Thermal coupling opportunities for floating natural gas liquefaction plants

I.J. Halvorsen [a,d,*], I. Dejanović [b], Ž. Olujić [c], S. Skogestad [d]

[a] SINTEF Digital, Mathematics and Cybernetics, Trondheim, Norway
[b] University of Zagreb, Faculty of Chemical Engineering and Technology, Zagreb, Croatia
[c] Retired, The Netherlands
[d] Norwegian University of Science and Technology (NTNU), Trondheim, Norway

ABSTRACT

In order to meet current carbon dioxide emissions reduction challenges, natural gas processing and refining industries have to find new ways to minimize energy requirements. Building on foundations laid down in a preceding effort, this paper shows that this could be achieved in a cost-effective way for the distillation section in natural gas liquids fractionation plants. A conventional demethanizer column, combined with either a thermally coupled direct sequence of deethanizer and propane–butane recovery columns or a dividing wall column, produces significant savings in capital and hot utilities costs as compared to conventional distillation with a direct sequence, without any temperature penalty on cold utilities side. The choice between available options will largely depend on important process considerations that may differ for offshore and onshore plants and specific site requirements.

© 2019 Published by Elsevier B.V. on behalf of Institution of Chemical Engineers.

1. Introduction

Modern floating and on-land natural gas liquefaction plants employ direct sequences of three or more distillation columns to recover C1 to C5+ hydrocarbons. In a recent paper addressing potential for energy conservation in the fractionation complex of a floating natural gas liquefaction plant (Halvorsen et al., 2016), it is shown that replacing conventional deethanizer and depropanizer columns by a three-product dividing wall column (DWC) could lead to significant overall energy saving as well as substantial weight and footprint benefits. However, this was achieved at the expense of an increased cold utility demand with respect to that of a conventional deethanizer column in a direct three-column sequence. Indeed, being a dominating factor, an increase in refrigeration costs may render a standard DWC industrially unviable in this and similar applications.

The present paper explores means and configurations that could help to overcome this refrigeration-demand related burden and open the door for implementing thermal coupling and DWC technologies in natural gas liquids (NGL) fractionation plants in a cost-effective way, both as a new design and as a retrofit option. The latter is

particularly interesting because it provides natural gas processing and refining industries with an opportunity to meet legislation imposed energy and carbon dioxide reduction challenges in existing NGL fractionation plants.

2. Previous work

A detailed description of process requirements for a natural gas liquids (NGL) fractionation complex within a state-of-the-art floating natural gas liquefaction plant can be found elsewhere (Halvorsen et al., 2016). This includes a real design case with feed and product specifications formulated to serve as a basis for the design of a planned floating natural gas liquefaction plant for exploitation of a marginal field. An NGL-rich feed is available at 34 bar and 54 °C, and is to be separated into four fractions: a methane rich stream (C1), an ethane rich stream (C2), a propane and butane rich stream (C3 + C4), and a pentanes and heavier stream (C5+), according to following product specifications. The top product of the deeth-

Fig. 2 – Conventional demethanizer combined with full DWC.

Nomenclature

B	Bottoms flow rate, kmol/h
D	Distillate flow rate, kmol/h
D/F	Distillate to feed ratio
L/D	Reflux ratio
L/F	Liquid to feed ratio
L	Liquid (reflux) flow rate, kmol/h
V	Vapour (boil-up) flow rate, kmol/h
V/B	Boil-up ratio
V/F	Vapour to feed ratio

Abbreviations

C1	Methane rich product stream
C2	Ethane rich product stream
C2+	Demethanizer bottom product, feed to deethanizer
C3 + C4	Propane and butane rich product stream
C5+	Pentane and heavier components rich product stream
DS-DWC	Direct sequence DWC
DWC	Dividing wall column
NGL	Natural gas liquids
TCDS	Thermally coupled direct sequence
TCSR	Thermally coupled side-rectifier arrangement
SR-DWC	Side rectifier DWC

pentanes, heavier alkanes (C6–C10) and aromatics (benzene to *m*-xylene).

As described in Halvorsen et al. (2016), to arrive at a feasible energy saving option, the thermal coupling should consider the second and third column only, leaving the demethanizer column as it is. Among a number of options considered and evaluated for the remaining C2+ separation, a conventional three-product DWC, operated at 8 bar, appeared to be the most promising. This arrangement, with a centrally placed partition wall, called "full-DWC" in the further text, is shown schematically in Fig. 2. The overheads vapour enters a partial condenser that liquefies only the fraction of vapour required as reflux. In other words, the top product (C2) is a vapour stream. Also the side stream, i.e. the (C3 + C4) fraction, which is three times larger than the ethane-rich stream, was drawn off as vapour. This effectively reduces the vapour load above the side product draw-off stage. The remaining vapour is mixed above the upper end of the partition wall with that coming from the prefractionator; thus the condenser operating at 8 bar and −39 °C receives more vapour, and more cold utilities are needed to liquefy the excessive vapour, i.e. −265 kW compared to −164 kW in case of a conventional deethanizer column with a condenser operating at 17 bar and −27 °C.

