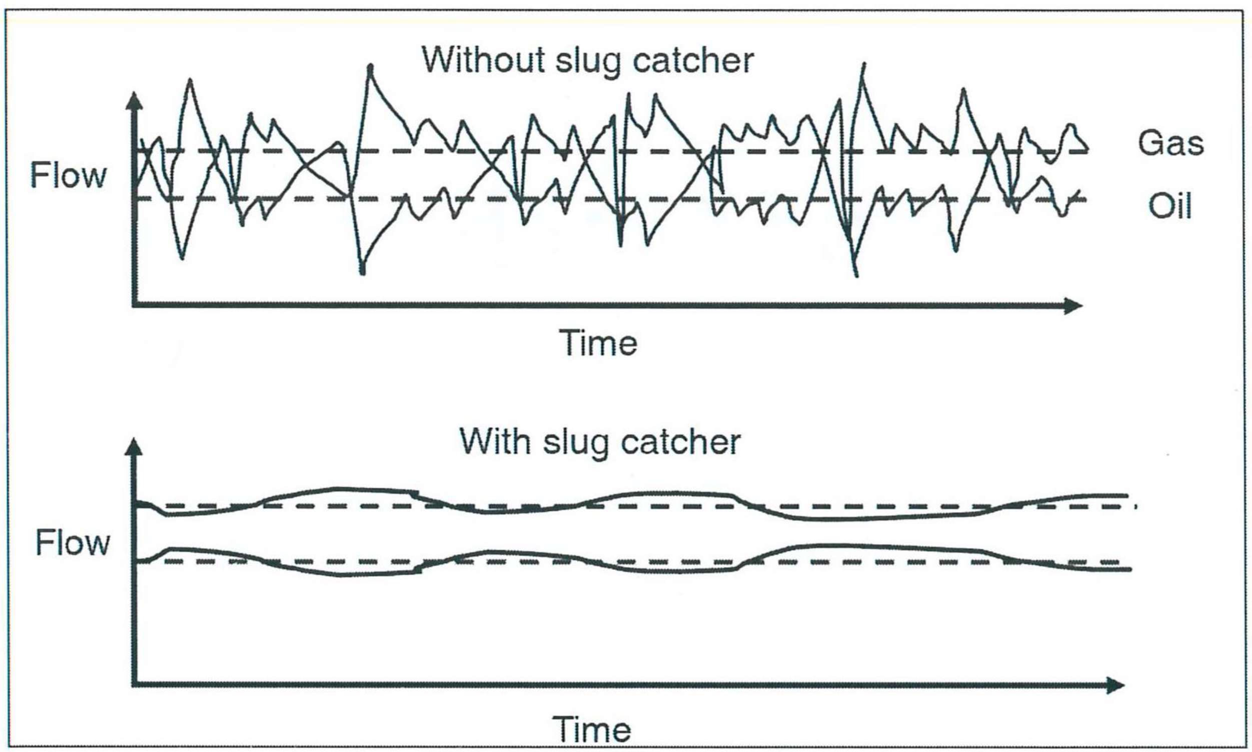
q_g in MMm^3/d, T in K, d in m, P in kPa, A in 5.19 (SI)

