Liquefaction Process Evaluation for Floating LNG

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Outline

- Introduction
- Expander processes for natural gas liquefaction
- Proposed process schemes for FLNG
- Basis for process evaluation and comparison
- Result and discussion

Introduction

Background

- A challenge to find the optimal liquefaction process technology for use in offshore environment
- Growing interest to apply expander-based liquefaction processes for FLNG
- A challenge to get objective comparison between the various technology efficiencies

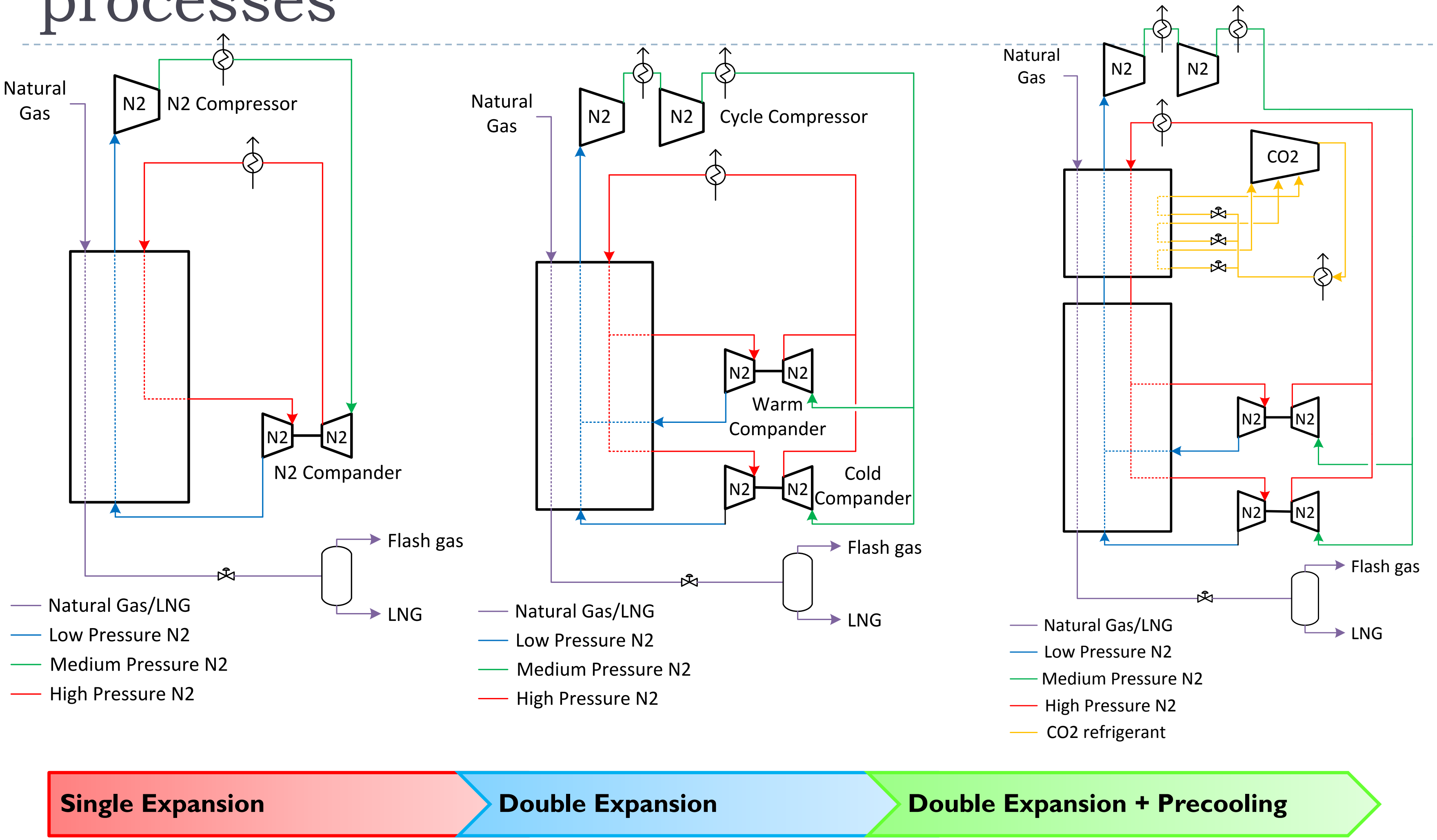
Objective

- To do a comparative evaluation of several expander-based processes for FLNG on a identical basis focusing on capacity, efficiency, integration into energy system, complexity, and hydrocarbon inventory

Scope of work

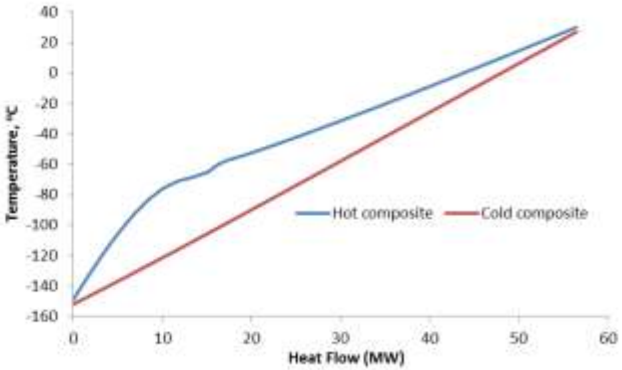
- Literature study, establishment of a identical evaluation basis, HYSYS model development, case studies and systematic comparative analysis of performance data resulted from HYSYS

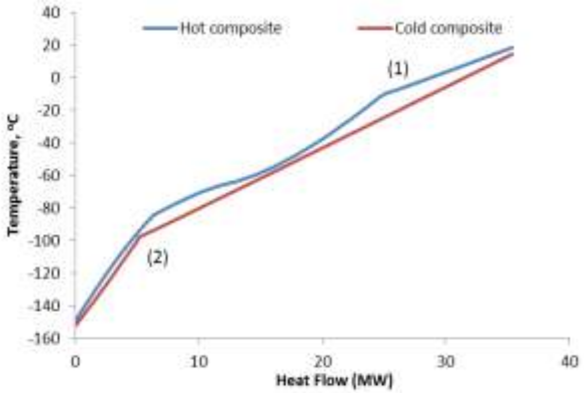
A typical expansion-based liquefaction processes

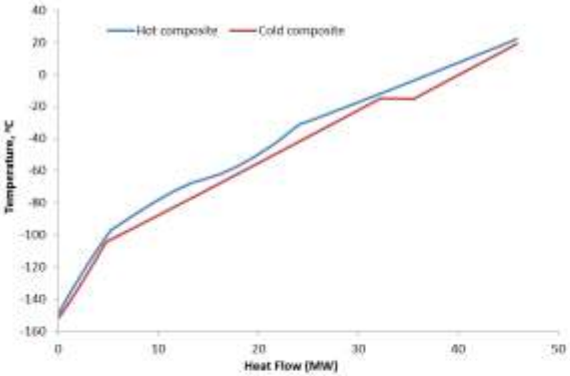


Efficiency and Capacity Improvement

The Cooling Curves







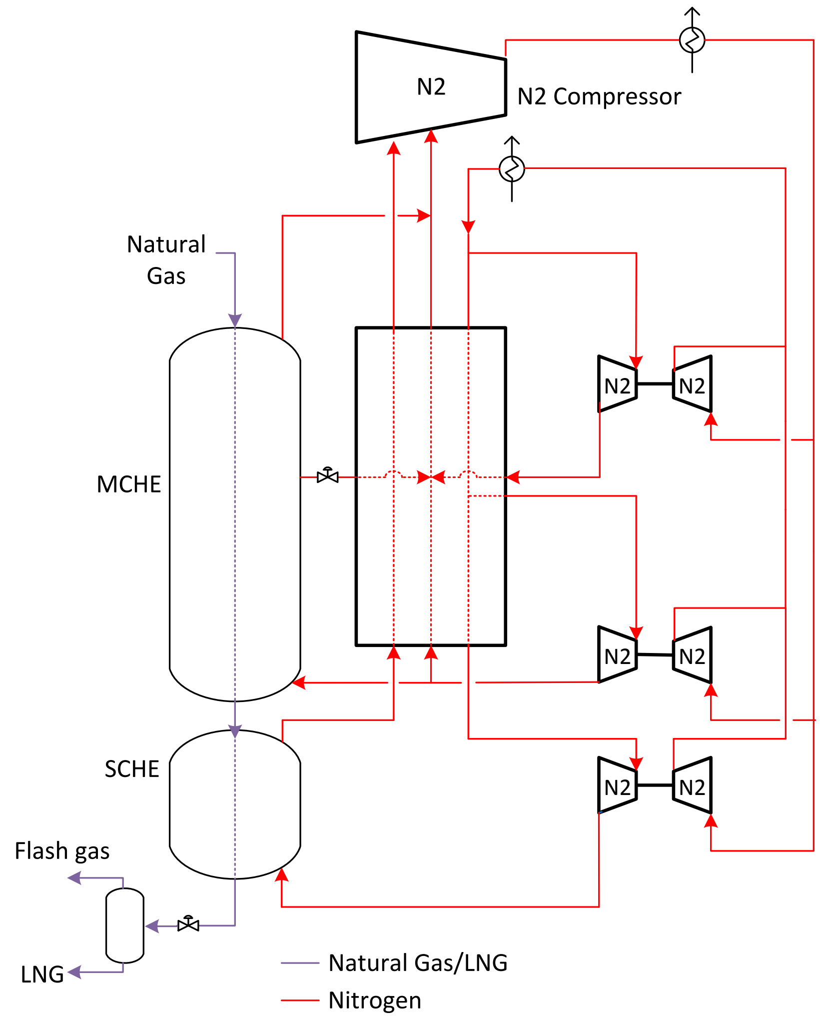
Single Expansion → **Double Expansion** → **Double Expansion + Precooling**

Efficiency and Capacity Improvement

Proposed Expander Process Schemes for Floating LNG

Process A

- Based on US Patent 2010/0122551 A1
- N2 expander-based process with two pressure levels and three expander temperatures
- ARS (LiBr/water) is used for precooling system
- The LiBr process driven by gas turbine waste heat and provides cooling of feed gas, N2 loop and gas turbine air intakes.

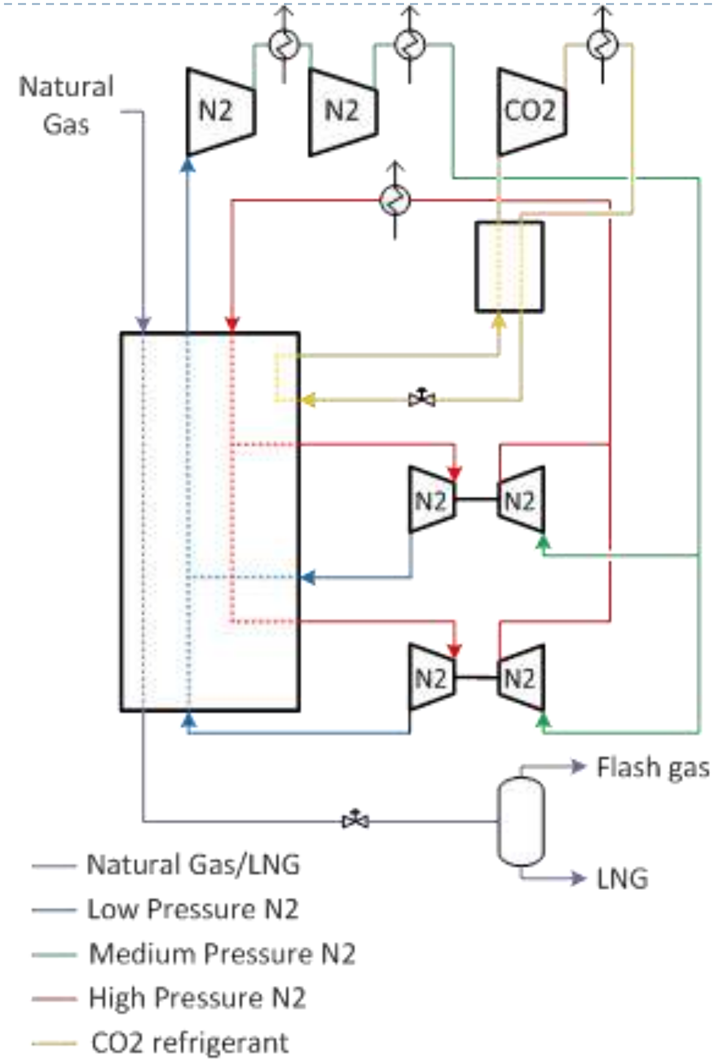


ARS :Absorption Refrigeration System

Proposed Expander Process Schemes for Floating LNG

Process B

- Based on US Patent 5,768,912
- a CO_2 precooled dual nitrogen expander liquefaction process
- the CO2 cooling is also used for gas turbine air intake cooling.

