Continuous desalting concept on ionic liquid-mediated de-acidification process of crude oil: A pilot study

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Abstract. Desalting process concept was tested using methyltrimethylammonium methylcarbonate [N$_{4441}$][MeCO$_3$] treated Pyrenees crude oil (initial Total Acid Number (TAN) of 1.6 mg KOH/g oil) with the aim to gain empirical evidences on the effectiveness of in-line water washing and electrostatic aided phase separation as mean to recover the naphthenic acid derivatives for recycling. The treated crude oil (final TAN value of less than 0.3 mg KOH/g oil) was subjected to typical operating scheme such as single stage desalting and effects of water wash volumes. The novelty of the work comes from the utilisation of ionic liquids to neutralise acid components of the crude oil. Furthermore, the work is also able to test the hypothesis of whether naphthenate salts behave as is its inorganic counterpart and quantify the solubility behaviour in water as extraction medium. The effectiveness of such scheme will be measured against naphthenic acids derivative percent recovery in the wash water. The results indicate the electrostatic conditions can facilitate the recovery of the naphthenate salts post neutralization with high recovery rate of average of 70.6 % with 30 % water wash volume in a single-stage contact, observed over 12 hours steady-state operation. The water wash weight was observed to increase post separation which indicate hydrocarbon carry-over in the heavy phase due to formation of tight water – oil emulsion. The technique is viable should the amount of water required is available and the process water can be recycled safely into the desalter again without causing tripping to the desalter. Ionic liquid can be used in conjunction with desalter and the presence of electrostatic field did hasten the separation of the phases, however the amount of water used may hinder the viability of the solution.

1. Introduction
Desalting process is a common feature of a refinery and is designed to remove inorganic salts and dehydrate crude oil through water-oil contact and hasten phase separation through coalescence of water droplets in a vessel that has electrostatic field generated in it [1-3]. The removal of salts and water is vital to the downstream process as the absence of the two will reduce fouling and corrosion tendencies in the hot heat exchanger train, boiler tube protection and water load reduction at overhead section of crude distillation unit [4]. Typical desalter package will have an electrostatic generator, separation vessel, pressure drop mixer and mud wash. There are two types of desalters which are Alternating
Current (AC) and Dual Polarity. The main difference is the type of electrostatic field use as the latter employs both AC and DC field to induce coalescence of water droplets.

Recent reviews provide an overview of techniques used to reduce or eliminate naphthenic acids (NAs) in crude oils which include esterification, decarboxylation and neutralisation with metal hydroxides [5, 6]. Naphthenic acids are carboxylic acids with one or more carboxylic acid functionalities attached to one or more saturated cyclic structure, in general. Naphthenic acids are commonly written as C₈H₂n+2O₂ with z values corresponds to the degree of hydrogen deficiency that starts from 0 to negative even integer. They can also refer to other chemical structures containing carboxylic acids such as heteroatom (S, N, B) or aromatics derived structures [7-12]. Ionic liquids or ionic liquid-based chemistry were explored as a method to reduce or eliminate NAs due to its tuneable properties i.e. solubility, low volatility and basicity [13-16]. Anderson et al. [17, 18] demonstrated the usage of two different classes of ionic liquids i.e., amino acid based ionic liquids (AAILs) and carbonate based ionic liquids on solvent-assisted removal of naphthenic acids from crude oil matrix. The work also demonstrated that upon reaction with ionic liquid, the solubility of the resulting naphthenate salts are more in the methanol phase as compared with metal naphthenates form, suggesting good recovery of the ionic liquid for regeneration effort. Hydroxide based ionic liquids were found to be effective with tetraalkylammonium cation being more efficient than tetraalkylphosphonium cation and can be regenerated using ion-exchange method [19]. There are few examples in the open literature that exemplified the synergy between desalter and ionic liquid technology. Vadara et al. [20, 21] utilised aqueous alkoxyalted tertiary amine and contacted it with crude oil containing naphthenic acids in the normal desalter setup. Similarly, like BASIL™ process, the amine shall form quaternary amine salts with the naphthenic acids and partitioned in the aqueous phase [22]. However, there was evidence of 4% oil loss to the aqueous phase and no information on alkoxyalted amine salt cross solubility into the oil phase was offered. Simple organic carboxylic acids were shown to have different solubility behaviour between organic phase and ionic liquid. Manan et al. demonstrated the phase behaviour of cyclohexanoic acid, a model naphthenic acid with different ionic liquids of which the solubility of cyclohexanoic acid is dependent on lipophilicity of the cation and coordinating capability of the anion such as trifluoroacetate and triflate [23]. However, with increasing alkyl chain of the cation, so does the solubility of dodecane indicating higher loss of hydrocarbon. Similar observations were also made with phenolate based ionic liquids [24]. Coto et al. recently disclosed 1-ethyl-3-methylimidazolium ethyl sulfate ([EMIM][EtSO₄]) achieved 25% TAN reduction post contact with model oil system [25]. Functionalization of the 1-butylimidazolium tetrafluoroborate ([Bmim][BF₄]) cation moiety was found to increase the interaction between naphthenic acid and substituional groups such as carboxylic acid (-COOH) and methyl ester (-COOCH₃) through formation of hydrogen bond [26]. This paper shall discuss in detail the work described by Anderson et al. [27]. The aim of this work is to examine and quantify process stability, mass balance generation and the behaviour of ionic liquid treated crude oil in a continuous desalting process. Desalter’s operating parameter of water wash volume and one versus two stage desalting will be benchmarked against the recovery percentage of the cation. Ultimately, the findings from this work is to aid engineering design and operating philosophy of a facile process that utilises ionic liquid and desalter as means to reduce TAN value of crude oils of which shall increase flexibility of refinery, widen crude oil diet for higher refining margin and mitigate adverse risks to the vacuum distillation units due to high temperature corrosion.