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



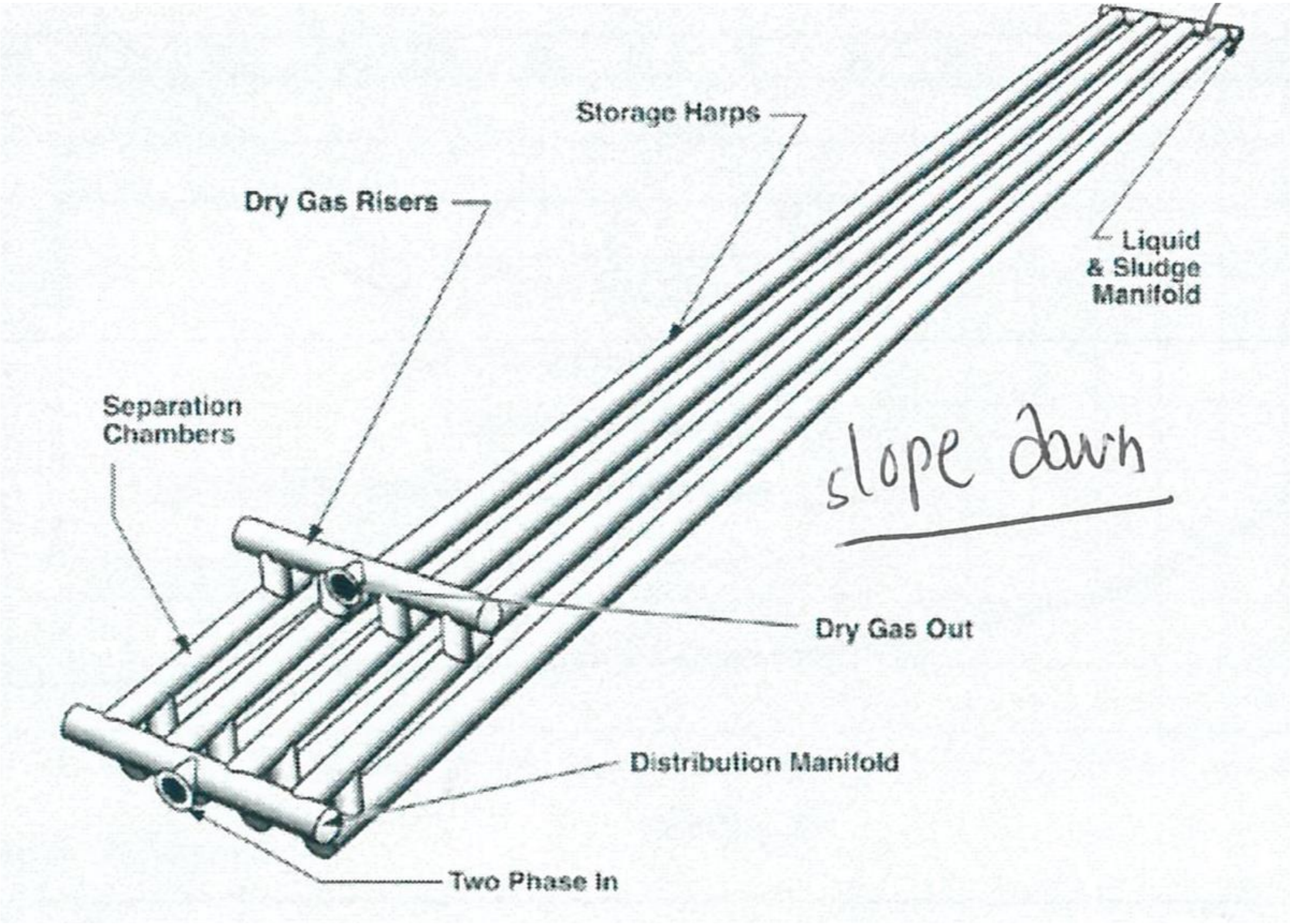
Slug catcher configurations

- This section briefly describes two kinds of slug catchers, manifolded piping and inlet vessels.
- The most difficult part of a slug catcher design is the proper sizing. Sizing requires knowledge of the largest expected liquid slug, as liquid pump discharge capacity on the slug catcher will be trivial compared with the sudden liquid influx.
- Manifolded Piping

  : One reason piping is used instead of separators is to minimize vessel wall thickness. This feature makes piping attractive at pressures above 500 psi (35bar).

  : The simplest slug-catcher design is a single-pipe design that is an increased diameter on the inlet piping. However, this design requires special pigs to accommodate the change in line size.