This additional refrigeration demand is a real burden that outweighs reboiler heat duty saving as well as all other potential benefits of implementation of a conventional full-DWC. This burden needs to be overcome to make thermal coupling in general, and a DWC in particular, to be an industrially viable option for natural gas processing plants.

anizer (C2) should contain more than 90 mol% ethane and less than 2 mol% propane, and the total amount of pentanes in the top product of the (C3 + C4) column should not exceed 0.5 mol%.

The common approach in this case is to utilize a direct three-column sequence (Fig. 1), and in a previous detailed simulation study the operating absolute pressure at the top (condenser) of the demethanizer was 34 bar, that of deethanizer column was 17 bar, and that of propane–butane column was 7 bar. The corresponding condenser temperatures are −91 °C, −27 °C, and +25 °C, respectively, which implies that refrigeration is required in the condensers of the first two columns.

The feed composition, the compositions of the product streams as estimated using detailed simulation, and related processing conditions are summarized in Table 1. The stream denoted "C2+" represents the bottom stream leaving the demethanizer column, which upon cooling to the required level enters as a feed to the deethanizer column. The product stream (C5+) leaving the bottom of (C3 + C4) column contains
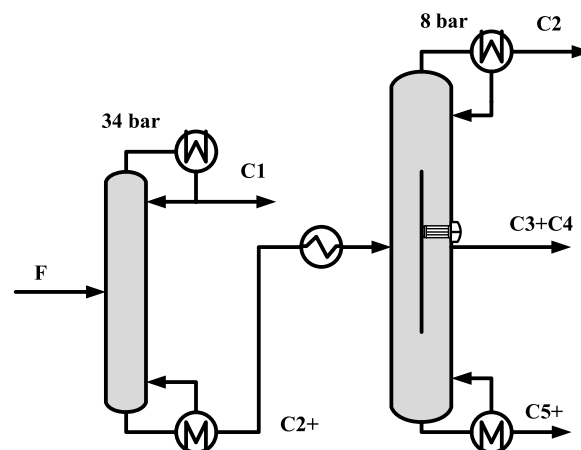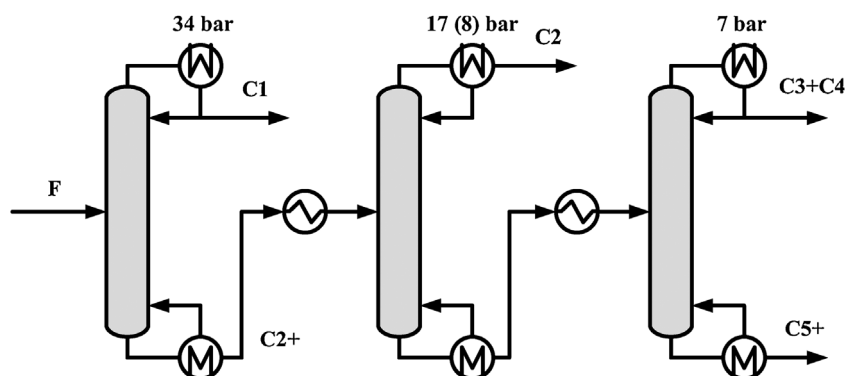

Fig. 1 – Conventional direct three column sequence.

Table 1 – Feed and product specifications for NGL fractionation in a floating natural gas liquefaction plant.

Stream name	Feed	C1	C2+	C2	C3 + C4	C5+
Temperature, °C	54.0	−91.7	159.0	−14.2	33.6	153.4
Pressure (absolute), bar	34.0	34.0	34.0	17.0	7.0	7.3
Vapour mole fraction, −	0.09	0.00	0.00	1.00	0.00	0.00
Molar flow rate, kmol h^{-1}	108.5	17.4	91.1	9.2	29.9	52.0
Mass flow rate, kg h^{-1}	6951.0	282.0	6669.0	270.8	1559.7	4838.5
Composition, mol%						
Methane	16.38	98.97	0.59	5.80	0.00	0.00
Ethane	8.19	1.02	9.56	92.82	0.58	0.00
Propane	11.73	0.00	11.73	1.38	42.09	0.00
I-Butane	6.08	0.00	6.08	0.00	22.04	0.00
N-Butane	9.81	0.00	9.81	0.00	34.87	0.40
2-2-Dimethylprop.	0.46	0.00	0.46	0.00	0.39	0.74
I-Pentane	6.45	0.00	6.45	0.00	0.02	13.46
N-Pentane	4.76	0.00	4.76	0.00	0.00	9.93
Heavies	36.14	0.00	36.14	0.00	0.00	75.47

Table 2 – Operating conditions and design data for conventional- and thermally coupled direct sequence[a].

	Conventional direct sequence		TCDS/DS-DWC	
	Deethanizer	(C3 + C4)	Prefractionator	Main column
Feed temperature	28 °C	78 °C	28 °C	–
Feed cooler duty	−643 kW	−43 kW	−643 kW	–
Condenser pressure	8 bar	7 bar	8 bar	8 bar
Condenser temperature	−39 °C	34 °C	−39 °C	+39 °C
Condenser duty	−52 kW	−381 kW	−52 kW	−308 kW
Distillate vapour fraction	1	0	1	0
Reboiler duty	315 kW	593 kW	–	830 kW
Total reboiler duty		908 kW		830 kW
Number of stages	20	56	17	54
Feed stage	6	27	5	20

[a] The demethanizer column is not included as it is identical for the two alternatives.