2. Materials and methodologies

Pyrenees crude oil was supplied by PETRONAS from Australia (TAN of 1.63 mg KOH/g oil). Tributylamine (99 %), n-hexane (HPLC), methanol (HPLC), deuterated benzene (C₆D₆) and deuterated dimethyl sulfoxide (DMSO) were supplied by Sigma-Aldrich. Dimethylcarbonate (99 %) was supplied by Alfa Aesar. Purified water with conductivity of 18 mS/cm was used throughout. All chemicals are used as received unless stated otherwise.

2.1. Synthesis of methyltributylammonium methylcarbonate [N₄441][CH₃CO₃]
The preparation of methyltributylammonium methylcarbonate \([\text{N} \text{4441}][\text{CH}_3\text{CO}_3]\) was performed according to published method [28]. Typically, the ionic liquid obtained by reacting tributylamine (37.07 g, 0.2 mol) with dimethyl carbonate (36.03 g, 0.4 mol) in 50 cm\(^3\) of methanol in an autoclave at 6 bar and 393.15 K for 12 hours. After the reaction the excess methanol was removed under reduced pressure and was washed with n-hexane (3 \(\times\) 100 cm\(^3\)) to remove excess dimethylcarbonate affording a light-yellow liquid. The yield (calculated using NMR) was 78\% for the \([\text{N} \text{4441}]^+\) based ionic liquid. The ionic liquid was then further dried under high vacuum (1.0 \(\times\) 10\(^{-3}\) Pa) at 338.15 K for 24 hours. The resultant ionic liquid crystallised upon standing at room temperature.

2.2. Total acid number (TAN) measurement

Typical procedure used for TAN determination using an automatic potentiometer with combined glass electrode (Metrohm Titrino Plus model 785), in accordance of ASTM D664, is by weighing a known amount of oil in 125 ml beaker which is then diluted with titration solvent. The titration solvent is a mixture of methanol/isopropanol, toluene and water in the ratio of 45:50:5 by volume. The electrode needs to be washed with titration solvent before usage. The homogenous solution is then titrated using 0.1 N of methanolic potassium hydroxide solution (Sigma Aldrich) until it reaches equilibrium point. The equilibrium point is then calculated using equation (1) below to get the final TAN value.

The Total Acid Number (TAN) is then calculated based on the equation (1).

\[
\text{Acid Number} = \frac{(A - B) \times M \times 56.1}{w}
\]  

where \(A\) is the volume of KOH titrated to reach the equilibrium point, \(B\) is the volume of KOH for blank titration, \(M\) is the concentration of the KOH solution and \(w\) is the sample mass. The accuracy of the method is dependant of mass of sample with reproducibility around 2\%. The electrode is conditioned by immersing in pH 4 buffer solution (Metrohm UK) for a minimum of 10 minutes before reusing it again for titration. After several titrations, it is recommended for the electrode to be lightly wiped with soapy water and reconditioned in its electrolyte solution (3.0 M potassium chloride in ethanol, Methrohm UK) for overnight.