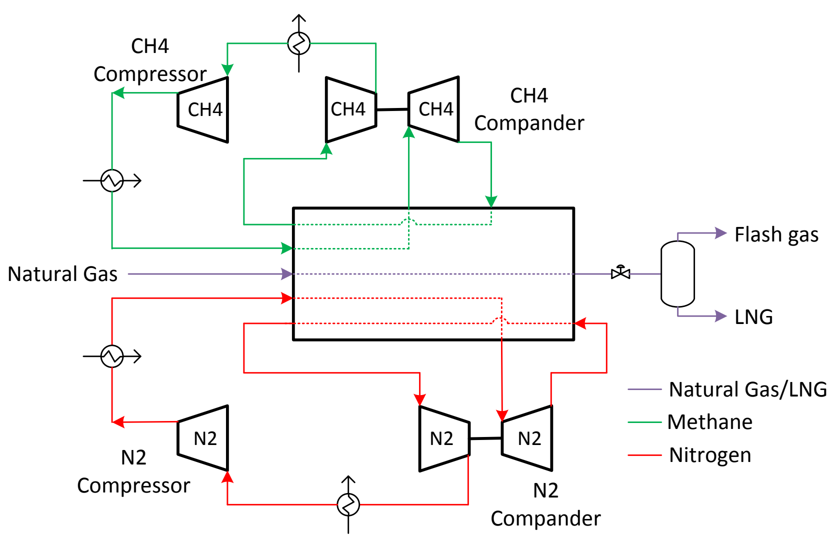


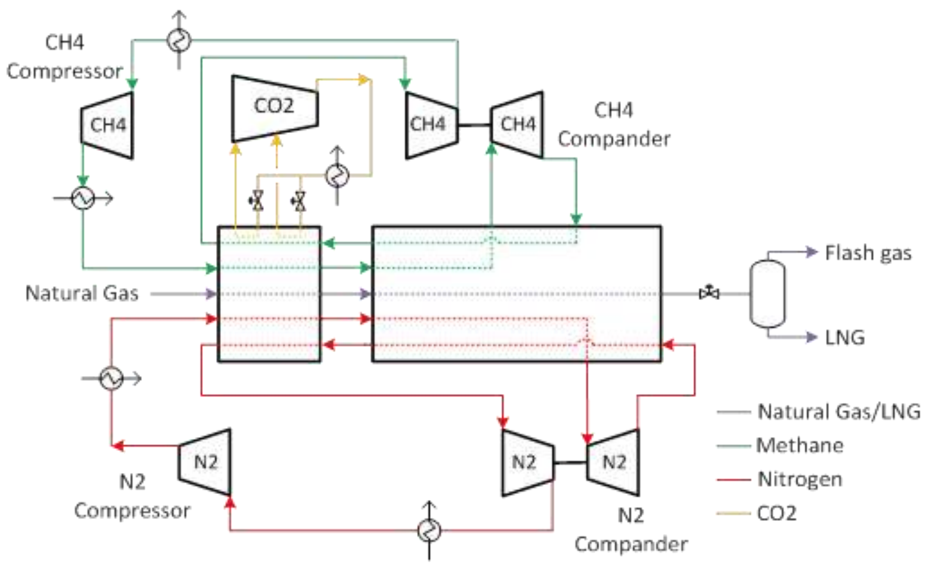
Proposed Expander Process Schemes for Floating LNG

Process C

Process C with precooling

Based on US Patent 6,412,302





Evaluation and Comparison Basis

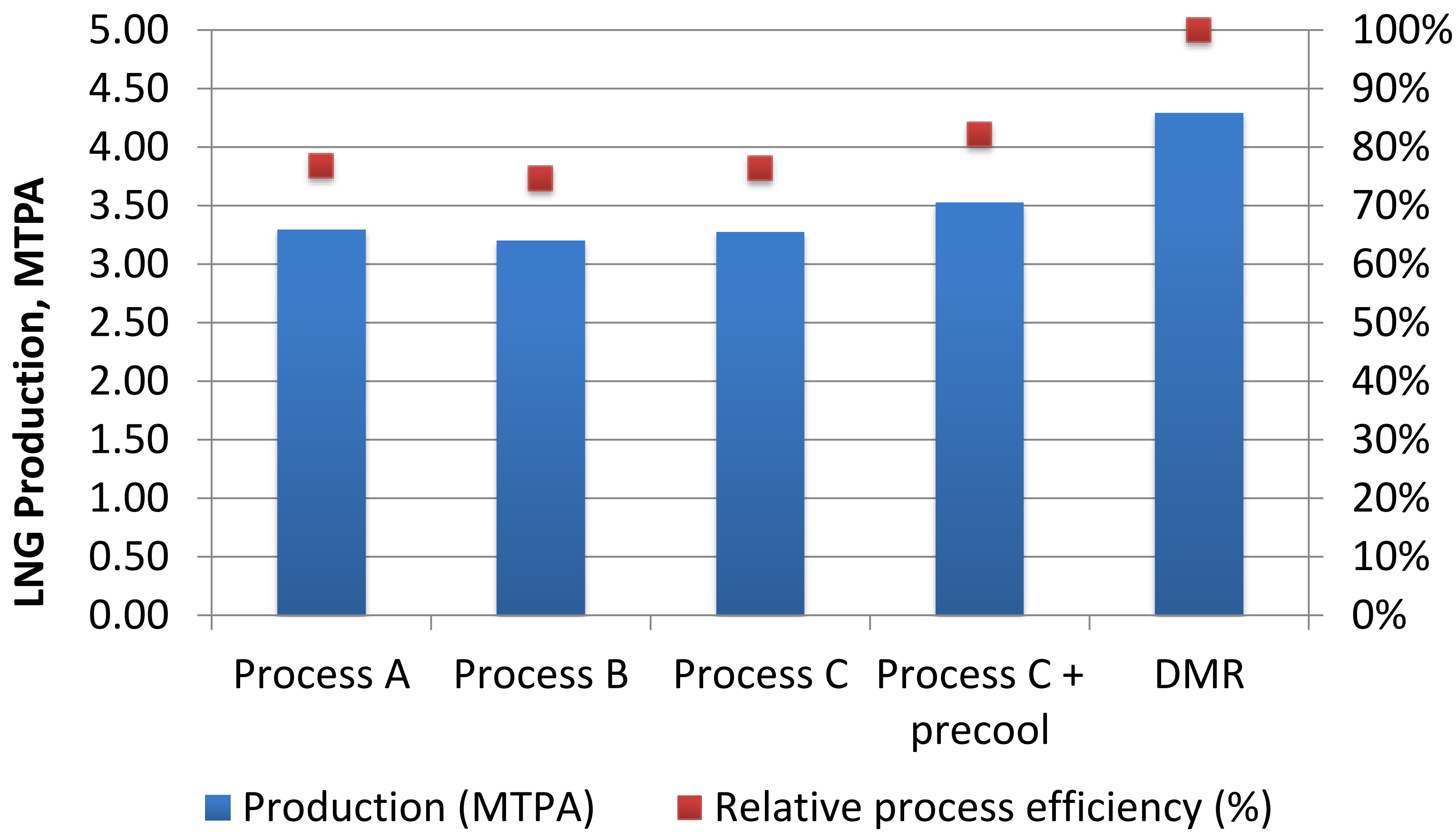
Simulation scope	• The pretreating units are out of the scope • Covers the process of the treated feed gas to LNG product
Feed gas	• Same condition (60 bar, 22°C) • Medium gas (91% of methane)
Site conditions	• 27 °C of air temperature • 17 °C of cooling water temperature • 5 °C min temperature approach of cooling water cooled HX
Drivers	• GT GE LM6000 • At 27 °C air GT gives 35 MW
Cryogenic HX	• 3 °C min temperature approach • 85 bar max pressure
Component efficiency	• Compressor polytropic efficiency of 78% • The compander polytropic efficiency was assumed of 73% for the compressor and 83% for the expander
LNG product	• Same condition (1.38 bar, -149 °C) • At LNG spec

Result & Discussion

Key Parameter

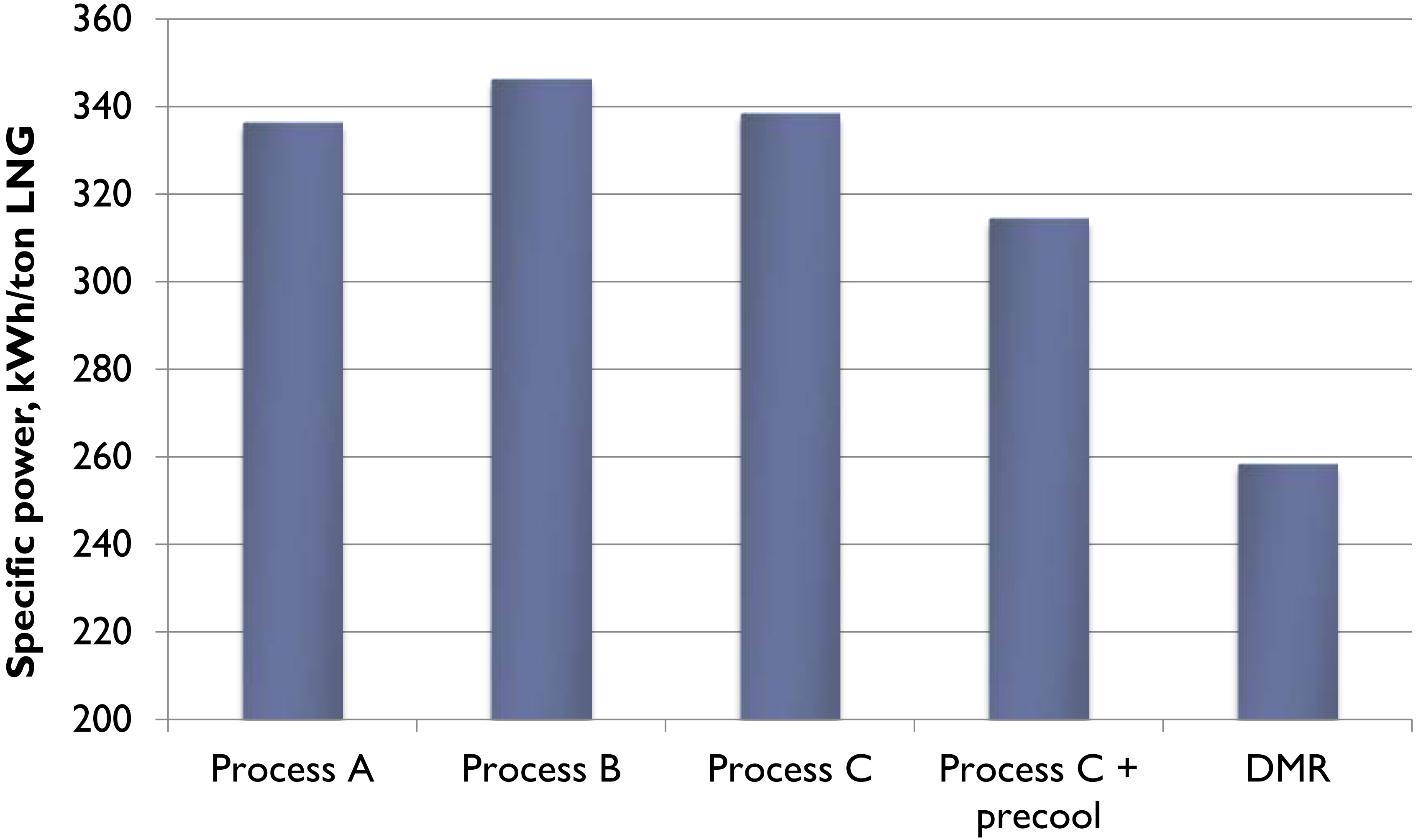
Cycle	Process A	Process B	Process C	
			Basic	CO_2 precooled
Cycle compression power, MW	35	35	35	35
Precooling power, MW	-	1.3	-	1.4
Precooling heat duty[1], MW	10	-	-	-
Total power, MW	35	36.3	35	36.4
LNG production, MTPA	0.83	0.83	0.82	0.91
Specific power, kWh/ton LNG	336	347	339	315
Compander size – the biggest in a single train, MW	9.83	13.23	11.45	8.53