Finger type slug catcher

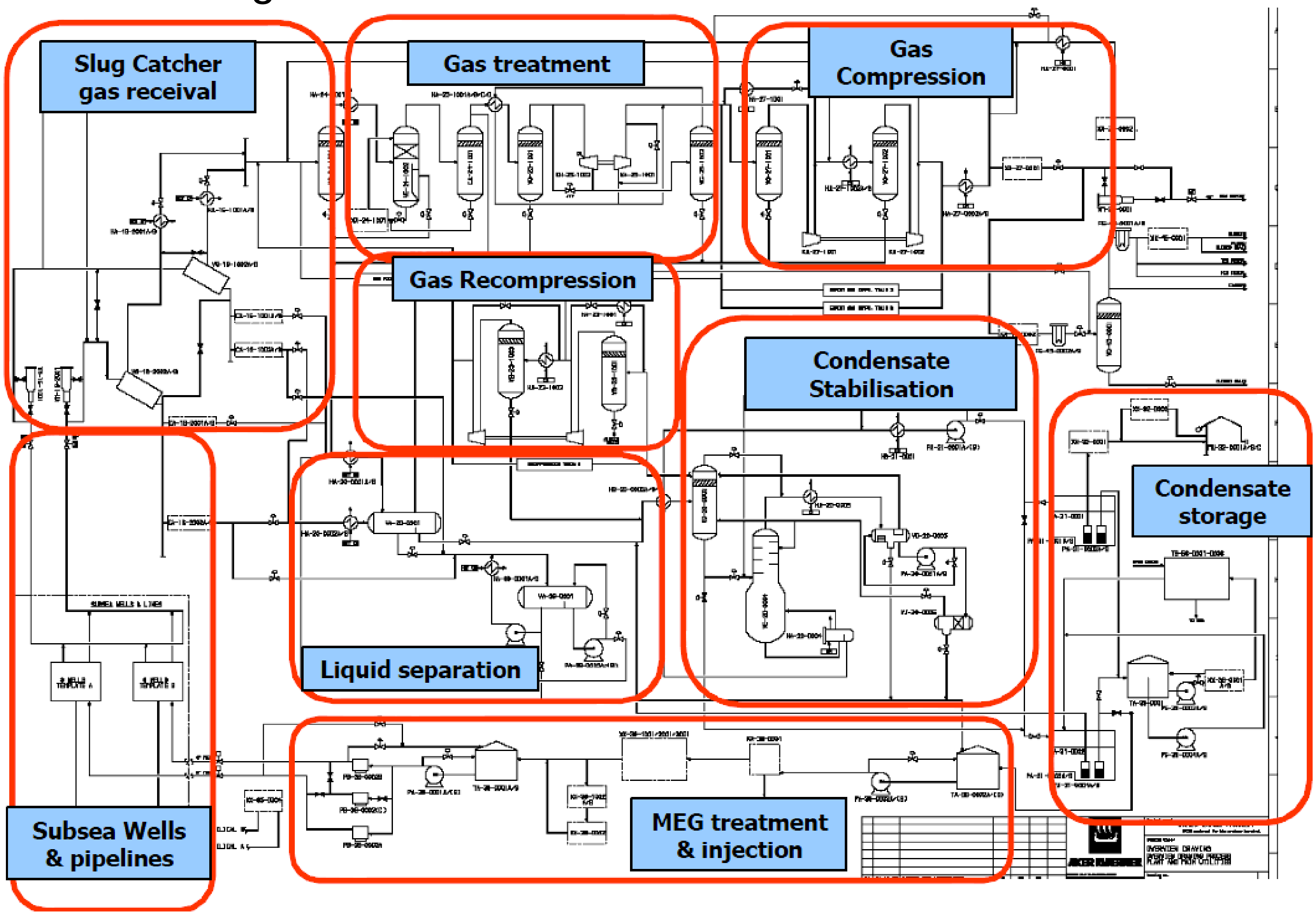


- The number of pipes varies, depending upon the required volume and operating pressure. Also, some designs include a loop line, where some of the incoming gas bypasses the slug catcher.
- Primary separation occurs when the gas makes the turn at the inlet and goes down the pipes. Liquid distribution between pipes can be a problem, and additional lines between the tubes are often used to balance the liquid levels. In harp designs, the pipes are sloped so that the liquid drains toward the outlet.
- Gravity settling occurs as the gas flows to the vapor outlet on the top while the liquid flows out the bottom outlet.
- Pipe diameters are usually relatively small(usually less than 48 inches [120 cm]), so settling distances are short.