2.3. Desalter setup and crude oil neutralisation

The high throughput desalter pilot unit is located at Intertek PARC at Pittsburgh, USA and was manufactured by Howe Baker Engineers (Serial No.: 1498; Year built: 1980). The unit consists of two process stages unit, shown in figure 1. The pilot unit has capability of 53 litres/hr (0.053 m\(^3\)/hr), injection pump facility of maximum 0.5 litres/hr, max temperature of 230 °C and working pressure of 10 bar. The setup is equipped with mixing valve with max pressure drop of 2 barg and flexibility of running the experiment in one or two-stages.
Ionic liquid and crude oil were pre-mixed in a holding tank with recirculation pump using a predetermined ratio between 0.7 wt% to 1.0 wt% at constant stirring of 500 rpm at 80 °C for 24 hours. The neutralisation effectiveness was monitored using ASTM D664 method for determining total acid number [29]. The crude oil is considered completely neutralised if the value reached <0.3 mg KOH/g of oil. The neutralised crude oil was pumped according to the conditions listed in table 1 and pre-heated using a series of heat exchanger. Wash water of pre-determined ratio between 10 – 30 v = vol%, was contacted with the oil phase using static mixer and pressure drop mixer of 10 barg to induce mixing and formation of droplets. The mixed phase was then subjected to the electrostatic field and allows to phase separate based on the vessel designed residence time of 15 mins. All the incoming and outgoing flows were stored and measured using weight balance recorder. Samples of both water and oil phase were taken from the vessels at predetermined 30 mins period for analysis. The process was allowed to reached equilibrium when the mass of wash water drawn became stable. As for this work, a minimum of 15 hours of continuous flow was set or until the crude oil runs out, whichever comes first. All experiments were performed in duplicate unless stated otherwise.

2.4. Quantification of naphthenate salts in water phase
The method is described in detail elsewhere [28, 30]. The ionic liquid recovery is expressed in mol% based on $^1$H NMR spectroscopy using DMSO as an internal standard and the methylene protons adjacent to the ammonium centre (N-CH$_3$) as the cation signal. The molar amount of the recovered ionic liquid, $n_2$ is calculated by integrating the signal attributed to the cation in reference to the internal standard signal, expressed by equation (2).

\[
\text{Recovery of naphthenate salts } n_2 = \alpha \times n_{\text{std}} \times \frac{P_1}{P_2} \quad (2)
\]

\[
\text{Recovery percentage, } \% = \frac{n_2}{n_1} \times 100 \quad (3)
\]

where $\alpha$ is the total mass of extracted material/mass used for NMR analysis, $n_1$ is the amount of IL used and the expression $P_1/P_2$ is the ratio of the amount of proton of the cation signal over the amount of proton in the internal standard and $n_{\text{std}}$ is the mole amount of the internal standard used. The percentage IL recovery is then calculated from equation (2).
3. Discussions
The operation conditions of the desalter pilot plant are summarised in table 1 below. The operation conditions were chosen based on typical operating conditions used for this type of crude oil and hydraulic limitation in PETRONAS refineries [31]. Table 2 below is the properties of the crude oil used in the pilot plant and commonly related to the operation of a desalter such as relative density, salt content and water content. All the properties tested and found it to be consistent with its crude assay [32]. Consistency of the sample and the published crude assay is important because of two reasons which are it is a new crude oil produced in Australia and is being considered to be a good crude oil to be part of the refinery crude oil diet that will use this technology. Any data generated here will be used in the refinery as base line for refinery crude oil and yield analysis. Secondly, it is important that the variation of the acid content to be at the minimum as this would confirm that the oil was drawn from the same hydrocarbon field and co-mingled in the same vessel prior to sampling. Sampling was done at the charter crude oil carrier off the concerned refinery. As such the heterogeneity of the naphthenic acids present in the oil was kept at minimum, which was gauged from the density, API value and ASTM D664.

| Parameter                      | Unit    | Value |
|-------------------------------|---------|-------|
| Flowrate                      | m³/hr   | 30    |
| Temperature                   | °C      | 90    |
| Pressure                      | barg    | 10    |
| Water flowrate                | vol%    | 10 and 30 |
| Electrostatic field strength  | kV      | 6.0   |
| Pressure drop across the mixing valve | barg | 0.5 |