3. Effective thermal coupling arrangements

Following the suggestion from the conclusions of our previous paper (Halvorsen et al., 2016), it is expected that fine-tuning the feed condition and prefractionator split can reduce the cold utilities demand markedly. Indeed, it appeared that by subcooling the feed C2+ to 28 °C (in previous case it was 43 °C), the cold utilities demand could be reduced from −265 kW to −175 kW. This, however, is still higher than for a conventional deethanizer operating at 17 bar, and, in fact, it is much higher than what would be needed if the conventional deethanizer would be designed to operate at the same absolute top pressure as the DWC, i.e. 8 bar in the condenser. In that case, i.e. at the same condenser temperature (−39 °C), the deethanizer condenser duty would drop to −52 kW. Therefore, −52 kW was taken as proper target for the thermal coupling effort considered in the present study.

Important design and operating parameters for the direct sequence with the deethanizer and (C3 + C4) recovery column operated at top pressures of 8 bar and 7 bar, respectively, are shown in first two columns, on left hand side of Table 2.

Note that subcooling the feed for the deethanizer operating at 8 bar is quite demanding (−643 kW). The temperature level of the feed of both columns as well as overheads vapour leaving the top of the (C3 + C4) recovery column is so high that the required temperature drop in both cases can be easily achieved using sea water as the cooling medium. The deethanizer column is equipped with a partial condenser and both columns utilize a partial reboiler, which act as external equilibrium stages.

Note that in the present case in Table 2, i.e. with a pressure of 8 bar in the partial condenser of the deethanizer, the corresponding condensation temperature is much lower (−39 °C) than in the original 17 bar case (−14 °C for distillate delivered as vapour, or −28 °C for liquid as employed in the preliminary study). Nevertheless, this is not a problem because natural gas liquefaction facilities usually have complex, multi-level compression refrigeration systems that can easily accommodate the refrigeration level needs of a NGL fractionation plant (Smith, 2005). However, the cost of sub-zero cooling depends on the refrigeration temperature, and at the levels considered in the present study, it is absolutely a dominating factor in operating costs, largely outweighing potential gains on the heating side due to a reduced reboiler temperature. An illustrative example indicating how the cold utility costs compare to hot utility costs at various temperature levels with water cooling cost as a reference can be found elsewhere (Jobson, 2014). A more recent paper by Luyben (2017) provides updated refrigeration cost data for processes operating over a much wider range of temperatures (−25 °C to −190 °C).

Considering the present NGL fractionation case, the key to success is to arrange thermal coupling, i.e. the internal configuration of a DWC, to have the same cold utilities demand as the deethanizer column in the conventional direct sequence (−52 kW at −39 °C). The full DWC, though it is the thermodynamically most efficient configuration, cannot achieve this, because, as mentioned before, more vapour needs to be condensed at the lowest temperature. However, as indicated during detailed preliminary evaluations using the Vmin-diagram method (Halvorsen and Skogestad, 2011),
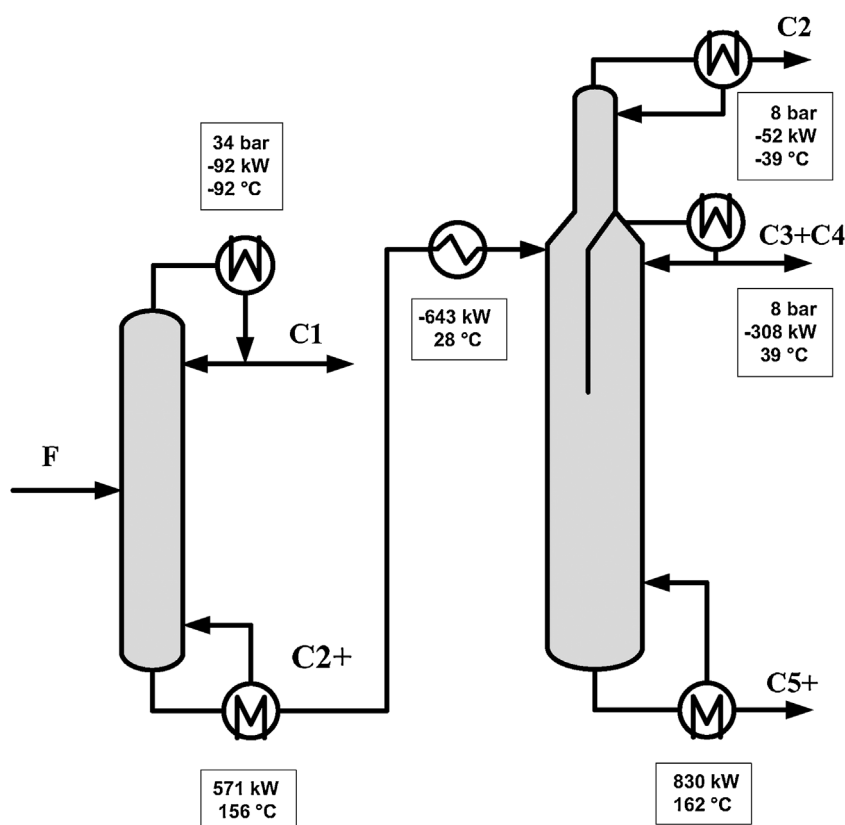
Fig. 3 – Conventional demethanizer combined with a modified full-DWC, i.e. a partitioned thermally coupled direct sequence (DS-DWC).