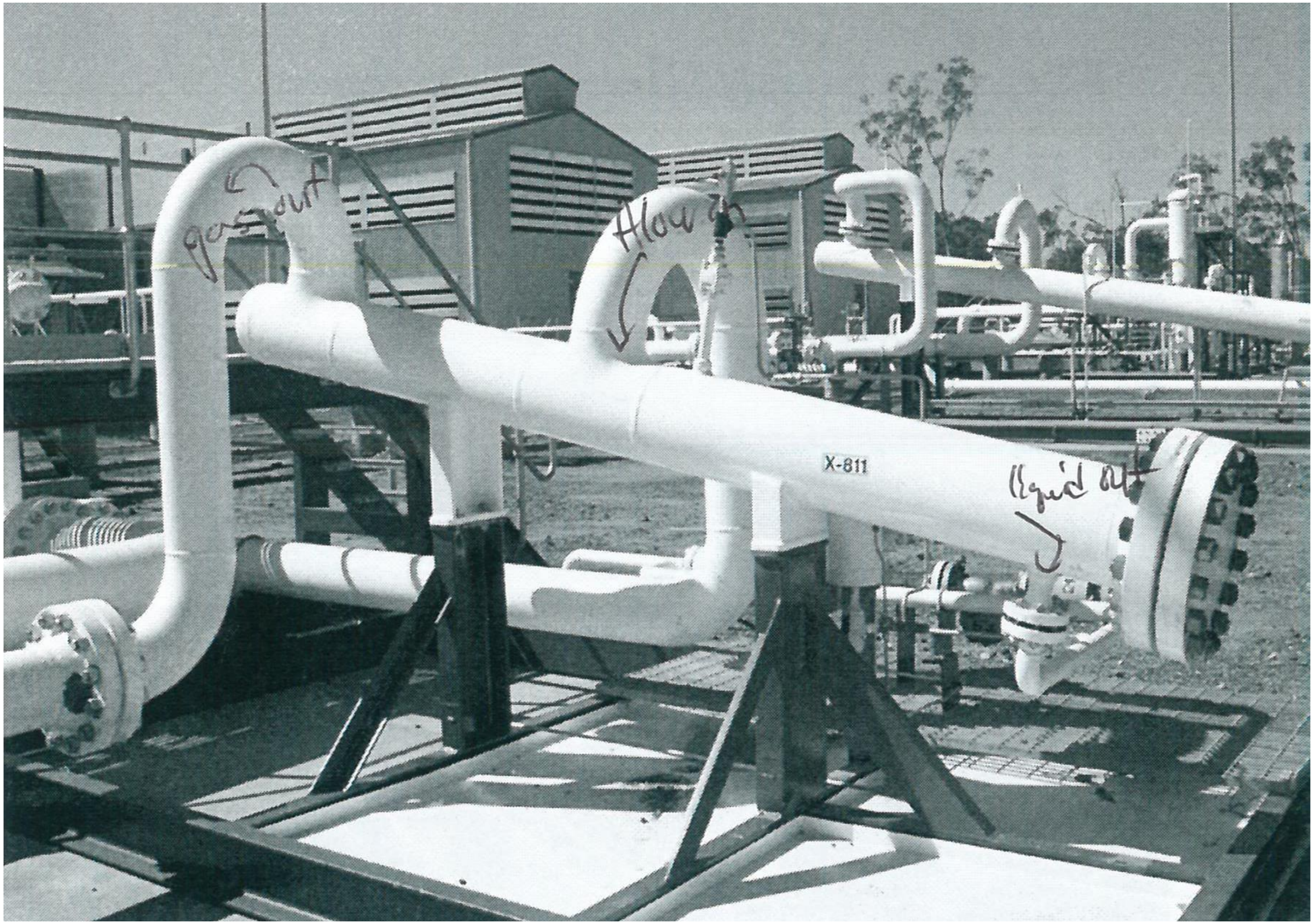
- Because manifolded piping is strictly for catching liquid slugs, demisters are usually installed downstream in scrubbers. Likewise, liquid goes to other vessels, where degassing and hydrocarbon-water separation occurs.
- Several advantages to the pipe design include the fact that design specifications are based upon pipe codes instead of vessel codes.
- Also, the slug catcher can be underground, which reduces maintenance costs and insulation costs if the slug catcher would otherwise need to be heated.