Production and Efficiency Comparison

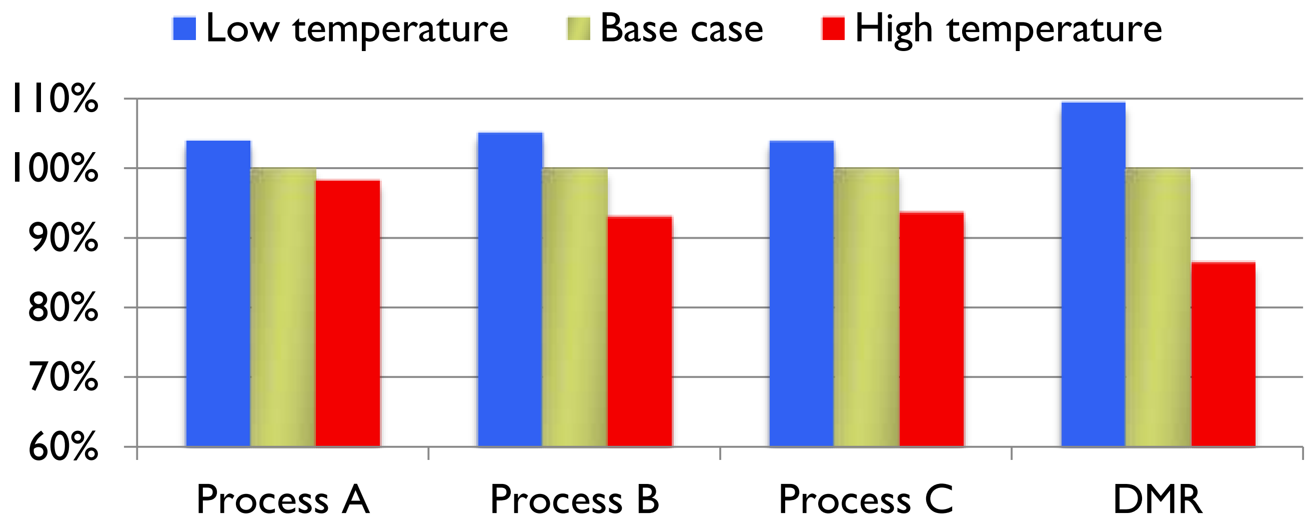


Note: At the same given power about 140 MW (4 GE LM6000) for liquefaction and precooling cycle

Specific Power Comparison

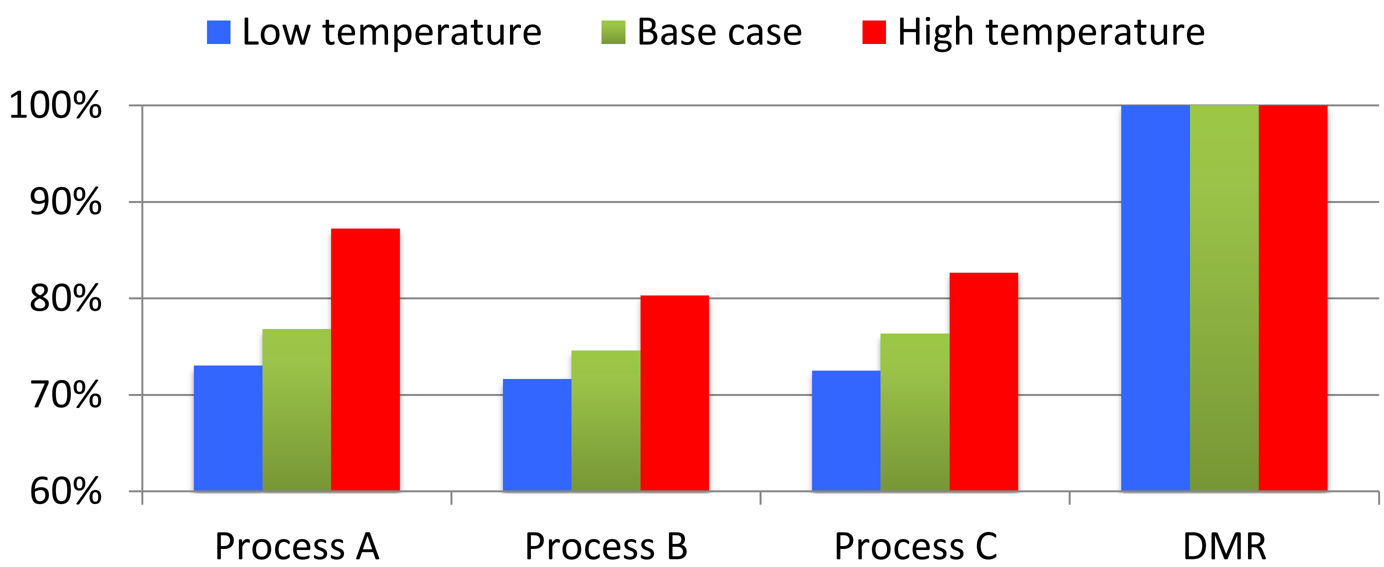


Effect of Cooling Water Temperature on Production



- All processes give higher production at lower temperature with the same given power (more efficient) since less power required for rejecting the heat into a colder sink (ambient)
- It is influencing more for DMR since the refrigerant condensation pressure depends on cooling water temperature
- The condensation of refrigerant typically rejects a large part of the energy removed from the natural gas to make LNG thus the variation of the pressure gives significant effect to the power consumption

Effect of Cooling Water Temperature on the Relative Production to the DMR



- Each case was compared to the DMR and represented as a percentage of the corresponding DMR production (100% = production for each DMR case)
- The difference in production between expanders vs DMR at the same cooling temperature is smaller at high cooling temperature and vice versa.
- High ambient temperature (e.g. tropical) reduces the advantage of the DMR over the expanders in term of efficiency

Cooling Water System

Cycle	Process A	Process B	Process C		DMR
			Without precooling	With CO_2 precooling	
Cooling water flow, m^3/h					
- For precooling cycle per train	1710	336	-	418	-
- Total per train	5,448	4,434	4,371	4,181	13,180
- Total in a 3 MTPA plant	21,792	17,736	17,484	16,724	13,180

In/Out =17 °C/30 °C

- A higher specific power of the process give a higher need for CW
- Process A is the highest as a consequence of introducing a large amount of heat into the precooling (ARS)
- Process A need a chilled water loop in addition to CW system for the precooling

Train Configuration and Equipment Count

Cycle	Process A	Process B	Process C: Without precooling	Process C: With CO_2 precooling	DMR
Number of train per 3 MTPA plant	2	4	4	4	1
The key equipment count per train:					
- Compressors	6	2	4	4	6
- Pumps	0	0	0	0	1
- Compander	6	2	2	2	-
- Heat exchanger (CWHE/PFHE)	4 (2/2)	1 (-/1)	1 (-/1)	1 (-/1)	2 (2/-)
- Separators	1	1	2	2	4
- Water cooled exchanger	8	3	6	6	6
Total equipment per train	25	9	15	15	19
Common precooling:	YES	YES	NO	YES	NO
- Compressors	-	2	-	2	-
- Separator	-	-	-	1	-
- Heat exchanger	-	1	-	1	-
- Water cooled exchanger	6	2	-	2	-
- ARS package (@ 7 MW)	8	-	-	-	-
- Chilled water pump (plus back up)	2	-	-	-	-
- HRSG unit (additional)	2	-	-	-	-
Total equipment per plant	50 + 18	36 + 5	60	60 + 6	19

Note: numbers are indicative based on the equipment units shown in the HYSYS simulation and did not consider the size and the duty of the units which may result in several units in parallel in actual plant.

Complexity

- The number of equipment units indicates complexity of the facilities.
- Process A is more complex and Process B is the simplest
- Even though the DMR has less number of equipment, the DMR is still considered more complex particularly if operational complexity when the plant start up/shut down and or dealing with feed gas condition changes are taken into account

Footprint and weight

- Footprint is a function of equipment count and dimension
- The actual volume flow indicating how large the suction piping needs to be and it may represent facility dimension
- Mass and actual volume flow rate of refrigerants into LP compressor inlets:

Cycle	Process A	Process B	Process C		DMR
			Without precooling	With CO_2 precooling	
Refrigerant flow, ton/h	1412	1062	706 (C1+N_2)	891 (C1 + N_2)	2227
Actual volume flow at compressor suction[1]):					
- in m^3/s	10.5	11.9	8.7 (C1)	6.4 (C1)	25.1
- in m^3/h	37,800	42,840	31,320 (C1)	23,040 (C1)	90,360

Niche processes is the smallest

- Weight of a topside processing facility is a function of number of equipment, thickness and the material use, including structural material
- The process with higher number of equipment and operates at higher pressure will be heavier
- Process A is considered as the heaviest and having the largest footprint

Hydrocarbon inventory

- All the expander-based processes evaluated use of a safe non-flammable refrigerant, i.e. nitrogen, or a minimum use gaseous methane as in Process C
- The expander-based processes are therefore ideal for FLNG where a small hydrocarbon inventory is preferable from a safety point of view.

Motion

- Except the basic Process C, all the expander based processes subjected to two-phase operation in their precooling system
- They are all subjected to the vessel motion to some extent.

Comparison Summary

Criteria	Process A	Process B	Process C	
			Basic	Precooled
Process efficiency	Medium	Low	Medium	**High**
Complexity/Equipment count	High	**Low**	Medium	Medium
Footprint	High	Medium	**Low**	Medium
Weight	High	**Low**	Medium	Medium
Safety	**High**	Medium	Low	Low
Sensitivity to motion	Medium	Medium	**Low**	Medium

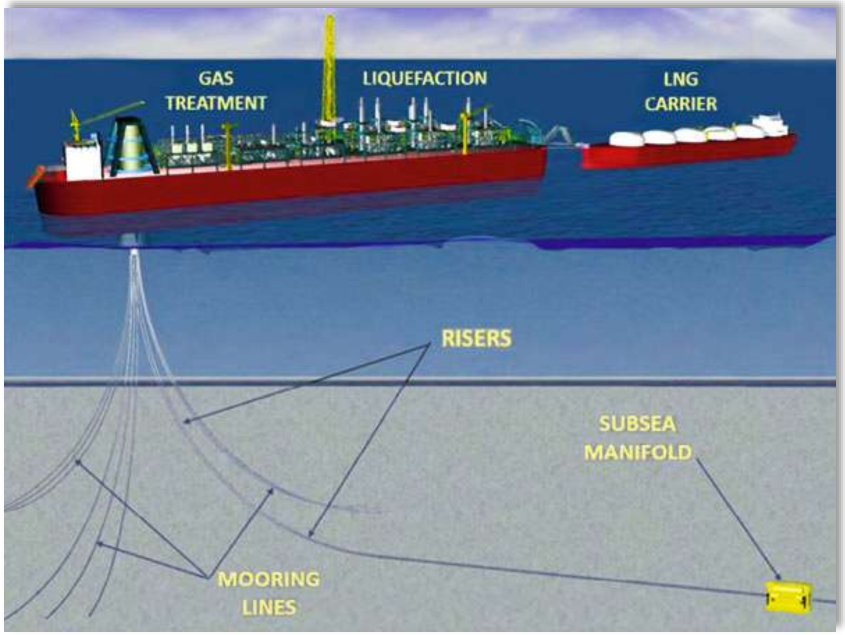
Note:

The process, which is considered the most suited for a certain criteria, is highlighted (**bold letter**)

TUSEN TAKK!!