3.1. Process mass balance and stability
One of the key questions in adapting the desalter process unit with the current application is how stable the extraction of ionic liquid and its derivatives under electrostatic field. The consensus coming off the performed series of experiments with varying conditions is there is no observable adverse effect of electrostatic field and presence of ionic liquid and its derivatives (naphthenate salts) on process parameters namely, temperature and pressure, voltage stability, resistance or current value and chemical structure of ionic liquids. Crude oil here shall refers to the de-acidified crude oil or crude oil that has a
TAN value of < 0.3 mg KOH/g post mixing with [N\text{4441}][\text{CH}_3\text{CO}_3]. Figure 2 below illustrates the amount of total feed to a process in which water is used in an amount of 30 % by volume of the mixture, as well as the amounts of crude oil product and water product from the process, as a function of process operating time. Followed by figure 3 which is a graph showing the amounts of feed mixture and feed water to a process in which water is used in an amount of 30 % by volume of the mixture, as well as the amounts of crude oil product and water product from the process, as a function of process operating time.

As mentioned before, ionic liquids’ properties can be tuned to have differ from the metal salt analogues in this case would solubility of the resultant organic salts or naphthenate salts. Again, by taking desalter operation as a model, water would be the best solvent as it is available in refinery system. Therefore, it is best to empirically determined the weight of the outgoing streams. The masses of incoming streams were determined using density and flow rate. Both figure 2 and figure 3 indicate that all the mass i.e., de-acidified crude oil and wash water, going into the desalter are accounted for which includes the ionic liquid. The outgoing stream of wash water post-contact with electrostatic field was found to increase in weight, attributed by the dissolved naphthenate salts and possibly small amount of oil carry-over. Whenever the flowrate was stable, the proportionate mass is consistent with every stream recorded. For example, the flow of each stream was compared to what phases they were. The observed weight in outgoing water streams increased by 0.9 kg (1.97 lb) of which similar loss in weight recorded in the product oil stream for recorded flow of 0800 hrs in Figure 3. The dips observed at 1400 hours and 0200 hours were due to crude oil feed tank change out.

![Figure 2. Flow of total feed and product streams over time.](image-url)
3.2. Effect of wash water volume on salt recovery
The amount of water typically used in refinery for desalting process ranges from 5 – 10 vol% and is a major factor contributing to salt removal efficiency (SRE) [33-35]. The standard operating conditions of desalting conditions in PETRONAS refineries was applied in the pilot trial [31]. The measure of success is analogous to SRE which is the maximization of mass recovery of naphthenate salts in the water phase.

10 vol% water wash relative to the volume of the treated crude oil were performed on the pilot plant as a baseline. The trial was monitored on hourly basis as to determine the hour of which the process has reached steady state. Figure 4 below summarizes the three observations made during the trial. The first observation is the maximum recovery of the ionic liquid – product or naphthenate salts using 10 vol% water wash is at average of 0.2 mol per unit mass of naphthenate salts recovered as compared of incoming stream of 0.7 mol/unit mass. This translates to 29 % recovery. The final observation is the amount of time to reach steady state is between 8 to 10 hours. The third observation which is the utilization of desalter as potential process unit to recover naphthenate salt and ionic liquid post neutralization is viable. The primary function of desalter in conventional refinery setting is to recover wash water containing inorganics from the crude oil prior to distillation process. The inorganics may cause scaling and fouling in the hot train and furnace potentially causing unscheduled shutdowns. And the electrostatic field help to hasten the process of water recovery by encouraging contact between water droplet and allows gravity to take over and form the continuous aqueous phase [37]. Clear separation between the oil phase and aqueous phase was observed and maintained throughout the run. As comparison, more than six (6) hours is required for a clear separation to takes place in absence of electrostatic field [36].

The low recovery rate of the ionic liquid derivatives in the water phase was discussed in detail elsewhere [18,30]. Factors cited were hydro- and lipophilicity behavior of the resultant ammonium cation and naphthenate anions, molecular structure of the acids themselves which some may exhibits surfactant-like propensity and formation of tight water – oil emulsion. Similar observation was made by
Hussain [36] whereby limited by solubility of naphthenate salts in simple alcohols around 0.26 mol/unit mass would require significant amount of alcohol to increase the recovery up to 80 %. This is further supported by analysis done by Anderson et al. [28] that shows the range of NA’s molecular weight can be up to carbon number of 34. The formation of tight emulsion of oil-water-oil was promoted by large surfactant – like naphthenic acid derivatives and get carried over by the aqueous phase.

A trial of 30 vol% water wash was performed and the results is presented in figure 5. Steady state was achieved after the 4th hour with maximum recovery of 0.49 mol of naphthenate salts/unit mass, which is equivalent to ~70 % mass recovery. Higher amount of water wash was not considered due to diminishing economic return on ionic liquid recovery. The additional mass recovery was in the single digit range against incremental volume of 10 % [36].