this can be achieved by rearranging the main column side of the full DWC, to include, as indicated in Fig. 2, an intermediate condenser placed immediately above the side product vapour draw-off, to condense the vapour in excess to that required. The resulting overhead vapour from the DWC has the same composition and rate as in the case of conventional deethanizer. Namely, in the case of a full DWC, the optimal prefractionator vapour rate is just slightly lower than the vapour rate needed to get the specified ethane product directly from the prefractionator.

However, such an internal condenser arrangement appears to be somewhat impractical, and a better solution is to simply rearrange the internal configuration of the full DWC to have the prefractionator separated from the main column, so that the former fully reflects the operation of the deethanizer from the conventional direct sequence. In a given case, this could be implemented by simply installing a lateral, sloped partition wall on main column side immediately above the draw-off stage and using an external, seawater cooled condenser to generate the reflux and (C3 + C4) product stream as liquid. Being useless, the complete section of the full DWC main column above the draw-off stage can then be omitted. This, as shown schematically in Fig. 3, would allow further reduction in shell diameter above the side-product draw-off position, to be dimensioned just to accommodate rectification section of the prefractionator.

Thus, the internal configuration of a conventional full DWC, which proved to be inadequate for NGL fractionation purposes because of excessive cold utilities demand, has been reduced to a simpler and more effective one. This arrangement with two separated column sections resembling fully the top arrangements of deethanizer and (C3 + C4) recovery column from the conventional two-columns sequence, is generally

known as a side-rectifier arrangement (Smith, 2005). More precisely, it represents a thermally coupled direct sequence (TCDS), because, as shown schematically in Fig. 4, the vapour requirement of the deethanizer column is supplied by the reboiler of the (C3 + C4) column.

This arrangement was considered about 20 years ago by Manley (1996, 1998), under the name "recycle coupled deethanizer and depropanizer", to improve the energy efficiency of conventional, on-land NGL fractionation complexes. He evaluated various options, and in his 1998 publication he proposed a highly thermally integrated four-column sequence, with ethane, propane, n-butane, i-butane and C5+ fraction (natural gasoline) as products, combining external and internal thermal coupling. Other related work further elaborates thermal coupling and DWC arrangements suitable for recovery of propane and heavier components (Long and Lee, 2012). It appears that thermal coupling, including a cold deethanizer column, is not considered to be promising, but as shown in what follows this may be a wrong perception.

Indeed, detailed simulation with the same subcooled feed and stage requirement, shows that the TCDS arrangement can achieve the same product specifications at the same cold utilities requirement as a conventional sequence with a deethanizer operated at the same top pressure. The relevant numbers are shown on the right-hand side of Table 2. The overall mass balance, internal liquid and vapour flows, reboiler and condenser duties and section stage requirement are shown in Fig. 5. A comparison with the direct sequence (Table 2), indicates that thermal coupling in this case requires 9% less hot utilities, without any penalty on the cold utilities demand side.