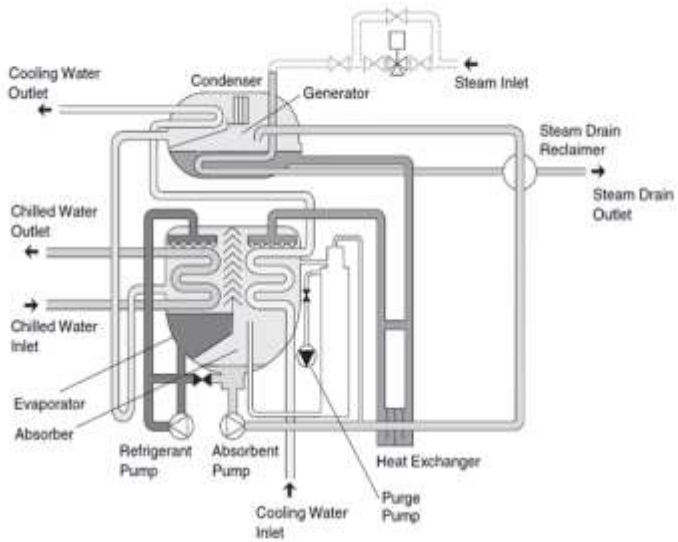
FLNG Overview

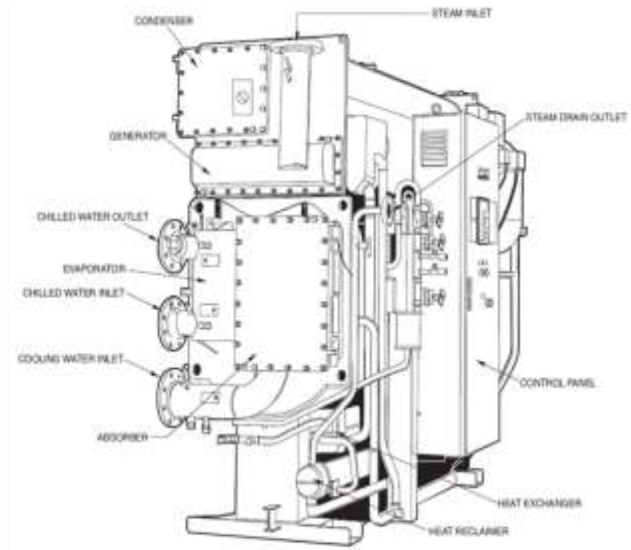
- Combining advance technology in land-based LNG and offshore FPSO
- Used for monetizing stranded offshore natural gas
- No FLNG currently exist
- Has different requirement compared to land-based LNG i.e. safety, simplicity, motion, low weight and small footprint

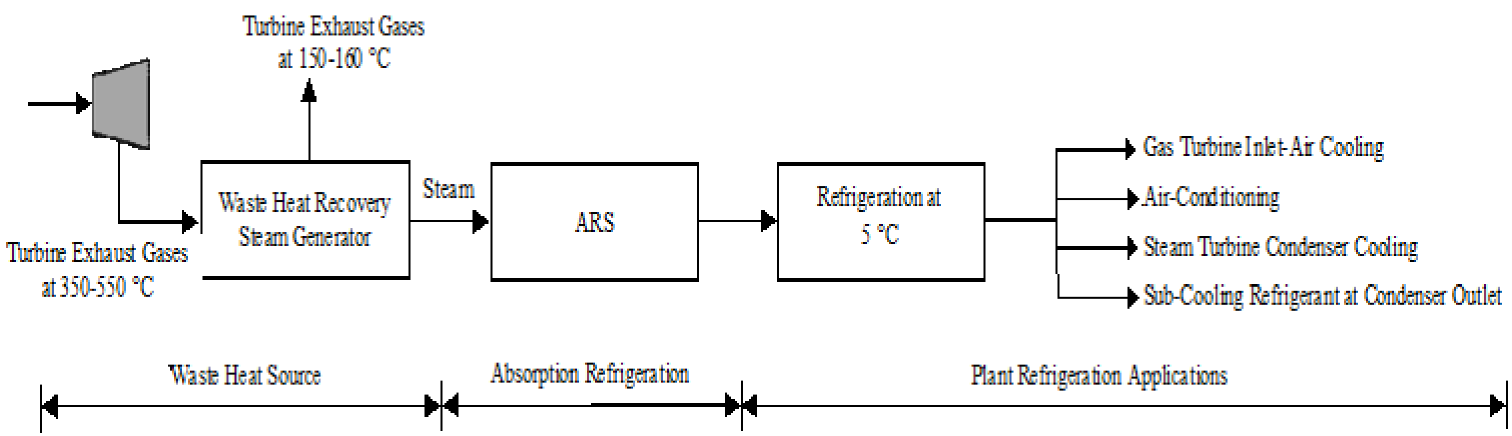


(ExxonMobile, 2013)

Absorption Refrigeration System



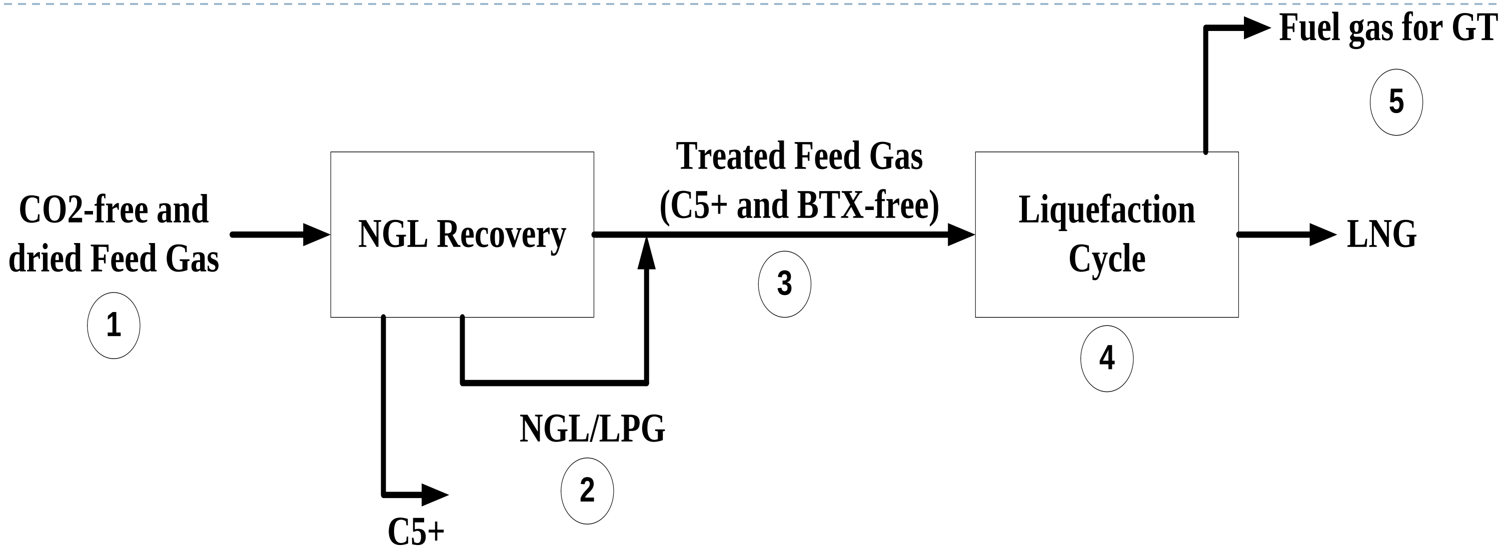




Evaluation and Comparison Basis

- The same set of conditions were applied to the proposed process schemes:
  - Simulation scope
  - Feed gas condition and composition
  - Site conditions (air/cooling water temperature)
  - Drivers (gas turbines)
  - Heat exchanger sizing (min temperature approach)
  - Component efficiency (compressor polytrophic eff.)
  - LNG product condition (i.e. end flash vapor quantity)

Simulation scope



1) The pretreating units, i.e. CO_2 removal, dehydration, mercury removal, are out of the scope of the simulation in this study. The feed gas is processed, sweet and dry natural gas coming from the pretreating units at the upstream.
2) NGL/LPG extracted in NGL recovery unit (modeled as turboexpander unit) is fully re-injected into LNG feed stream (assumes that no LPG production and no make up refrigerant required). The NGL/LPG extraction is only for removing BTX components in the feed gas stream. Those aromatic components are assumed to leave the LNG feed stream in the condensate (C5+) product.
3) Lean gas leaving the NGL recovery unit enters into the liquefaction circuit at the same temperature/pressure condition as when it enters the NGL recovery unit (it is recompressed by the booster and cooled by cooling water).
4) The proposed expander-based liquefaction process schemes and the APCI DMR were simulated in this thesis for analysis and comparison.
5) The LNG condition is set to provide constant end flash vapor quantity. The nitrogen content is assumed to be moderate and does not necessitate the implementation of a dedicated nitrogen rejection unit.

Feed Gas

Condition

Properties	Feed Gas Stream
Pressure	60 bar abs
Temperature	22 °C

Composition

Component	Composition (in %-mole)
Nitrogen	1.00
Methane	91.00
Ethane	4.90
Propane	1.70
i-Butane	0.35
n-Butane	0.40
i-Pentane	0.15
n-Pentane	0.15
n-Hexane	0.13
n-Heptane	0.10
n-Octane	0.04
n-Nonane	0.01
n-Decane	0.01
CO_2	0.00
H_2O	0.00
Benzene	0.03
Toluene	0.02
m-Xylene	0.01

Cooling Water Temperature

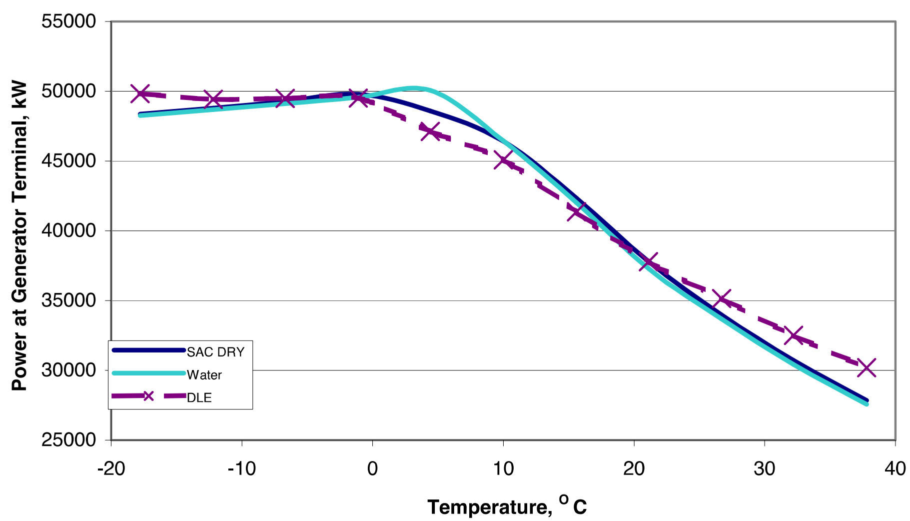
- The temperature was assumed at 17 °C and it is a closed-circuit cooled by sea water at 14 °C.
- A 5 °C cooling water cooled exchanger temperature approach was assumed meaning that the compressed refrigerant and feed gas was cooled to 22 °C by using the cooling water.

The Driver (Gas Turbine)

- Gas turbines (GT) used as mechanical driver of main refrigerant compressors and electrical power generation considered are General Electric (GE) LM6000 models

ISO rated power (MW)	TIT (°C)	Exhaust (°C)	Air flow (kg/s)	Pressure ratio	Efficiency (%LHV)
42.9	1260	456	124	30	41.7

- Air temperature was assumed at 27 oC



Heat Exchanger Sizing

- A minimum temperature approach of 3 °C in the cryogenic heat exchangers was assumed
- The refrigerant pressure of expander-based processes was limited to 85 bar

Component Efficiency

- Compressors used in the simulation were assumed centrifugal type that has moderate polytropic efficiency of 78%
- The compander polytropic efficiency was based on GE (Byrne and Mariotti 2010), it was assumed of 73% for the compressor and 83% for the expander