![Figure 4](image-url)  

**Figure 4.** Quantity of naphthenate salt extracted, utilising water in an amount of 10 % by volume of the mixture, as a function of process operating time. (◊ - naphthenate salts in recovered water phase; € - Calculated amount of naphthenate salts in the incoming stream).
Preliminary investigation on the feasibility of two-stage desalting with consecutive two (2) cycle of 30 vol% water wash, each consists of freshwater makeup, was performed. The second stage desalter only able to achieve additional 3.5 % recovery. Considering that, no more test was pursued on that idea. This confirmed the observation made by previous studies [18,30,36] on the solubility limitation posed by the anion of the salt which is naphthenate itself. The electrostatic field itself works as design which is to increase the movement of the water droplet and increased collision until the water droplet becomes big enough to be influenced by Stokes Law and merge with the separated phase [2]. Figure 6 illustrates the changing conditions of the recovered wash water containing the ionic liquid as the extraction process reaching steady state, suggesting more hydrocarbon being co-extracted in the water phase as a result of emulsion formation. The follow-up tests in Malaysia revealed that non-naphthenate based hydrocarbon was within 1 – 2 wt% using the method described in [28]. This is rather consistent with work done by Anderson et al. [28] whereby as the emulsion become tighter because of the lipophilicity of the naphthenate salt, more oil become entrapped as oil-water-oil formation which could not be easily broken up by mere droplet to droplet collision caused by electrophoretic forces [37]. To break such emulsion, chemical intervention is needed such as pH control and emulsion breaker to ensure lower hydrocarbon losses. Hydrocarbon losses would translate into higher end-of-pipe treatment cost and opportunity cost to the refinery.

Figure 7 compares the extraction efficiency of 10 vol% versus 30 vol% water wash with maximum recovery at 68 % - 72 %, with extra cautions to note on mixing and sample homogeneity. Compared to gravity-assisted settlement performed by PRSB of which requires visual inspection, bigger density difference between phases and long-standing time (more than one hour) for separation of two or more non-miscible phases, faster separation can be performed adequately with electrophoretic force induced by oscillating current and bigger droplets via gravity-assisted settlement (Stokes’ Law) of which is found adequate for desalter’s residence time of 15 mins and less.
3.3. Preliminary finding on in-line mixing for $[N_{4441}][CH_3CO_3]$ and pyrenees oil

The Intertek desalting pilot plant is equipped with a $\frac{3}{4}$ inch piping with optional heat tracing that can be used for injecting chemicals without the need to pre-mixed it with water or hydrocarbon itself (no figure available to illustrate this). The aim is to seek evidence that the residence time and mixing power required for fresh crude oil and $[N_{4441}][CH_3CO_3]$ to get to low TAN value can be achieved via pipeline volume/length and pressure drop mixing valve.

The mixing point (which is a three-way pipe connection) is located upstream of the mixing valve and based on the volume and flowrate used, the length of pipeline would allow seven (7) mins residence time. Samples were taken out from a sampling point made available at the desalter itself as no other sampling port are available. The result indicates that the concept is feasible as the resultant crude oil met the low TAN value of <0.3 mg KOH/g of oil. The reaction itself is quite fast but the mechanics of diffusivity of ionic liquid into crude oil bulk phase to achieve ‘deep de-acidification’ can be an issue.
The combination of pipeline volume and pressure drop mixing valve provide adequate mixing and time, in addition to that the hold-up volume of the desalter would provide additional 15–20 mins.

4. Conclusions
All objectives outlined for this work were achieved. The technical performance criteria were met as the desalter allows the recovery of the ionic liquid derivatives or naphthenate salts post contact with the aid of 30 vol% water wash, at temperature and pressure of 90 °C and 10 barg and within the residence time of the vessel. The novelty of this work which is the continuous extraction of the ionic liquid was demonstrated with good process stability and exemplified chemical stability pre and post extraction, a huge improvement in term design challenge as this will eliminate the need for huge buffer vessel to accommodate long standing period to aid phase separation. Furthermore, the ionic liquid is now can be regenerated with minimum fresh make-up needed.

Unconventional water amount needed may pose significant barrier for commercialization effort due to availability and finite nature of the resource for most refineries. New techniques and process strategy are being explored such as designing smaller cation as to induce better solubility in the water. Tight water-hydrocarbon emulsions were observed in the water phase and it will present itself as opportunity cost if it is not recovered.