Gas condensate flowline slug catcher

- Ormen Lange

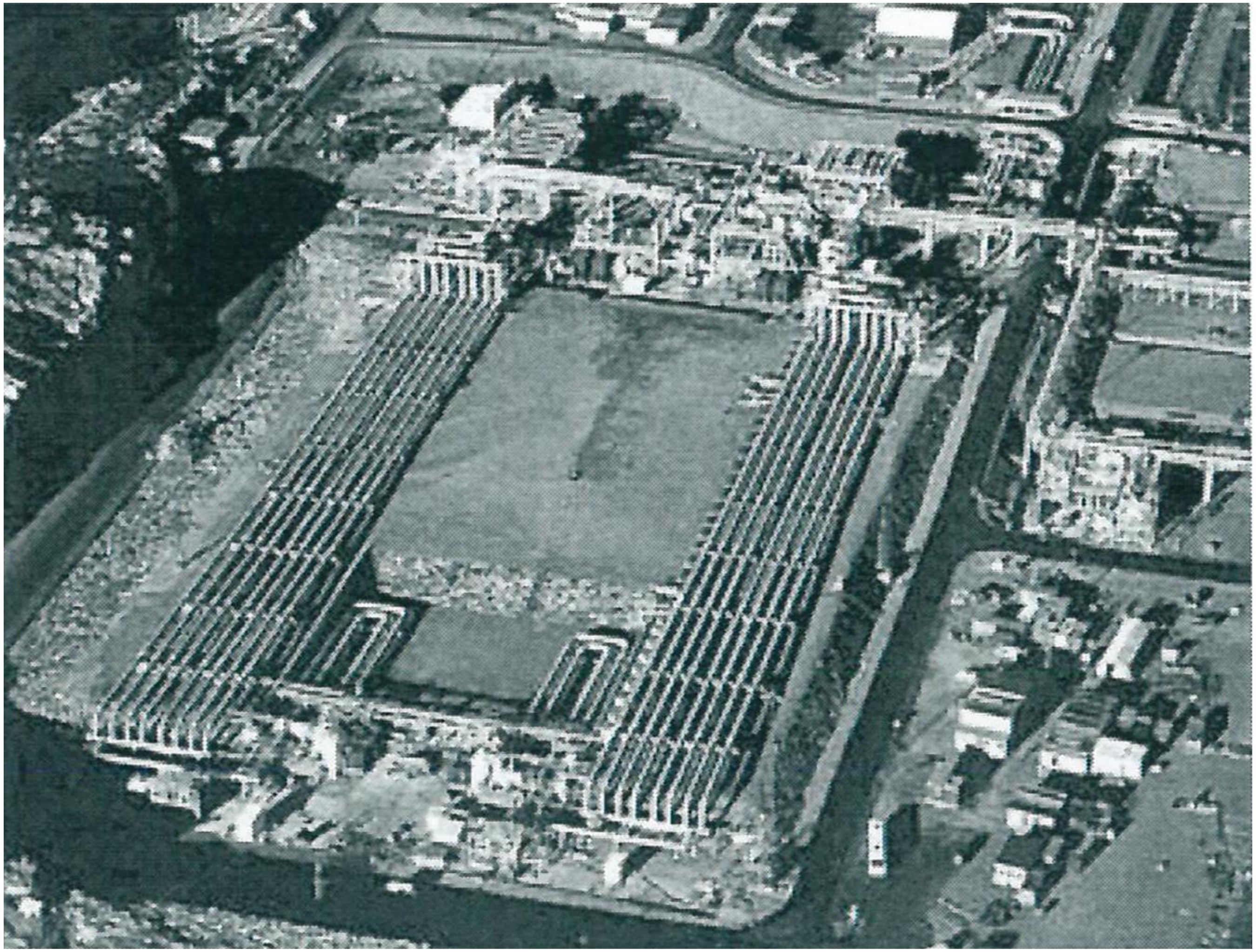


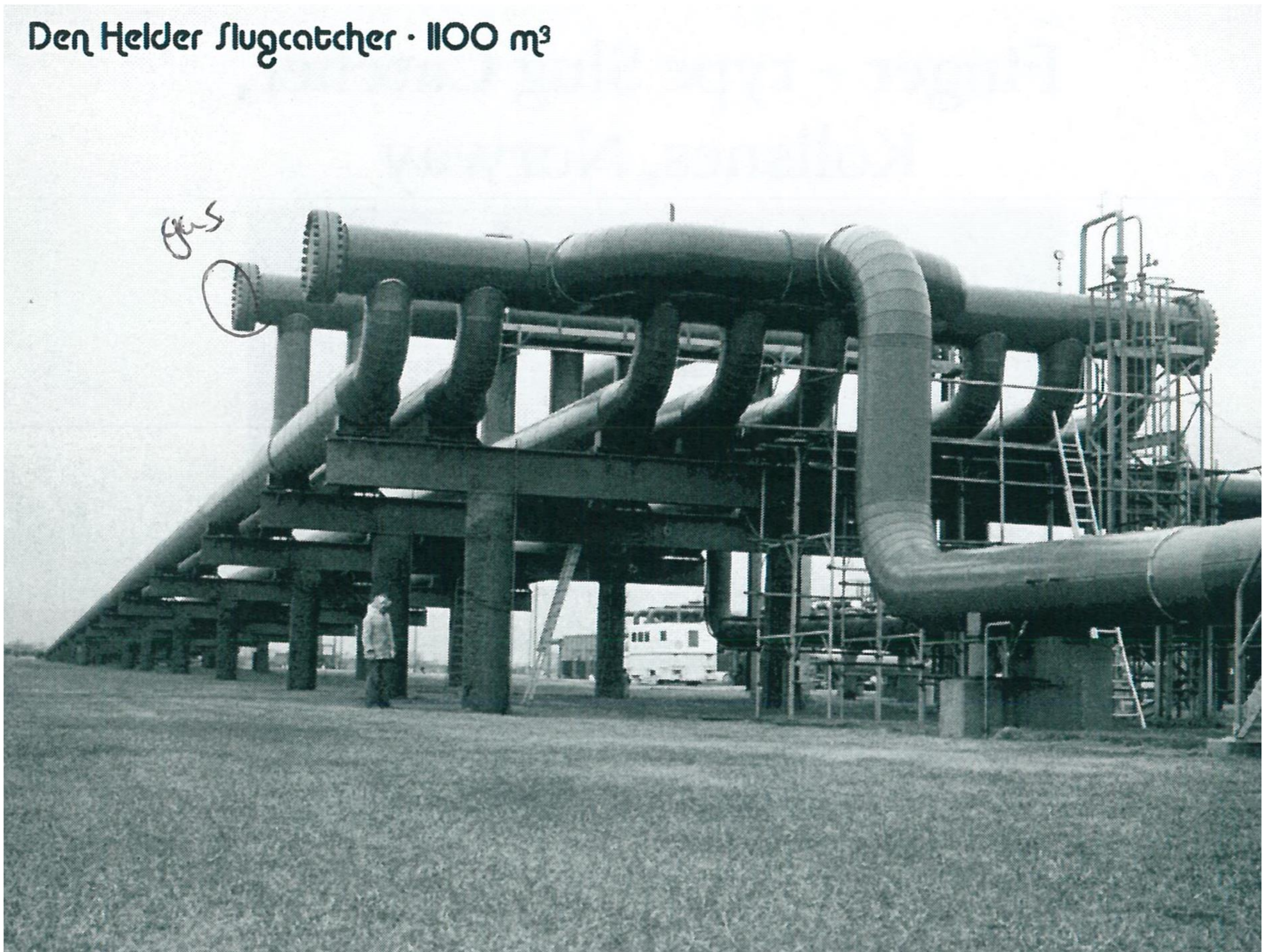

2 Slug Catcher units 1500m³ capacity