Production and Product Quality

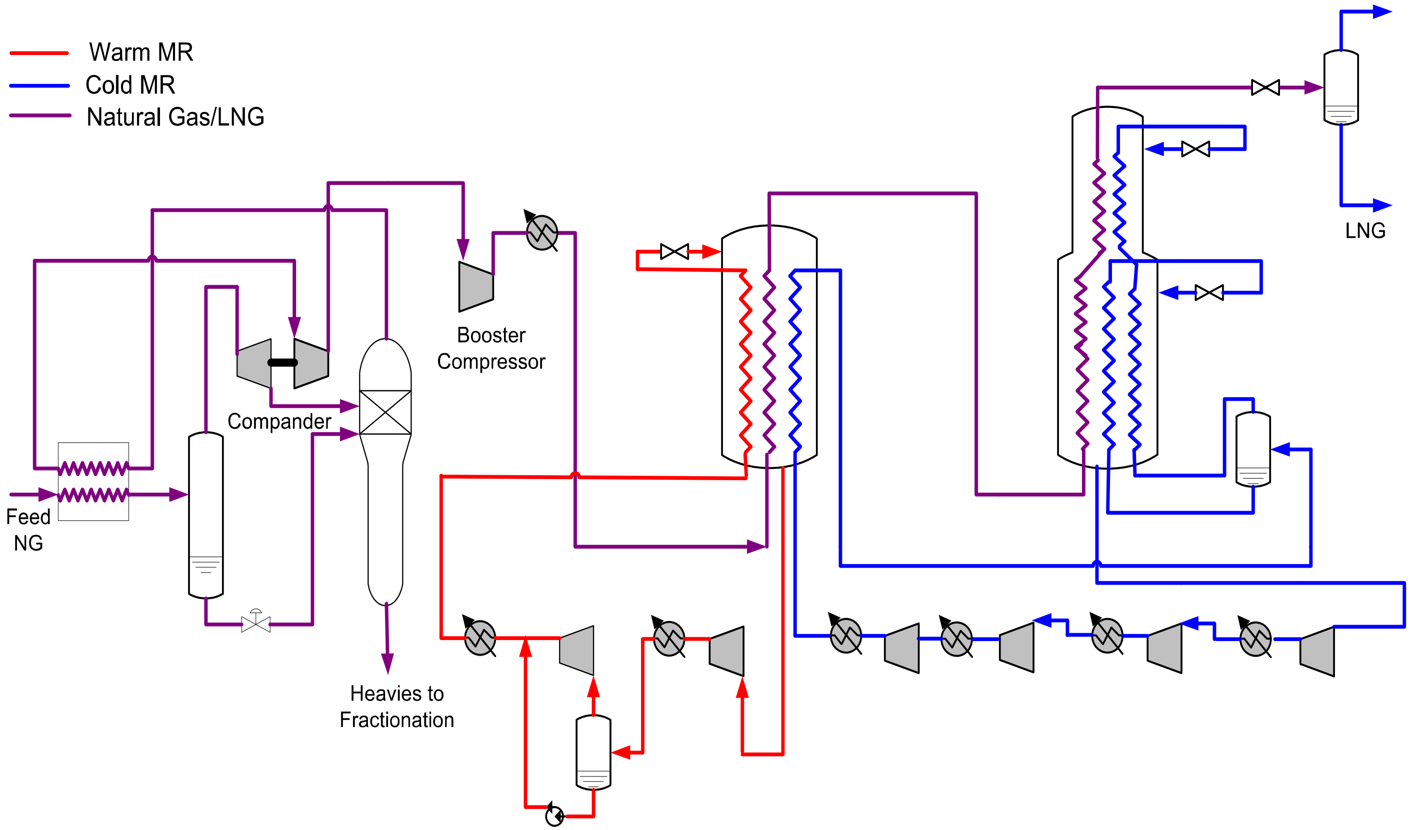
- The intended LNG production is about 3 MTPA with 330 days in a year for the plant availability.
- LNG product was assumed at -149 °C at the exit of cryogenic heat exchanger and was expanded in end flash column to pressure of 1.38 bar before going to the storage tanks.
- By this condition, the end flash vapor generated is about 8%-mass of LNG product and nitrogen content is within the LNG specification.
- It was assumed that the end flash gas from a single expander process train covers fuel needed for a gas turbine
- LNG Quality

Parameters	LNG
Nitrogen, %-mole	< 1
C5+, %-mole	< 0.1
BTX, ppm	10

HYSYS Model Development

- Methodology
  1. The literature review on the patents and publications of each proposed process scheme was used as basis information for model development
  2. Peng-Robinson EOS was used for calculation of thermodynamic properties
  3. A steady state mode calculation
  4. Optimizing by varying refrigerant flow rate to obtain the selected assumption of minimum approach in LNG heat exchangers.
  5. The production was determined based on the given gas turbine power as a mechanical driver for the refrigerant compressors.
  6. The key parameters recorded after optimization was LNG production, UA value of LNG heat exchangers, refrigerant flow rates and specific power.

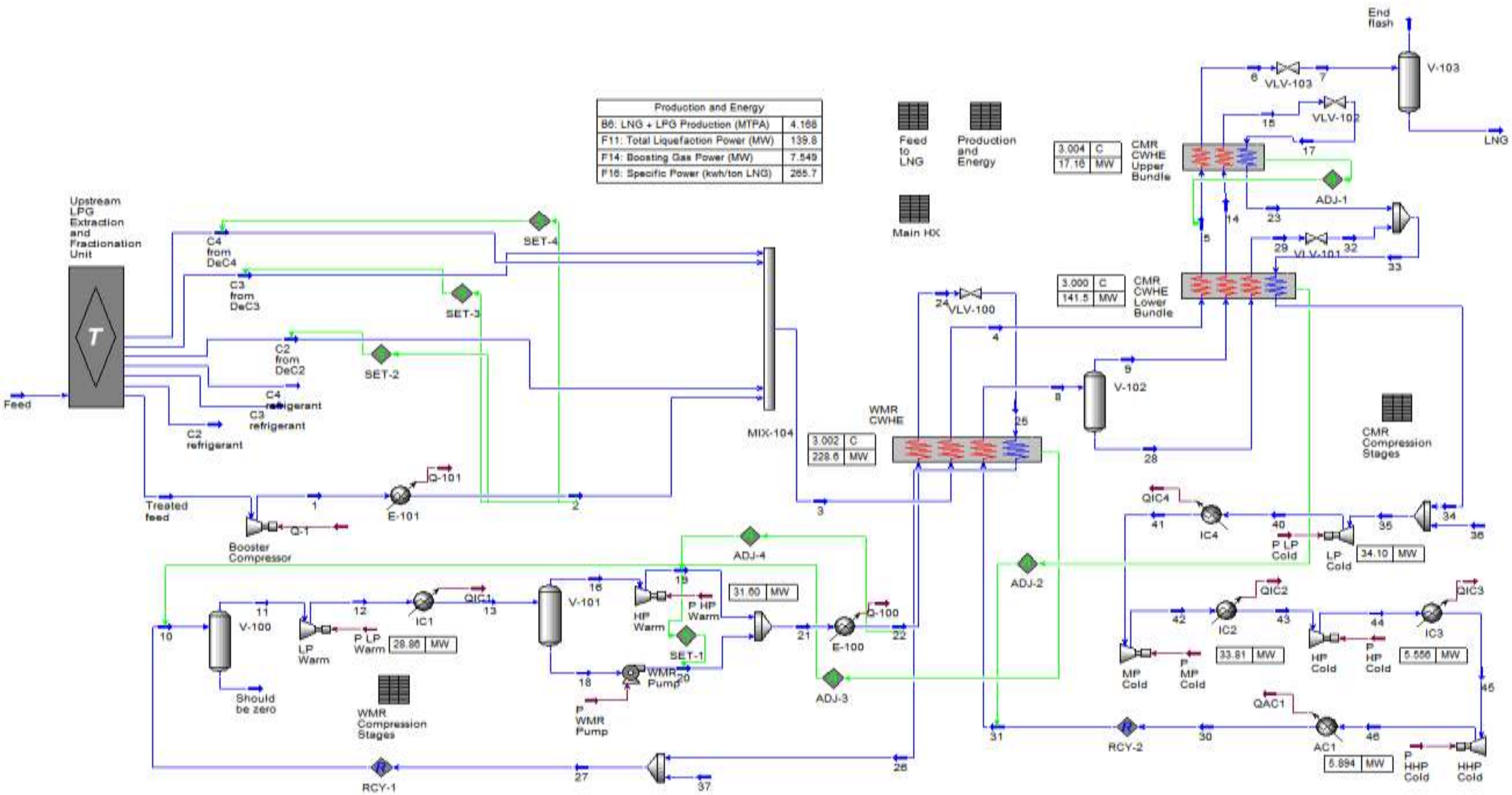
DMR process as basis for comparison



- This is a modification on an established HYSYS model from Statoil, which was adopted by the author during his previous work on the specialization project