This is slightly more than expected from the Vmin-diagram method. Fig. 6 shows the contours of Vmin diagram for each
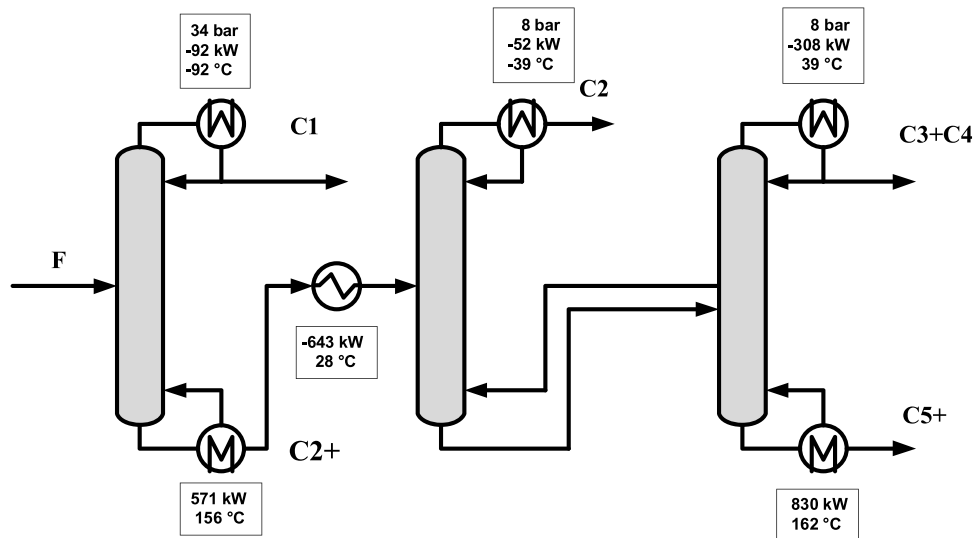
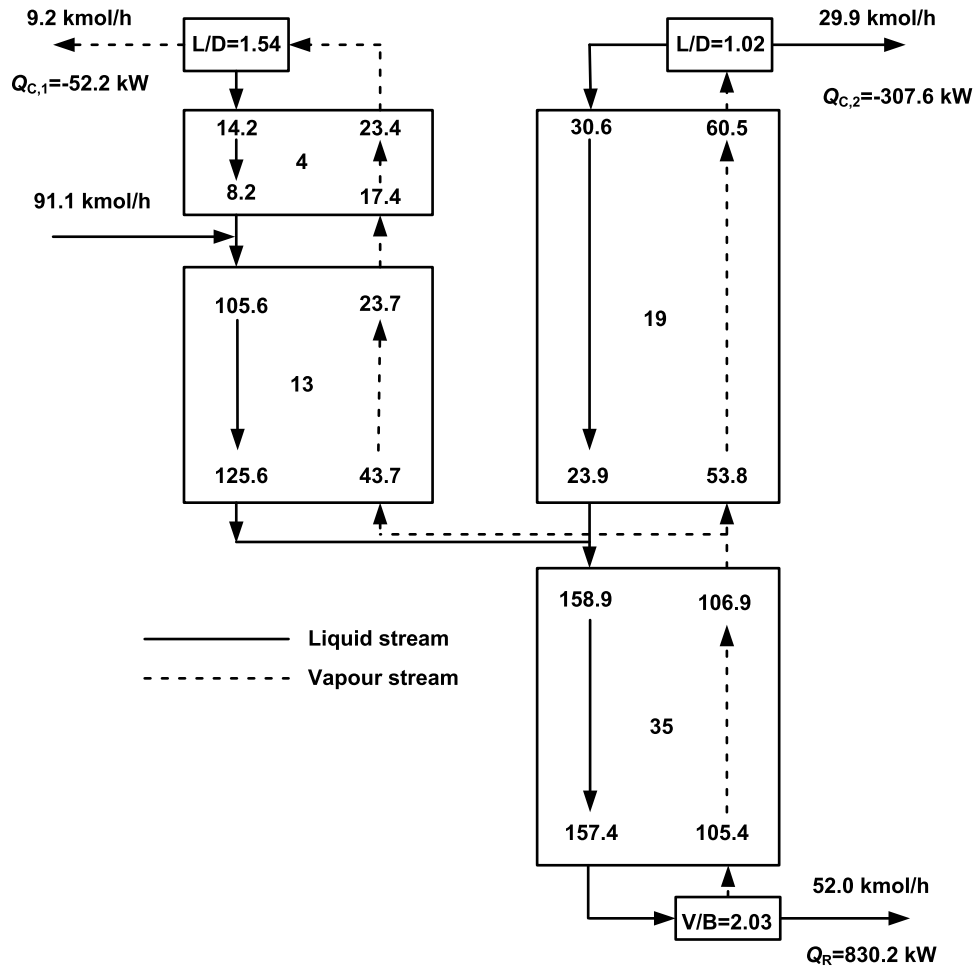Fig. 4 – Conventional demethanizer combined with a thermally coupled direct sequence (TCDS) arrangement.

Fig. 5 – Material and energy balance of TDCS arrangement with internal molar liquid and vapour flow rates, and stage requirement per section according to detailed simulation.

of three compared cases, a direct sequence (DS) with the demethanizer and (C3 + C4) column operating at 8 and 7 bar, respectively, and the two thermal coupling arrangements (TCDS and full-DWC) operating at 8 bar. The diagram corresponding to the full-DWC, i.e. a fully thermally coupled Petlyuk arrangement, exhibits the common two-peak contour, where the highest peak represents the total V/F requirement. The direct sequence has two independent columns

which gives two individual peaks. The first one (deethaniser column) is equal in height to that of thermally coupled arrangements, and the second one, starting at $D/F = 0.1$ (feed is assumed to be a saturated liquid), reflects the V/F demand for the (C3 + C4) column. The sum of these two peaks is the total V/F requirement for the DS configuration, i.e. $V/F = 0.21 + 0.76 = 0.97$. Note that for the full DWC and the TCDS configurations, all the vapour comes from a sin-
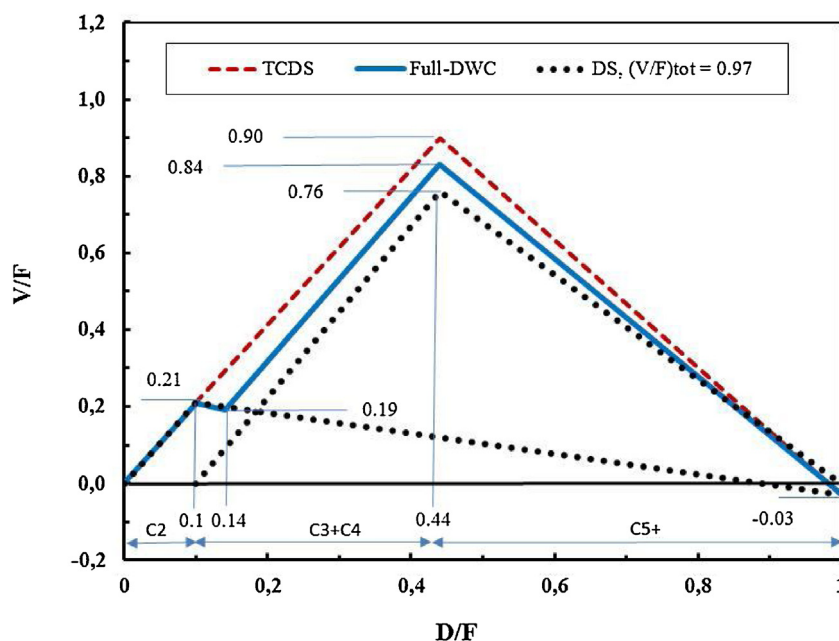
Fig. 6 – Vmin-diagram comparing boil-up (V/F) requirements of direct sequence (DS), a full DWC, and its thermally coupled direct sequence counterpart (TCDS).