Can be connected to provide 3000m³ capacity


gas out
Flow in
X-811
liquid out

Finger type slug catcher

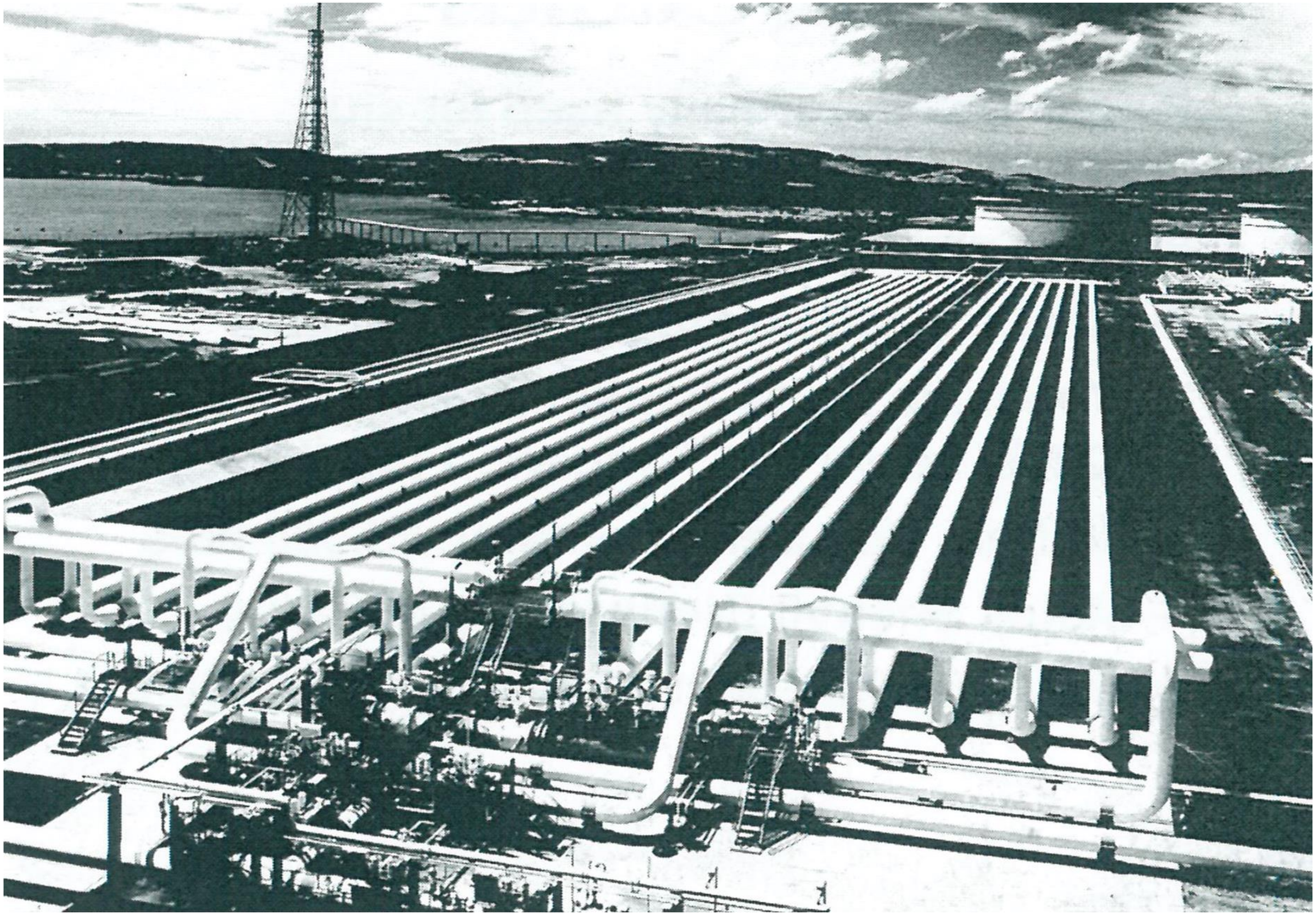
- Kollsnes, Norway





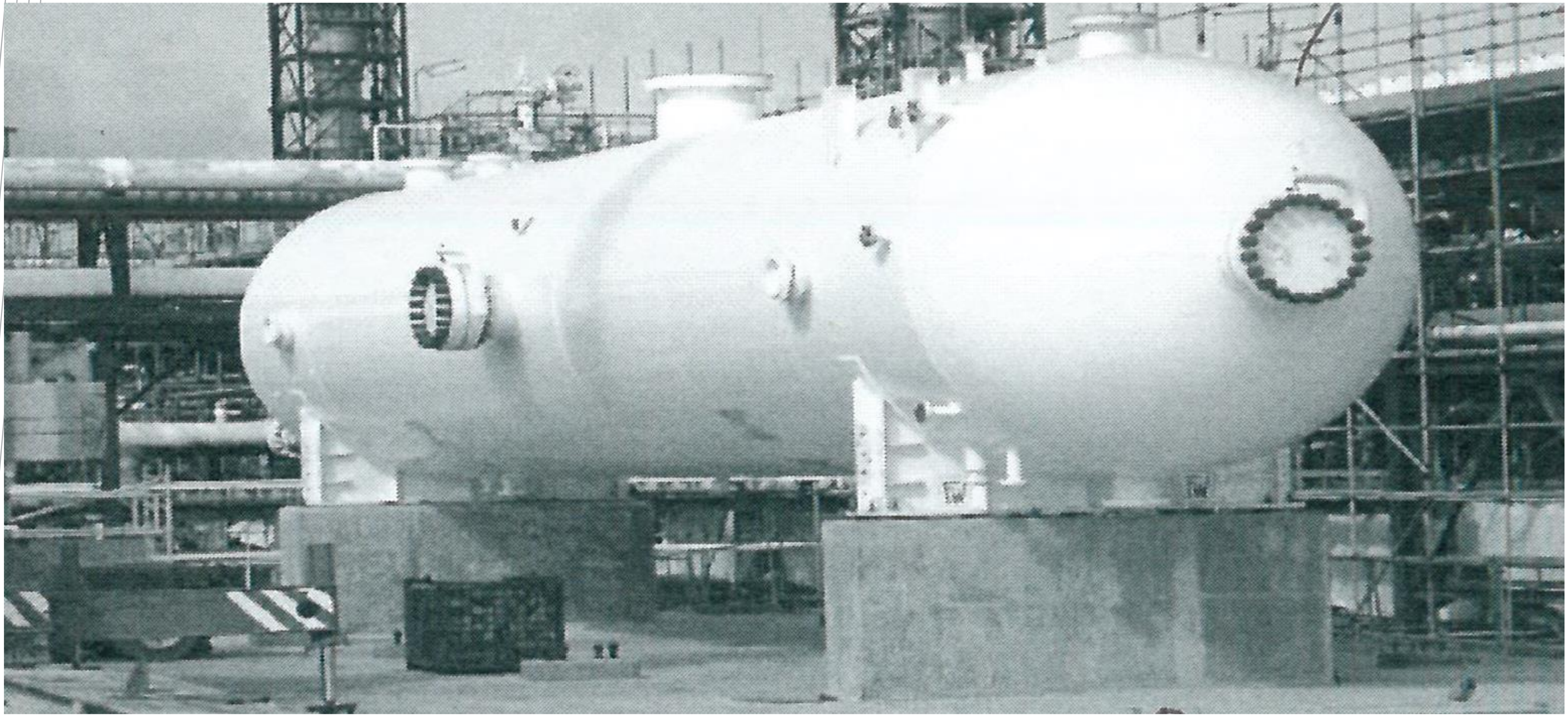
Den Helder Slugcatcher · 1100 m³

- 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

Thank you

Exercise.

- 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				

Exercise. Determine pressure drop

- In-situ Vapor MW = 20.6, 25 inch pipe
- Q_l =2000 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

* For all gases: 2635 lbmole/mmscf, g = 32.17 ft/sec^2