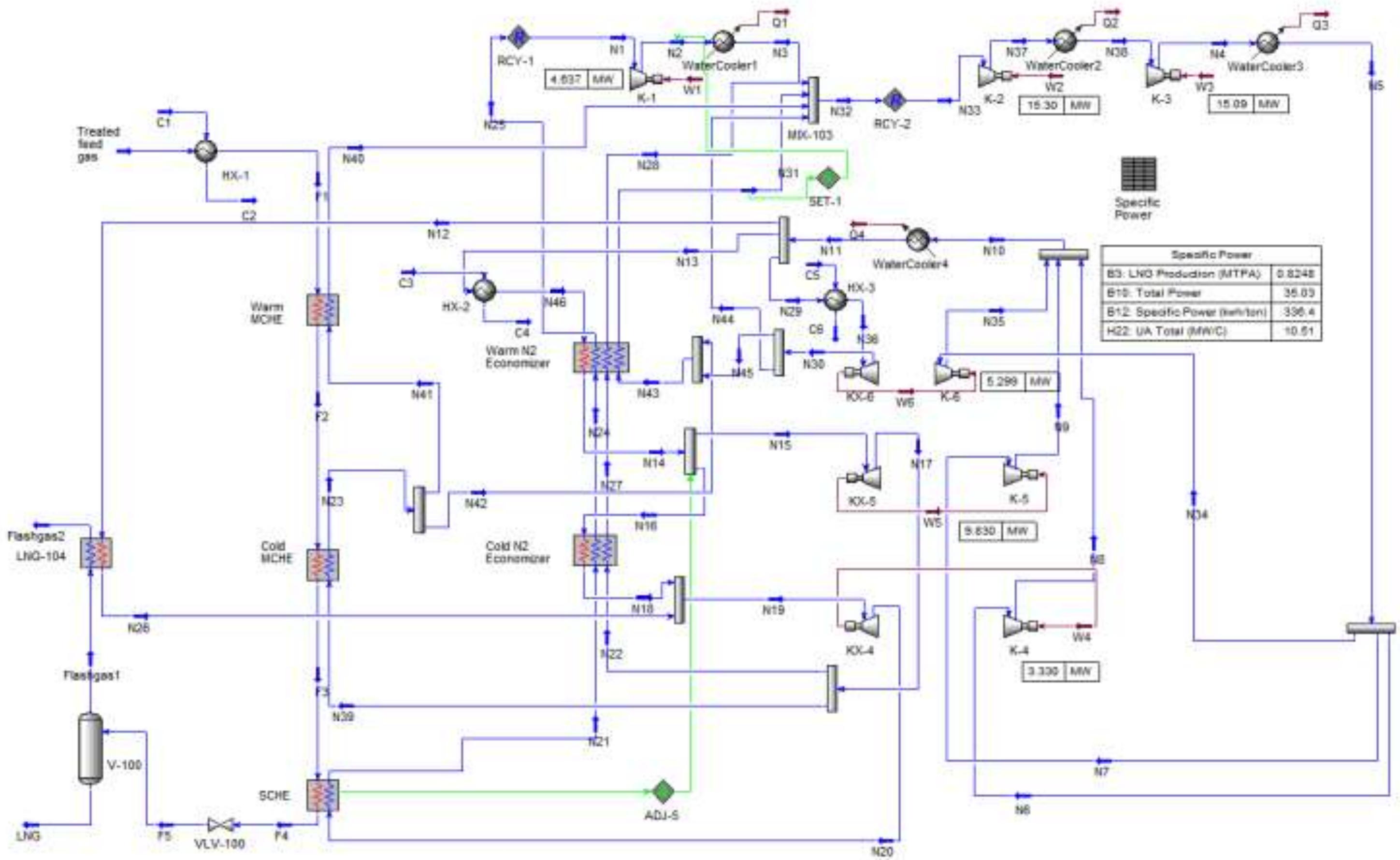
HYSYS Model Development

- DMR Process



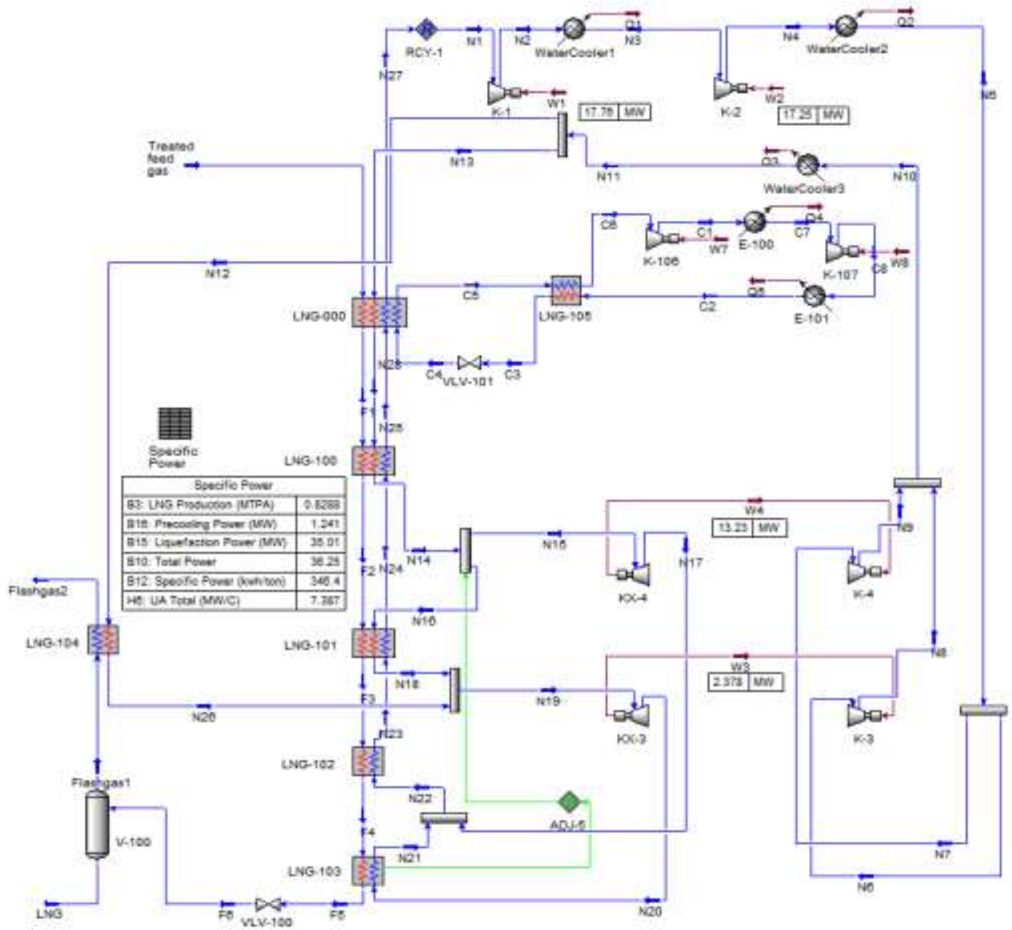
HYSYS Model Development

- Process A



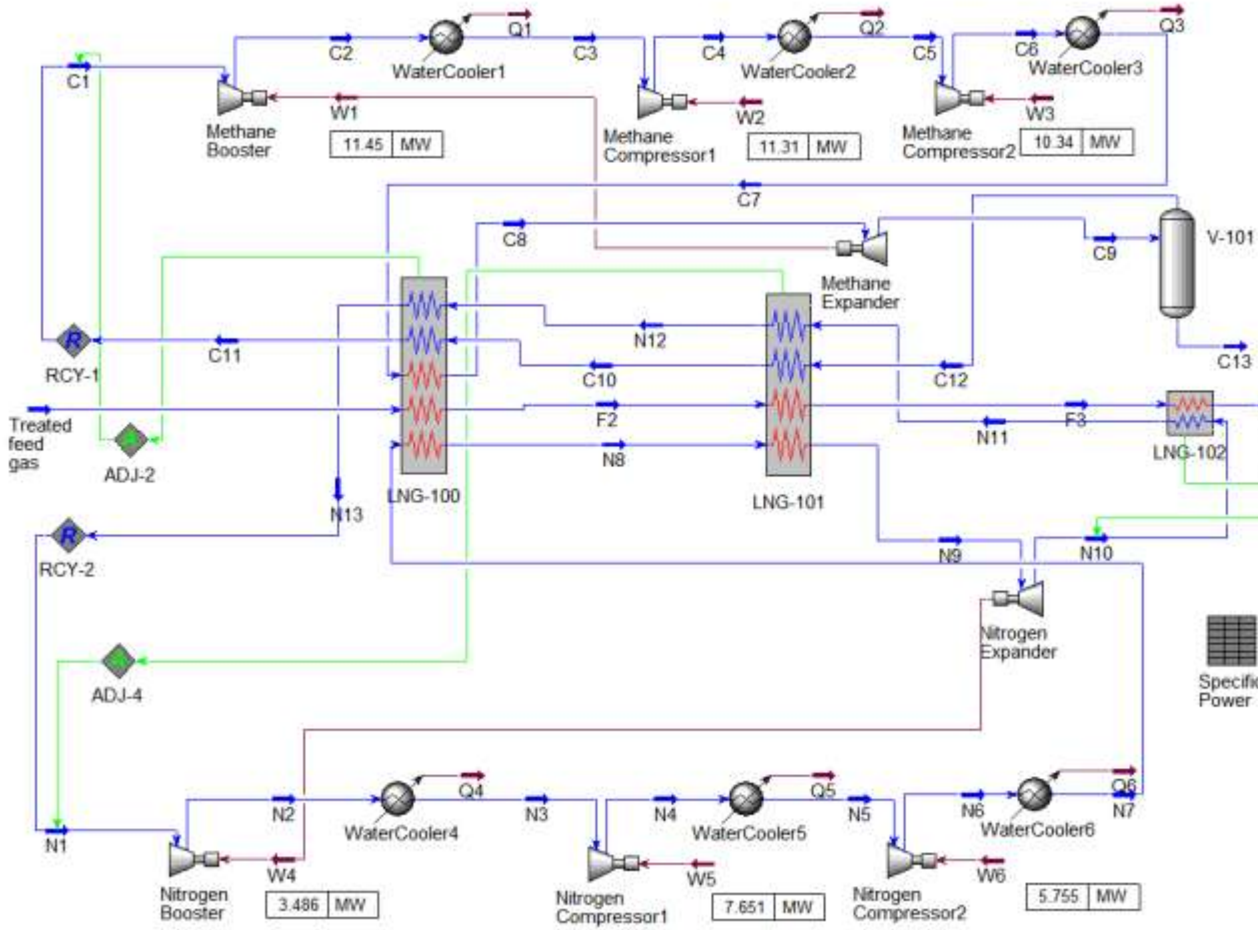
HYSYS Model Development

- Process B



HYSYS Model Development

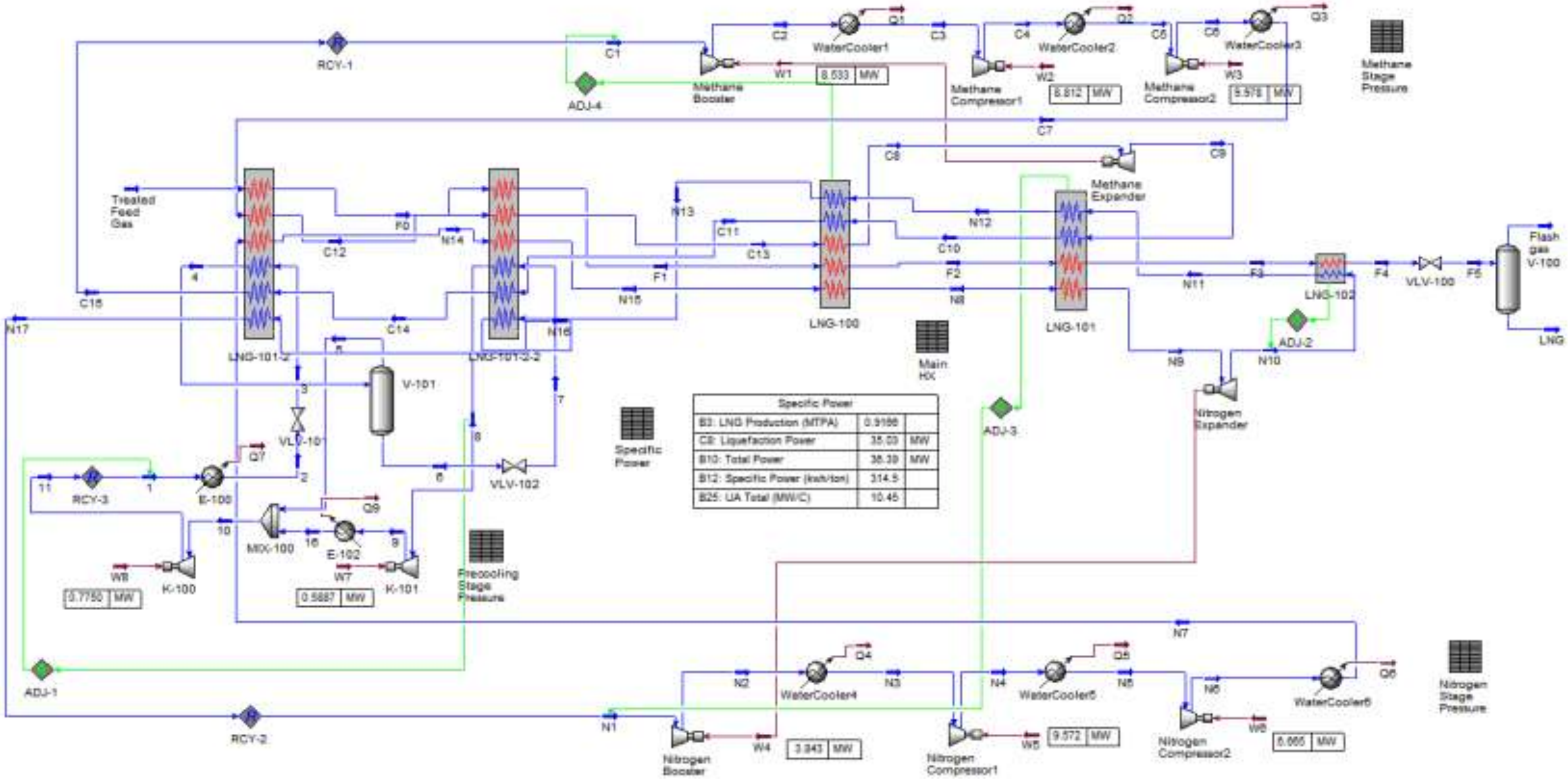
- Process C



Specific Power		
B3: LNG Production (MTPA)	0.8203	
B10: Total Power	35.06	MW
B12: Specific Power (kwh/ton)	338.5	
E14: UA Total (MW/C)	6.995	

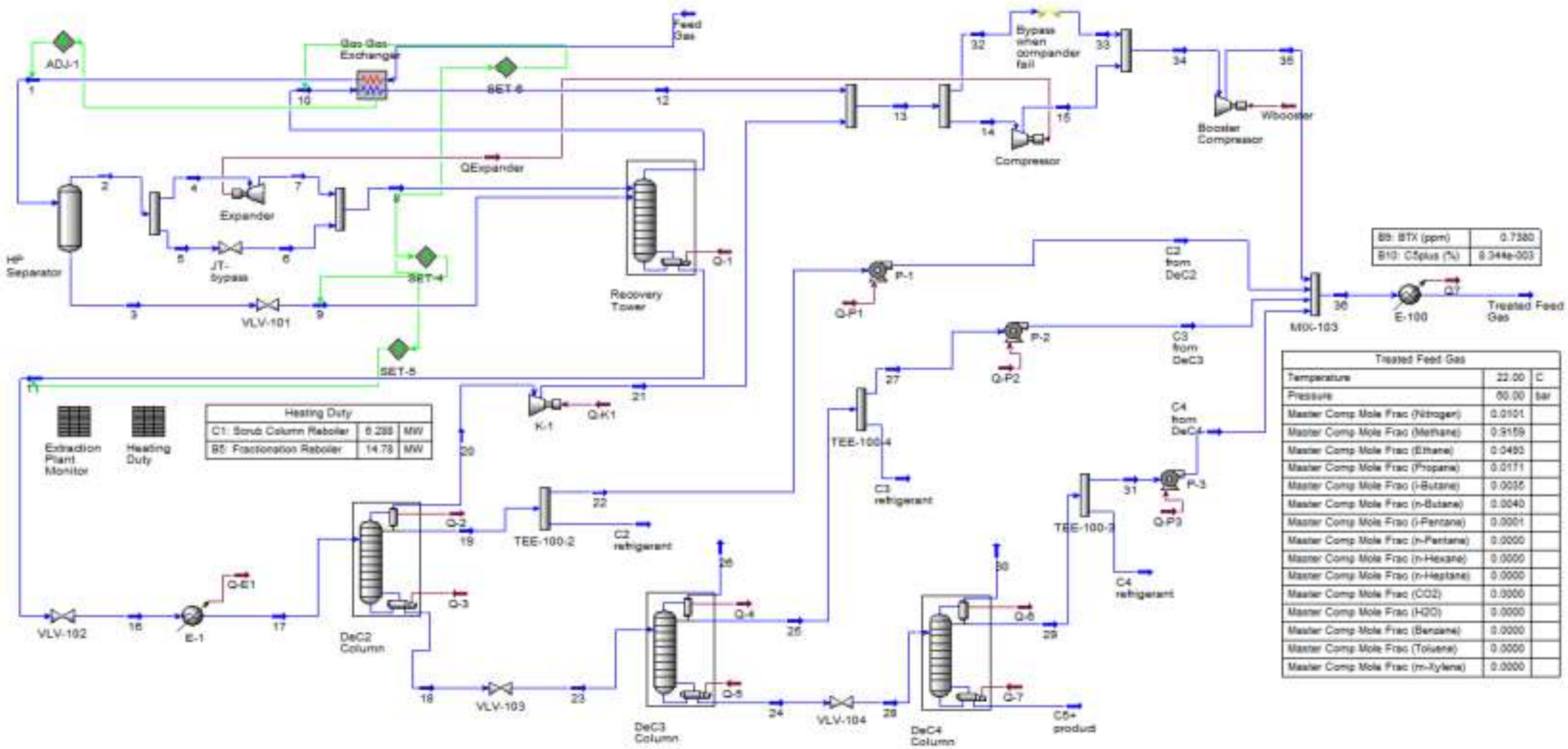
HYSYS Model Development

- Process C + precooling



HYSYS Model Development

- NGL Extraction Unit



B9: BTX (ppm)	0.7380	
B10: C5plus (%)	8.344e-003	

Heating Duty		
C1: Scrub Column Reboiler	6.288	MW
B5: Fractionation Reboiler	14.78	MW