gle reboiler and the highest peak represent the total vapour requirement.

The feed (C2+) is subcooled to the same temperature in all succeeding configurations, so in all cases a fraction of the vapour ascending from the stripping section of the deethanizer column is condensed at feed stage, which means that additional heat is required in the reboiler. In the present case, this is equivalent to a vapour-to-feed ratio of –0.03. This amount added to the (V/F) values from peaks gives the overall vapour flow rate that must be supplied in the reboiler.

For the full-DWC configuration, the preferred split is between C2 and C5+, and the corresponding vapour requirement ($V/F = 0.19$ at $D/F = 0.14$) for the prefractionator sub-column in a full Petlyuk sequence is lower than that required for sharp C2/C3+ and C4/C5+ splits occurring in downstream sub-columns. For the direct sequence (DS) and thermally coupled direct sequence (TCDS), the vapour requirement for the C2/C2+ separation is the same, i.e. $V/F = 0.21$ at $D/F = 0.1$, but in both cases the prefractionator (deethanizer) column needs to perform a sharp split, which, according to theory, increases the energy demand.

Because of a slightly increased V/F from the preferred split point to produce pure C2, the contour of the Vmin-diagram of the TCDS arrangement is characterized by a single peak with the side of the triangle that is practically an extension of that of the first ($V/F = 0.21$) peak. Note that the difference between V/F and D/F represents the fraction of overheads vapour condensed ($L/F = V/F − D/F$) and returned to the column as reflux. This is same in all three cases and sets the refrigeration requirement for the deethanizer (prefractionator) condenser. The horizontal distance between two peaks represents the relative size of (C3 + C4) product stream. Thus, this simple Vmin diagram provides complete mass balances for the given separation, indicating that the main effort in the present case is that related to separation between (C3 + C4) and C5+.

The difference in the heights of the relevant peaks (at $D/F = 0.44$) shown in Fig. 6 indicates that the thermally coupled arrangements have significantly lower minimum vapour

requirement than a conventional direct sequence for the same separation task, and that a TCDS arrangement is pronouncedly less efficient than a full Petlyuk arrangement.

Per definition, the boilup requirement, i.e. V/F delivered at reboiler, found from the Vmin-diagram is the minimum one and corresponds to an infinite number of equilibrium stages. For practical purposes, it appears that using four times the minimum number of stages (corresponding to infinite V/F), gives a good trade-off between actual V/F and the number of stages (Halvorsen and Skogestad, 2003; Dejanović et al., 2011).

To find the actual V/F, detailed simulations were carried out in CHEMCAD utilizing a sequential algorithm with optimization of vapour and liquid splits in an inner, and optimization of stage and reflux requirement in an outer iteration loop (Dejanović et al., 2011). The results of the Vmin-diagram method were used as initial guesses.

According to the results of the detailed simulations, summarized on right hand side of Table 2, a TCDS arrangement enables a saving of 9% in heating demand compared to a conventional direct sequence. This is, indeed, much less than the typical expected savings (∼ 30%) based on experiences with common thermal coupling applications. However, in the present application, the feed mixture is extremely wide boiling with large differences in the relative volatilities of the key components. This is unfavourable for thermal coupling. However, as mentioned before, it is not the energy-saving potential but its equivalent in reduction of carbon dioxide footprint that makes thermal coupling an interesting option for NGL fractionation plants.