References
[1] Zeidani K and Bahadori A 2006 J. Canadian Pet. Tech. 44 5
[2] Goode M 2001 5th Intl. Conf. Refining Processes AIChE
[3] Pereira J, Velasquez I, Blanco R, Sanchez M, Perenade C and Canelon C 2015 Advances in Petrochemicals (IntechOpen)
[4] Pruneda E F, Escobedo E R B and Vazques F J G 2005 J. Mex. Chem. Soc 49 4–19
[5] Adams F V, Chukwuneke C E and Agboola B O 2019 Processing of Heavy Crude Oils: Challenges and Opportunities (IntechOpen)
[6] Mandal P C and Sasaki M 2018 Recent Insights in Petroleum Science & Engineering (IntechOpen)
[7] Brown L D and Ulrich A C 2015 Chemosphere 127 276–90
[8] Clemente J and Fedorak P A 2005 Chemosphere 5 585–600
[9] Fan T 1991 Energy Fuels 3 371–5
[10] Othmer K 2007 Kirk-Othmer Encyclopedia of Chemical Technology (New Jersey: John Wiley & Sons)
[11] Qian K, Robbins W, Hughey C, Cooper H, Rodgers R and Marshall A 2001 Energy Fuels 6 1505–11
[12] Tomczyk N, Winans R, Shinn J and Robinson R 2001 Energy Fuels 6 1498–1504
[13] Wasserscheid P and Welton T 2007 Ionic Liquids in Synthesis (Germany: Wiley – VCH)
[14] Deetlefs M and Seddon K R 2006 Chim. Oggi 2
[15] Greaves T and Drummond C 2008 Chem. Rev 1 206–37
[16] Luo H and Luo H 2009 J Phys. Chem. B 13 4181
[17] Anderson K, Goodrich P, Hardacre C, Hussain A and Rooney D W 2013 Fuel 108 715–22
[18] Anderson K, Atkins M P, Goodrich P, Hardacre C, Hussain A S, Pilus R and Rooney D W 2015 Fuel 146 60–8
[19] Shah S N, Mutalib M I A, Pilus R and Lethesh K C 2015 Energy Fuels 29 106–11
[20] Varadaraj R, Savage D W and Wales W E 1998 US5961821A
[21] Varadaraj R, Pugel T M and Savage D W 2000 US6096196A
[22] Maase M and Massonne K 2005 Ionic liquids IIIB: fundamentals, progress, opportunitities & challenges ACS Symp. Ser.
[23] Hart P R and Cunningham T L 2008 WO137523A1
[24] Manan N A, Atkins M P, Jacquemin J, Hardacre C and Rooney D W 2012 Sep. Sci. Tech. 47 312–24
[25] Shah S N, Mutalib M I A, Ismail M F, Suleman H, Lethesh K C and Pilus R M 2016 J. Mol. Liquid 219 513–25
[26] Coto B, Suarez I, Chirita M, Conde J, Gimenez R, Rodrigues N, Alvarez N and Pena J L 2019 Sep. Purification Tech. 223(15) 234–42
[27] Anderson K, Hussain A S, Atkins M P and Basar J 2014 WO2014016425A1
[28] Anderson K, Atkins M P, Goodrich P, Hardacre C, Hussain A S, Pilus R and Rooney D W 2015 Fuel 146 60–8
[29] American Association for Testing and Materials International (ASTM) 2006 ASTM D664 Standard Test Method for Acid Number of Petroleum Products by Potentiometric Titrations
[30] Hussain A and Basar J 2021 Proc. Int. Conf. Process Eng. Adv. Mater. (ICPEAM) 287 04002
[31] Personal communications between project team and PETRONAS Penapisan Melaka
[32] Pyrenees Crude Oil Overview 2009 BHP
[33] Sellami M H, Naam R and Temmar M 2016 J. Pet. Env. Biotech. 271 71–6
[34] Al-Otaibi M, Elkamel A, Al-Sahhaf T and Ahmed A S 2003 Chem. Eng. Comm. 190 65–82
[35] Abdul-Wahab S, Elkamel A, Madhuranthakam C R and Al-Otaibi M 2006 Chem. Eng. Procs. 45568–577
[36] Hussain A S 2010 Extraction of acids using ionic liquids PhD Thesis Queens University of Belfast
[37] Sams G 2010 Challenges of desalting canadian crudes Presentation PETRONAS
[38] Hughes H B 1981 Operating Manual for Two Stages Single Phase 60kV Desalter (USA: Intertek PARC)

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