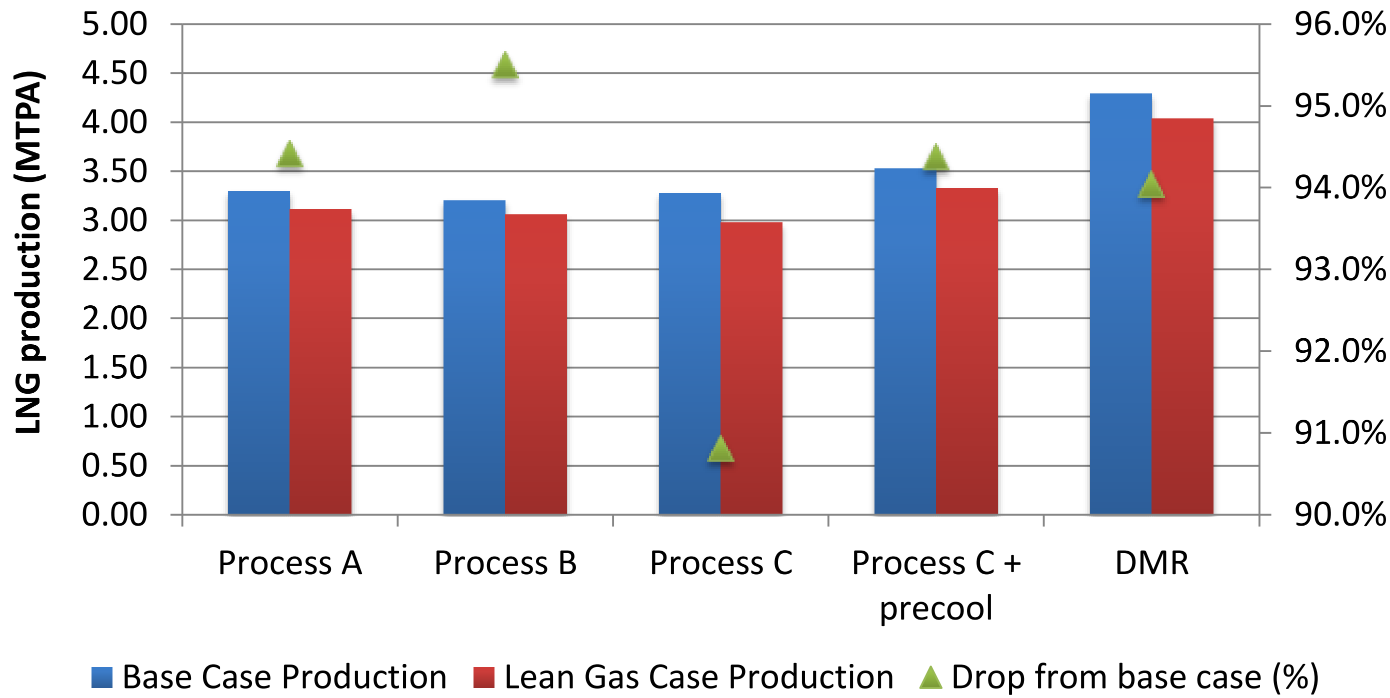
Treated Feed Gas		
Temperature	22.00	C
Pressure	60.00	bar
Master Comp Mole Frac (Nitrogen)	0.0101	
Master Comp Mole Frac (Methane)	0.9159	
Master Comp Mole Frac (Ethane)	0.0493	
Master Comp Mole Frac (Propane)	0.0171	
Master Comp Mole Frac (i-Butane)	0.0036	
Master Comp Mole Frac (n-Butane)	0.0040	
Master Comp Mole Frac (i-Pentane)	0.0001	
Master Comp Mole Frac (n-Pentane)	0.0000	
Master Comp Mole Frac (n-Hexane)	0.0000	
Master Comp Mole Frac (n-Heptane)	0.0000	
Master Comp Mole Frac (CO2)	0.0000	
Master Comp Mole Frac (H2O)	0.0000	
Master Comp Mole Frac (Benzene)	0.0000	
Master Comp Mole Frac (Toluene)	0.0000	
Master Comp Mole Frac (m-Xylene)	0.0000	

Sensitivity Analysis

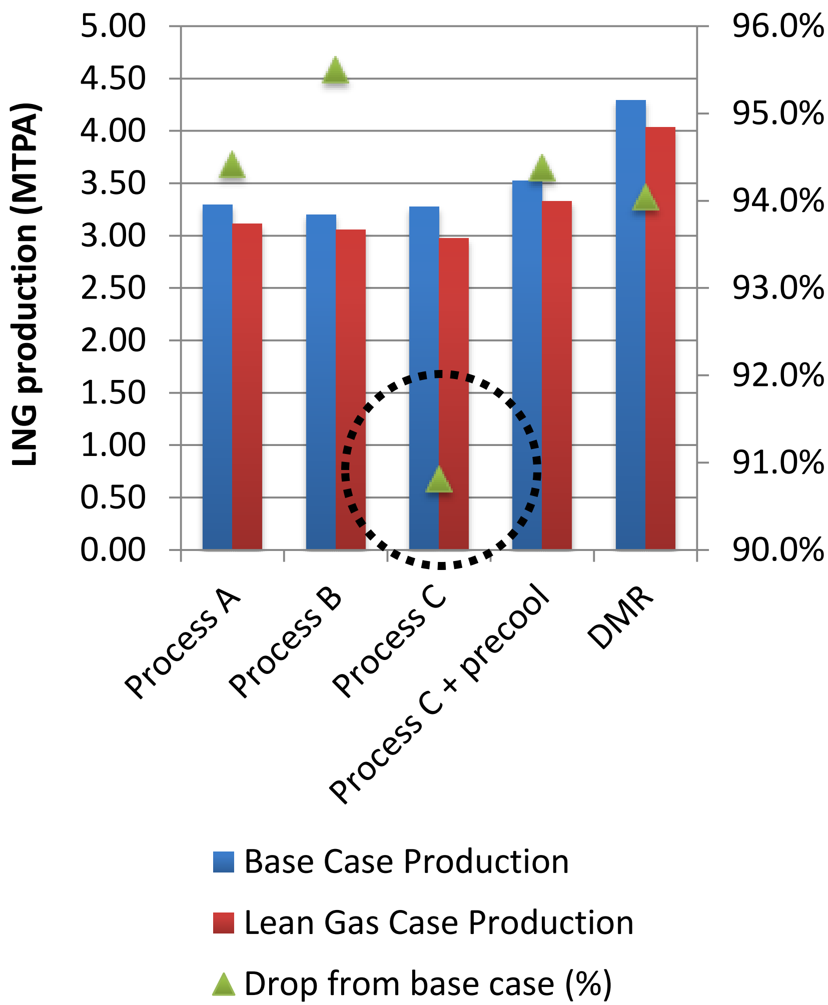
- Effect of feed gas composition
- Effect of treated feed gas pressure
- Effect of refrigerant pressure
- Effect of gas turbine intake air cooling
- Effect of cooling water temperature (heat rejection)

Effect of Feed Gas Composition

- A study case with lean feed gas was performed
- Production of all processes drops at the same given power (lower efficiency)
- The lean gas has a lower condensation temperature i.e. larger temperature lift thus higher work requirement



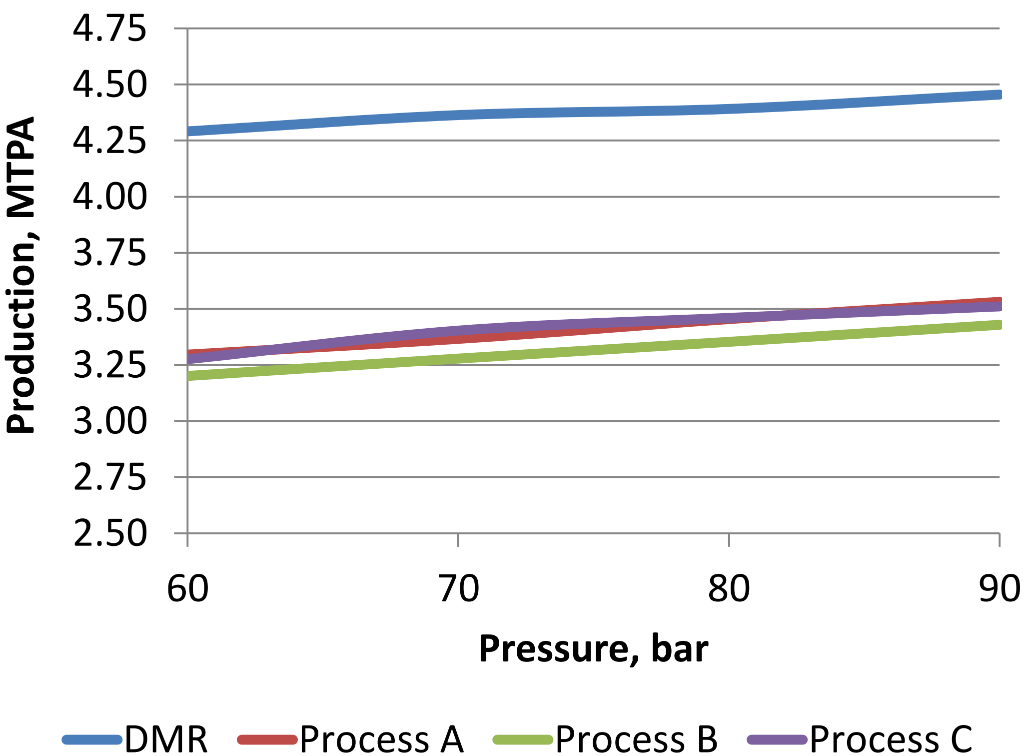
Effect of Feed Gas Composition (cont.)

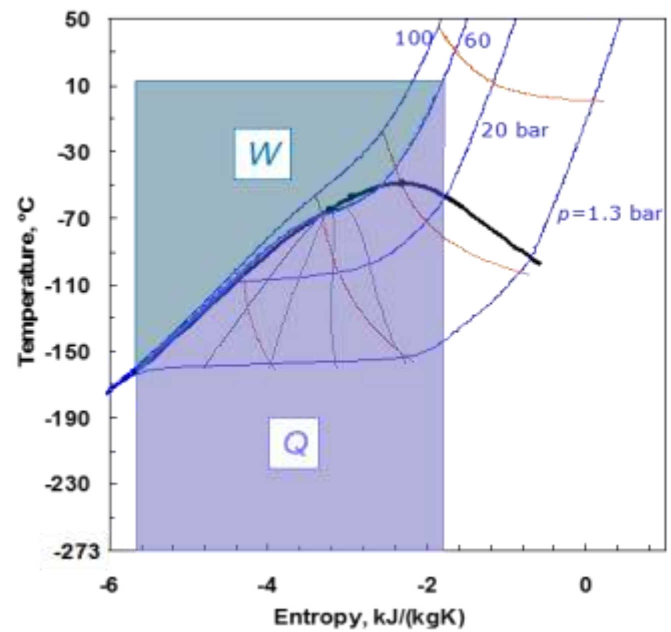


- The process C without precooling tends to suffer more than the other processes (production drops about 9% from its base case)
- The production drops is due to the feed gas is also the basis for refrigerant in one of the circuits i.e. the methane rich refrigerant circuit.
- Using a leaner feed gas for the refrigerant increases the specific compression power since the gas has a higher compressibility (z) and a lower molecular weight (M)