The single-shell implementation of the thermally coupled direct sequence in Fig. 3, with a rectification section of the prefractionator placed above the main column, will be called a direct sequence DWC (DS-DWC). This column is taller than any of the two conventional columns in TCDS arrangement. However, in the present case this will be not so pronounced because the rectification section of the prefractionator contains only five stages. Most importantly, in such an arrangement the very cold part of the column (the rectification section of the deethanizer or prefractionator) is placed
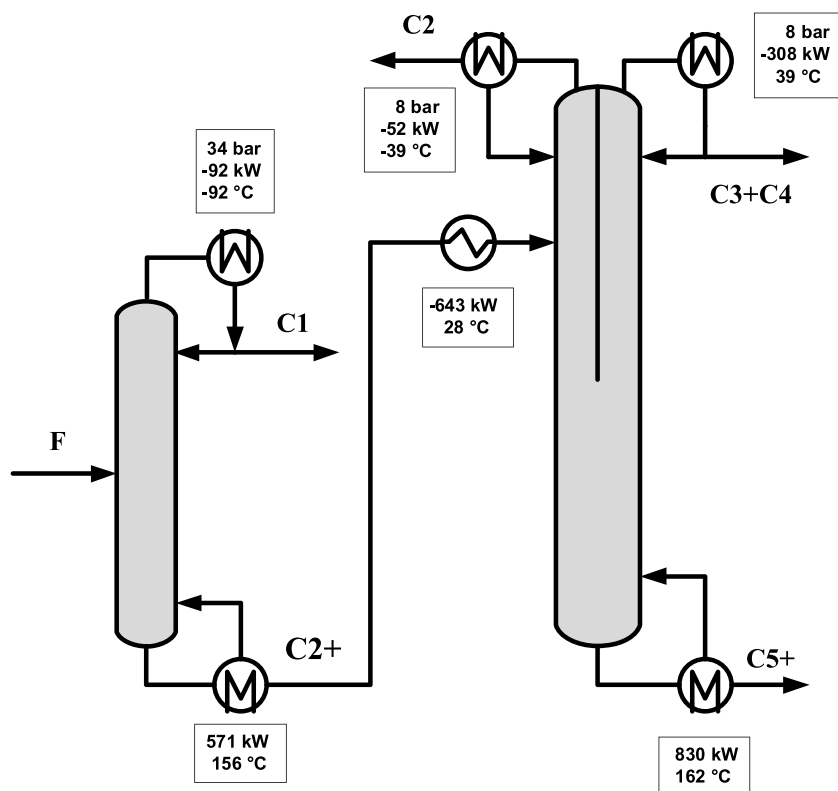
Fig. 7 – Conventional demethanizer combined with a partitioned side-rectifier (SR-DWC).

above the much hotter part of the column. In the latter, the temperature difference across the partition wall, separating the stripping section of the prefractionator from the rectification section of the main column, though significant (∼ 20 °C at the top of partitioned section), is not so large that it could induce performance deteriorating effects (Kaibel, 2014).

According to Smith (2005), a thermodynamically equivalent alternative for a thermally coupled direct sequence (TCDS) or its partitioned counterpart DS-DWC, is the before mentioned side-rectifier arrangement (TCSR), where the vapour required by the second column is supplied by the reboiler of the first column.

Both two-shell (TCDS and TCSR) configurations imply draw-off and transport of a vapour from one shell to the other. This requires selecting the operating pressures accordingly, and the additional pressure drop needs to be accounted for to ensure maintenance of the required vapour split. The pressure in the top of the deethanizer (prefractionator) and (C3 + C4) columns was kept at the same value, i.e. 8 bar. This pressure that has been taken as basis for detailed simulation of both cases is quite high, and the pressure drop involved, even in case of trayed columns, is relatively small to influence significantly phase equilibria and tray or packed column efficiency. In the detailed simulations a certain amount of pressure drop was adopted that correlates well with estimates in this respect. Details on hydraulic design of TCDS (Fig. 4) and DS-DWC (Fig. 3) arrangements employing packed and tray columns can be found in another, just published paper (Dejanović et al., 2018).

Importantly, this dual condenser arrangement allows control of the vapour split by balancing the condenser duties, which is a prerequisite for maintaining the prefractionator at the desired split for producing the top C2-product. A second advantage is in the simplicity of the construction, with two shells in parallel. Onshore, this allows utilization

of proven tray designs. On a floating natural gas liquefaction plant, a lower shell height makes a column less prone to motion-induced liquid maldistribution and accompanying column efficiency reduction effect. Last but not the least, both TCDS and TCSR could be highly interesting as retrofit options for existing NGL fractionation plants. This, however, may require installation of an additional, larger-diameter column to accommodate the increased vapour volume associated with low-pressure operation of the prefractionator, i.e. the deethanizer.

A side-rectifier arrangement can also be accommodated in a single shell, that is, as a dividing wall column with a partition wall in the upper section of the column (Smith, 2005). Similar to the DS-DWC, this partitioned side-rectifier arrangement (SR-DWC), shown schematically in Fig. 7, ensures additional plot area and weight benefits.

However, unlike the DS-DWC shown in Fig. 3, the top partition arrangement shown in Fig. 7 is not a suitable configuration for such a wide-boiling feed as considered in the present study, because it implicates a very large temperature difference across the partition wall, reaching up to 80 °C at the top. Although adequate insulation, as described in a patent by Kaibel et al. (1998) could reduce the heat transfer across the partition wall to a tolerable level, it is questionable whether a column can be designed and constructed to accommodate in a safe way, the process and mechanical integrity related adverse effects of excessive thermal expansion of the shell on the hot side.

In any case, the extent of thermal expansion of the shell on the hot side is a concern. If significant, it needs to be accounted for during installation of internals to ensure proper levelness under operating conditions (Kaibel, 2014; Jansen et al., 2014). This is demanding and related uncertainties may emerge as a reason to decline this compact construction. However, to avoid a too high temperature difference across the partition

wall, the cold rectification section of the prefractionator (C2 column) could be installed above the rectification section of the main (C3 + C4) column, in the same manner as shown for TCDS configuration in Fig. 3. In fact, both thermal coupling arrangements, TCDS and TCSR, when placed into one shell have the same internal configuration.