Effect of treated feed gas pressure

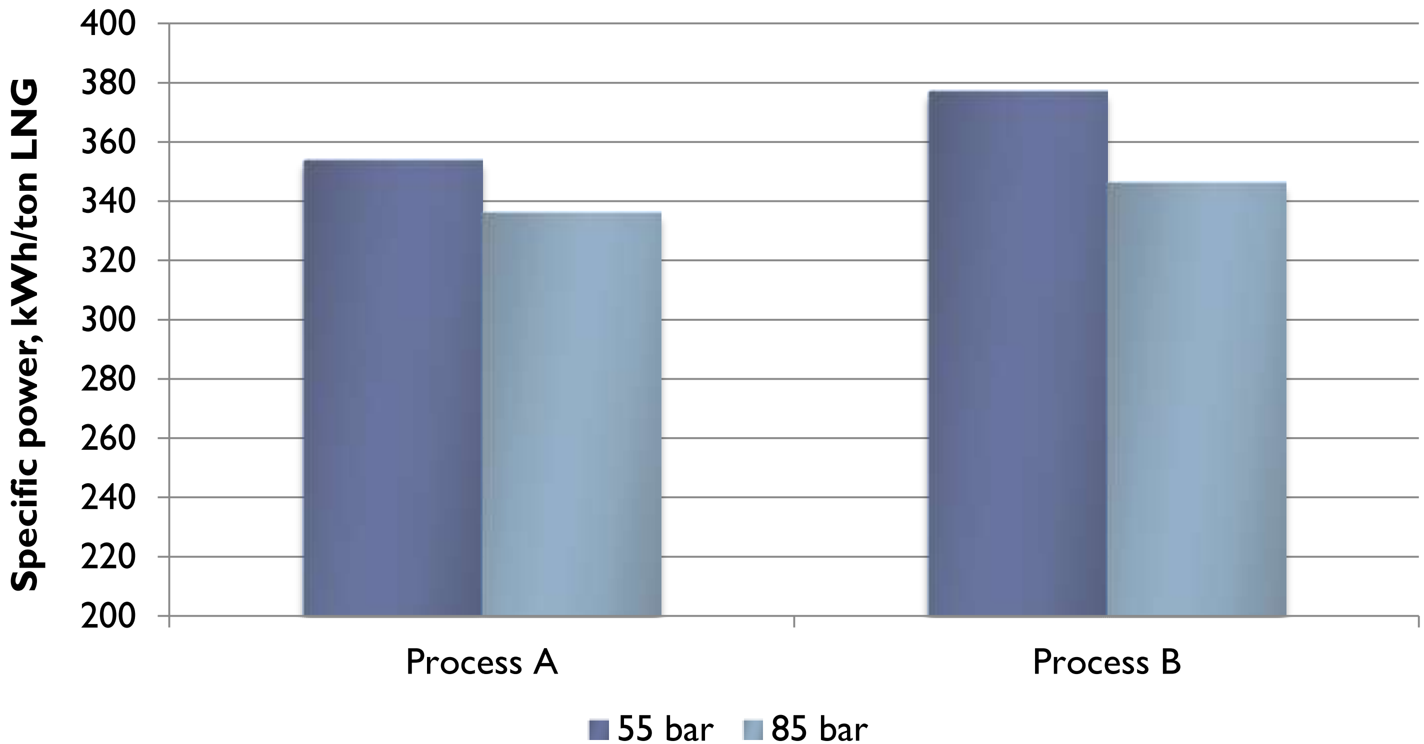
- Pressure was varied to the limit that the main HX still withstand
- For all processes, higher pressure increases the efficiency
- How the pressure effects to the efficiency in a T-S diagram
- At higher pressure, the min work and heat load to liquefy the gas is reduced





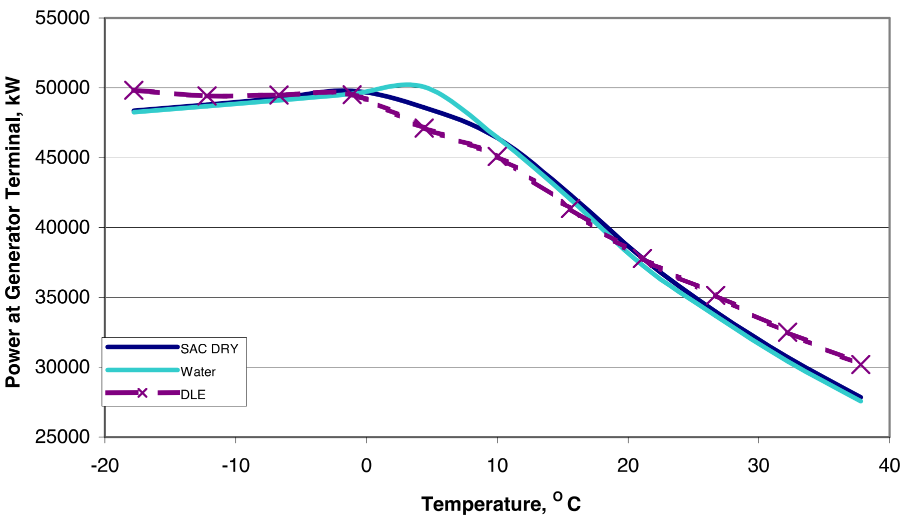
Effect of refrigerant pressure

- The refrigerant pressure was limited up to a practical limit of the cryogenic heat exchangers, which is in this study limited to 85 bar



Effect of gas turbine intake air cooling

- Lower air temperature higher gas turbine power output

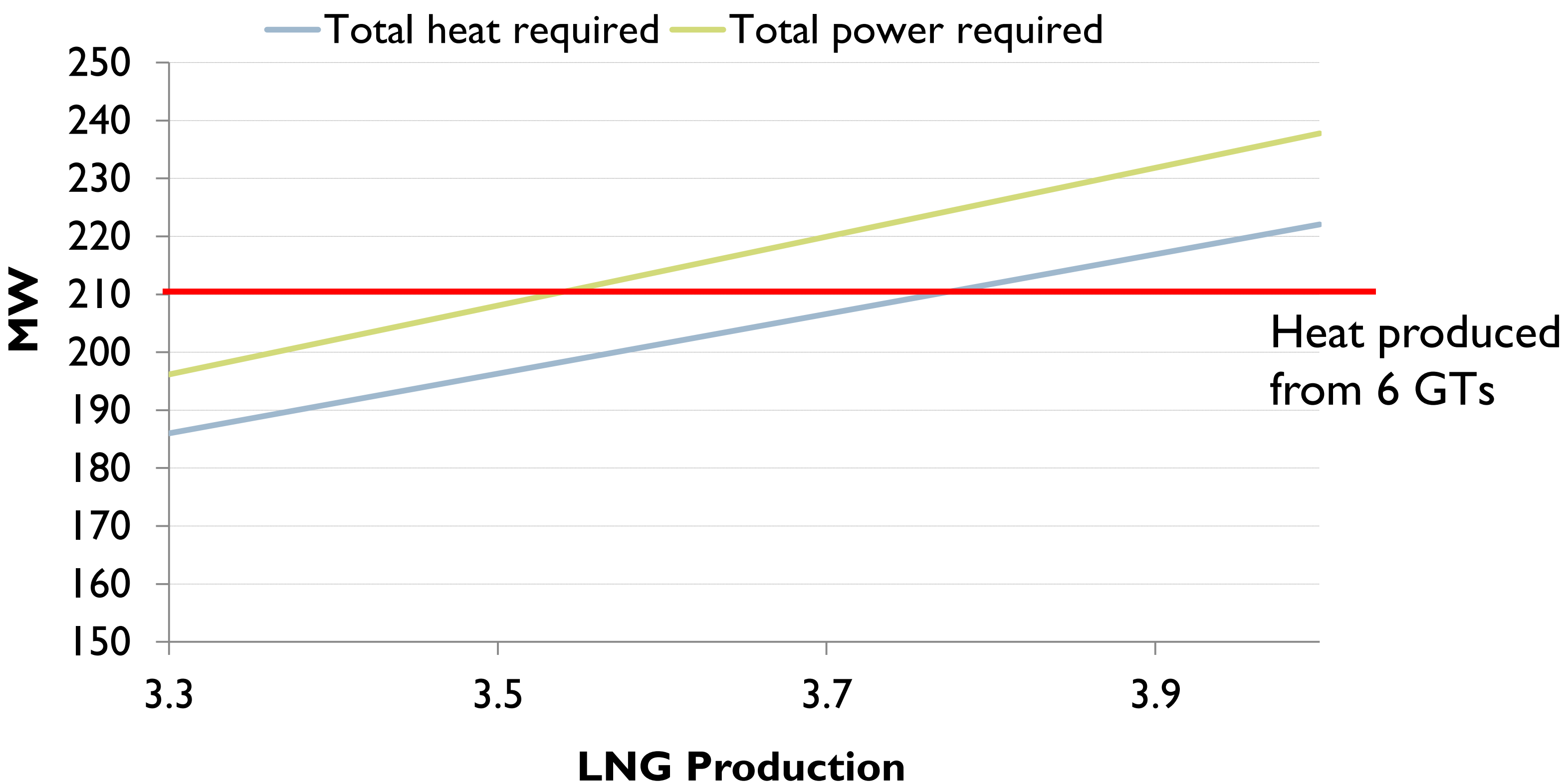


- The production capacity for all processes potentially increases over 15% when the precooling system is utilized to cool gas turbine air intake.

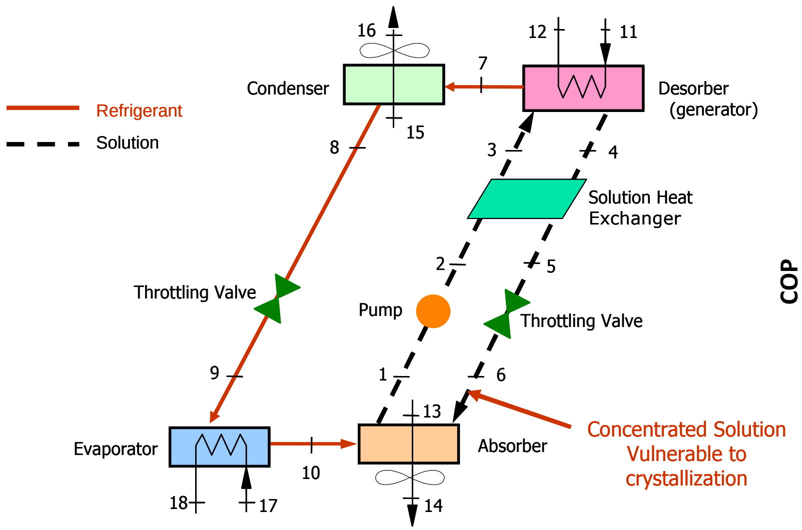
Cycle	Process A	Process B	Process C+ CO_2 precooling
Total power (MW)	4 x 45	4 x 45[1]	4 x 45[1]
LNG production (MTPA)	4.2[2]	4.1[3]	4.5
% increased in production	97%	95%	105%
Compander size (MW)	12.6	16.3	10.5

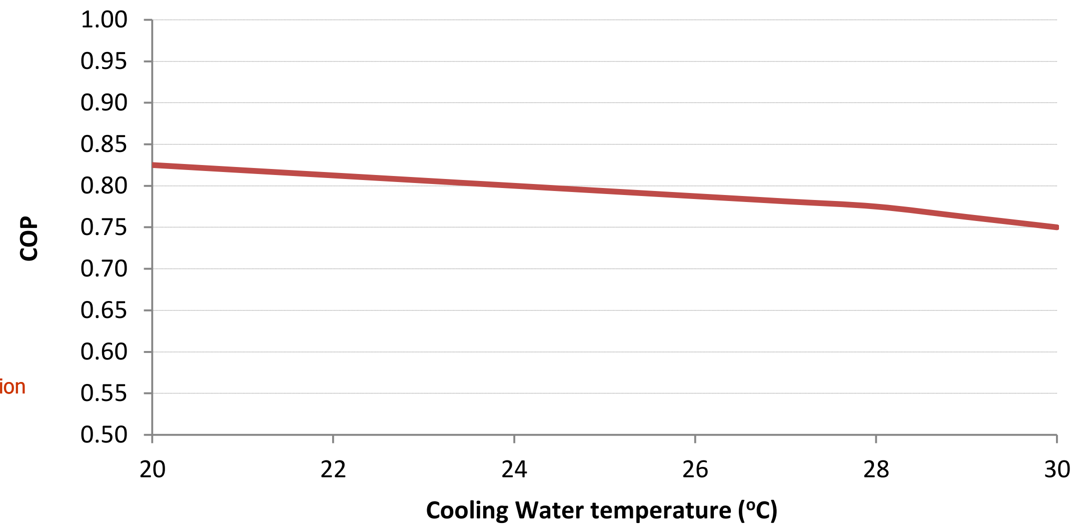
Integration into Energy System

- Power and Heat Balance (correspond to the Process A only since it requires larger amount of heat to drive its precooling)



Effect of Cooling Water Temperature to the ARS system used as Precooling in Process A





- The COP of the system is lower at higher cooling water temperature
- To provide the same amount of cooling duty from this system, increased in heat supply to the system is required. There will be not enough waste heat to provide that requirement.
- And at higher cooling water temperature, crystallization of the solution in the ARS is more likely occurs

Effect of Cooling Water Temperature to the Process A Heat Balance