Both the external (TCDS) and the internal (DS-DWC) thermal coupling arrangements, if primarily considered for installation on a floating facility, should utilize columns equipped with structured packings, as discussed elsewhere (Halvorsen et al., 2016). In a recent paper, Schultes et al. (2018) provide pilot-scale experimental evidence obtained using a packed column operating under constant tilt or under moving conditions with different frequencies and extent of tilt, indicating that random packings may exhibit even less pronounced loss of efficiency than structured packings under same operating conditions. Therefore, they recommend use of random packings for floating vessels applications similar to those proven in on-land applications.

In a similar onshore plant, with a relatively much larger capacity and equipment size, tray columns would be the preferred choice. A detailed description of a trayed SR-DWC can be found in a patent by Ognisty and Manley (1998). Practical implementation is growing slowly, but there is evidence that the number of trayed DWCs as well as the related construction and installation knowhow is increasing steadily (Bhargava et al., 2016).

4. Conclusions

The energy efficiency of fractionation complexes using refrigeration for condensation, as encountered in natural gas liquefaction plants, offshore and onshore, could be significantly improved by implementing thermal coupling whereappropriate, provided that the cold utilities demand will not exceed that of the columns in the conventional sequence.

As demonstrated in this study, combining a conventional demethanizer with a thermally coupled direct sequence (TCDS) that replaces the conventional deethanizer and (C3 + C4) recovery columns provides a realistic opportunity for natural gas processing and refining industries to achieve significant energy savings, e.g. carbon dioxide emissions reduction, in a cost-effective way (Figs. 3, 4 or 7).

This simple and effective thermal coupling arrangement is amenable for both new designs and as a retrofit option, and its partitioned counterpart, the DS-DWC in Fig. 3 would ensure further weight and plot area benefits.

Final choices in this respect, including the choice of the most appropriate column internals will largely depend on important design, construction, and operation considerations. This may differ for offshore and onshore plants to the extent depending on specific site requirements.

References

Bhargava, M., Kalita, R., Gentry, J., 2016. Improved Distillation Efficiency. PTQ, Q4, pp. 1–5 www.digitalrefining.com/article/1001301.

Dejanović, I., Matijašević, Lj., Olujić, Ž., 2011. An effective method for establishing the stage and reflux requirement of three-product dividing wall columns. Chem. Biochem. Eng. Q. 25, 147–157.

Dejanović, I., Halvorsen, I.J., Jansen, H., Olujić, Ž., 2018. Hydraulic design of thermally coupled columns and a DWC for NGL fractionation plants. Chem. Biochem. Eng. Q. 32 (4), 391–400.

Halvorsen, I.J., Skogestad, S., 2003. Minimum energy consumption in multicomponent distillation. 1. Vmin diagram for a two-product column. Ind. Eng. Chem. Res. 42, 596–604.

Halvorsen, I.J., Skogestad, S., 2011. Energy efficient distillation. J. Nat. Gas Sci. Eng. 3 (4), 571–580.

Halvorsen, I.J., Dejanović, I., Maråk, K.-A., Olujić, Ž., Skogestad, S., 2016. Dividing-wall Column for fractionation of natural gas liquids in floating liquefied natural gas plants. Chem. Eng. Technol. 39, 2348–2354.

Jansen, H., Dejanović, I., Kaibel, B., Olujić, Ž., 2014. New horizons for dividing wall columns. Chem. Eng. 121 (August), 40–48.

Jobson, M., 2014. Energy considerations in distillation. In: Górak, Andrzej, Sørensen, Eva (Eds.), Distillation Fundamentals and Principles. Academic Press/Elsevier.

Kaibel, B., 2014. Dividing wall columns. In: Górak, Andrzej, Olujić, Žarko (Eds.), Distillation Equipment and Processes. Academic Press/Elsevier.

Kaibel, G., Stroezel, M., Pfeffinger, J., 1998. Distillation column for separating a liquid mixture into plurality of pure fractions. United States Patent 5785819.

Long, N.V.D., Lee, M.Y., 2012. Improvement of natural gas liquid recovery energy efficiency through thermally coupled columns. Asia Pac. J. Chem. Eng. 7 (Suppl. 1), S71–S77.

Luyben, W.L., 2017. Estimating refrigeration costs at cryogenic temperatures. Comput. Chem. Eng. 103, 144–150.

Manley, D.B., 1996. Distillation of natural gas liquids. In: Proceedings of the 75th Annual Convention, Gas Processors Association, March 11–13, Denver, CO, USA, pp. 67–74.

Manley, D.B., 1998. Thermodynamically efficient distillation: NGL fractionation. Lat. Am. Appl. Res. 28 (4), 211–216.

Ognisty, T.P., Manley, D.B., 1998. Partitioned distillation column. United States Patent 5755933.

Schultes, M., Brauer, J., Chen, P., Doong, S., 2018. Marinization of mass transfer columns for FLNG applications. Chem. Eng. Trans. 69, 301–306.

Smith, R., 2005. Chemical Process Design and Integration. John Wiley & Sons Ltd.